

Development Report No. 9

V.T.E. Process Development

Freeport Test Facility

Freeport, Texas

United States Department of the Interior



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FOREWORD

This is one of a continuing series of reports designed to present accounts of progress in saline water conversion and the economics of its application. Such data are expected to contribute to the long-range development of economical processes applicable to low-cost demineralization of sea and other saline water.

Except for minor editing, the data herein are as contained in a report submitted by the contractor. The data and conclusions given in the report are essentially those of the contractor and are not necessarily endorsed by the Department of the Interior.

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A. INTRODUCTION

Development Run 15 lasted from April, 1968 through September, 1968; Development Run 16 lasted from December, 1968 through May, 1969.

Development Runs 15 and 16 are covered in this report. Their discussion is combined because of the similarity of the test objectives studied.

Major importance was placed on evaluating the double-fluted tube bundle and the titanium tube bundle to see how well they perform with time in a large desalination plant. Condensate handling-restrictions in the 5 effect module were studied. A major sump to sump brine transfer system modification was designed, installed and tested. An attempt was made to improve brine distribution by the use of porcelain spray nozzles in two effects. Enhanced surface spirally grooved tubes were evaluated both in a brine preheater and in the high temperature auxiliary test unit. Deaerator control at Run 14 optimum conditions and minimum acid addition rate were closely monitored. A full sized system of oxygen scavenging was operated, modified and evaluated.

B. OBJECTIVES

The major objectives of Runs 15 and 16 were to evaluate the overall plant performance operation and to monitor the performance of the new heat transfer surface area in the plant.

The actual stated objectives at the start of Run 15 were as follows:

- (A) Obtain complete heat and material balance data, utilizing conventional and high accuracy instrumentation, to establish process conditions for evaluation of overall seventeen-effect performance.
- (B) Obtain complete heat and material balance data, utilizing conventional and high accuracy instrumentation to establish process conditions within the five-effect module for evaluation of modular evaporator performance.
- (C) Evaluate the heat transfer performance of new evaporative heat transfer configurations or materials, including:
 - 1. Double-Fluted Copper Alloy 194 tube bundle in Effect 13.
 - 2. Titanium tube bundle in Effect 9.
 - 3. 316 SS tube bundle in Effect 14.
- (D) Evaluate new or revised methods of fluid handling including:
 - 1. Sump-to-sump brine transfer from Effect 10 to Effect 14.
 - 2. Modified condensate transfer within the five-effect module.
 - 3. Brine distribution to the tubes of Effect 11, utilizing SVL distributors and procelain flow nozzles.
- (E) Evaluate Deaerator-Decarbonator performance, with improved acid-sea water mixing, at various rates of acid injection, with residual alkalinity of the effluent as the acid feed control criteria.
- (F) Evaluate the relative performance of competitive mist eliminators in Effects 13 and 14.
- (G) Investigate various venting rates and modes in the five-effect module.
- (H) Evaluate stabilization of product water by the injection of liquid sodium silicate into the Freeport product stream.
- (I) Evaluate the effect of periodic shock chlorination of sea water feed and heat rejection condenser cooling water with quantities of chlorine lower than previously used.

- (J) Operate the Auxiliary Test Unit (ATU) at various evaporating brine temperatures for investigation of heat transfer performance and calcium sulfate (anhydrite) scaling limits.
- (K) Perform special chemical analyses related to scaling and corrosion.
- (L) Evaluate the performance of various materials of construction.
- (M) Begin shakedown operations and calibration of Rapid Data Acquisition System (Data Logger).

Objectives D-3, F and M were postponed to Run 16 because of time and/or equipment limitations. Other objectives were modified or expanded as the need for more or less information became evident during the Run.

Target operating conditions for the plant and the ATU for Development Run 15 are in Table B-1. These were maintained throughout the Run except when variations were required to satisfy specific objectives or when operation philosophy was changed.

TABLE B-1

DEVELOPMENT RUN 15 TARGET CONDITIONS

17-EFFECT EVAPORATOR

First Effect Sump Brine Temp, °F	266-268
First Effect Condensing Press, psig	31.5 maximum
Heat Rejection Condensing Temp, °F	90-100
Clarifier-Thickener in Service	Yes, from cold leg
Heat Rejection Condenser Flow Pattern	Alternate weekly
Scale Inhibitor Injected at P-26 Disch.	Nalco 918, 4 ppm \pm 1
Concentration Factor of Feed to Eff. 1	0.90 - 1.00
Final Concentration Factor	3.10 - 3.20
Sea Water Feed Rate, lb/hr	As required
Tube Feed Rates, lb/hr	Max., Effects 10-17
Forward Feed to Effects 11 to 14	Sump-to-sump
Atmospheric Vent from Effect 4, lb/hr	400-600
Vent from Effect 12 to Stm. Jets, lb/hr	200-400
Cascade Vents	1 to 4, 5 to 12
Module Vents	Vent Tubes
Desuperheater Water Rate, GPM	To give 300-325°F
Stripping Steam to DA-DC from Eff. 17, lb/hr	700-800
DA-DC Sea Water Inlet pH	As required to give 15 ppm CaCO ₃ residual alkalinity
DA-DC Sea Water Outlet pH as Adjusted	5.8 - 8.0
Freeport Product Stabilizer	N-silicate
Freeport Product pH	8.5 - 10.5
Chlorination of Sea Water Feed	Weekly, Shock

AUXILIARY TEST UNIT

The operation of the ATU should duplicate the operation of the First Effect as closely as possible; equal quantities of flow per tube, of preheating, and of sea water concentration. The following table summarizes the desired operating conditions.

	A	B	C
Sump Brine Temperature, °F	266 - 268	271 - 273	276 - 278
Inlet Sea Water Temp., °F	263 - 264	268 - 270	273 - 275
Sea Water Feed Rate, lb/hr	As required for tube loading equivalent to Effect 1		
Steam Supply, lb/hr	As required for concentration equivalent to Effect 1		
Degree of Concentration	6-9%, as required for concentration equivalent to Effect 1		
Inlet Concentration Factor	0.90 to 1.00	0.90 to 1.00	0.90 to 1.00
Operating Time, Hours	500	500	500

The actual initial objectives for Run 16 were as follows:

- (A) Obtain complete heat and material balance data, utilizing conventional and high accuracy instrumentation, to establish process conditions for evaluation of modular and overall 17 effect performance alone and with time.
- (B) Evaluate the heat transfer performance of the enhanced tubes and the vapor shear mode of operation of the new test preheater.
- (C) Evaluate the heat transfer performance of new evaporative heat transfer configurations or materials.
 - 1. Evaluate double fluted CDA 194 tube bundle in Effect 13.
 - 2. Evaluate titanium tube bundle in Effect 9.
- (D) Evaluate new or revised methods of fluid handling.
 - 1. Investigate sump-to-sump brine transfer from Effect 10 to Effect 14.
 - 2. Evaluate sump-to-sump brine transfer through the bulkhead from Effect 13 to 14.
- (E) Evaluate the Deaerator-Decarbonator Systems.
 - 1. Monitor the routine performance while operating at the optimum conditions as defined by run 14, by measuring the dissolved oxygen and carbon dioxide in the brine effluent.
 - 2. Study the effectiveness of catalyzed sodium sulfite in scavenging residual oxygen from the deaerator brine effluent.
 - 3. Determine the feasibility and practicability of running the deaerator at an unadjusted pH out of as high as 6.5.
 - 4. Evaluate the operability, stability, and response of the automatic acid addition rate control instrument system.
- (F) Evaluate Mist Eliminator Performance.
 - 1. Monitor the relative performance of competitive mist elimination in Effects 13 and 14.
 - 2. Determine the relative performance of modular effects with and without mist eliminators.

(G) Vent Studies

1. Determine the effect of changes in vent configuration and/or venting rate.
2. Evaluate the effect of blocking off the top of vent tubes and compare vent tubes with external steam chest bottom vents.

(H) Evaluate the performance of LTV evaporators to determine the optimum tube feed rate while operating in the sump-to-sump mode.

(I) Determine the operating capabilities of various types of tube feed weirs and the effect of said weirs on such things as brine distribution and overall heat transfer coefficient.

1. Determine the effectiveness of SVL weirs in feeding the double fluted tubes in Effect 13.
2. Study the brine distribution to the tubes of Effect 11 utilizing porcelain flow nozzles.
3. Evaluate the effect of vortex flow distributors on the operations of Effect 9.

(J) Operate the auxiliary test unit (ATU) with enhanced surface tubes at various evaporating brine temperatures for investigation of enhanced surface heat transfer performance and determination of calcium sulfate anhydrite scaling limits with enhanced surface.

(K) Perform special chemical analyses related to scaling and corrosion.

1. Determine the amount of ionic calcium and sulfate in brine and sea water samples.
2. Continue to investigate methods of improving precision and accuracy of the chlorinity determination used in making material balances.
3. On a routine basis, determine the copper and iron composition of the condensate and brine stream into and out of the 5 Effect module.
4. Set up and monitor systems for routine control sample analyses.
5. Set up and perform special tests as required.
6. Develop procedures for cation analyses with the atomic absorption unit.

- (L) Evaluate the performance of various materials of construction using the probolog, test spools and making metal inspections as necessary, recording all observations on the appropriate materials performance form.
- (M) Carry work on data logger through installation, calibration, and shake down phases up to the point at which the data acquisition system is ready to be used to collect data on a routine basis.
- (N) Determine the reliability and the usefulness of a corrater as a continuous corrosion monitor.
- (O) Determine the cause of and recommend solution for any operational problems that had developed in Run 15 or do develop in Run 16.

Objectives D-2, E-4, F-1, I-3 and M were postponed because of equipment limitations. Objectives G, H, I-1, J, and O were not finished because of the abrupt termination of Development Run 16. Other objectives were modified or expanded as the need for more or less information became evident during the run.

Target operating conditions for the plant and the ATU for Development Run 16 are in Table B-2. These were maintained throughout this Run except when variations were required to satisfy specific objectives.

TABLE B-2

17-EFFECT EVAPORATOR

	<u>Target</u>	<u>Control Limits</u>	<u>Remarks</u>
<u>A. Seawater Feed Circuit</u>			
1. Seawater Feed Source			
a. Intake Pit Entrance	-	-	North Entrance for all Heat & Material Balances. Switch weekly for alternate chlorination.
b. Clarifier-Thickener	-	-	In service as settling basin. Feed from cold leg.
2. Brine Feed Conditions			
a. Blended Brine Concentration factor	1.01	0.99 to 1.04	P-27 discharge is source of blending brine.
b. Blended Brine Feed rate, lb/hr	-	Maximum Controllable	
3. Deaerator-Decarbonator			
a. Stripping Stm., lb/hr	750	500 to 1000	No. 17 Effect chest is source of stripping steam.
b. Flashdown, °F	0.8	0.6 to 1.0	Regulate vapors out of Deaerator-Decarbonator to control flashdown
4. Feed Treatment			
a. Acidity of Deaerator-Decarbonator outlet brine before caustic neutralization, pH	5.7	5.5 to 5.9	Regulate acid injection rate to control Deaerator-Decarbonator outlet brine pH.

TABLE B-2 (CONTD)

17-EFFECT EVAPORATOR

	<u>Target</u>	<u>Control Limits</u>	<u>Remarks</u>
<u>A. Seawater Feed Circuit (CONTD)</u>			
4. Feed Treatment (Contd)			
b. Acidity of Deaerator- Decarbonator outlet after caustic neutralization, pH	6.8	6.6 to 7.0	Regulate caustic addition rate to control pH
<u>B. First Effect Operations</u>			
1. Sump Brine Temperature, °F	268.0	267.5 to 268.5	Regulate by adjusting steam flow to first effect steam chest.
2. First Effect Chest Press, psig	-	<u>31.5 Maximum</u>	-
3. Desuperheater Water Rate, gpm	-	-	As needed to keep desuperheater steam inlet temperature between 300-330°F.
<u>C. Last Effect Operations</u>			
1. Final Effect Blowdown (P-27 discharge) Concentration Factor	-	<u>3.0 Maximum</u>	No polyphosphate scale inhibitor added to P-26 discharge.
2. Final Condensing Temp., °F	as low as possible	-	Pumps P-1 & P-2 on full open to Heat Rejection Condenser (HX-318)
3. Heat Rejection Condenser Flow Pattern	-	-	Alternate Weekly

TABLE B-2 (CONTD)

17-EFFECT EVAPORATOR

	<u>Target</u>	<u>Control Limits</u>	<u>Remarks</u>
<u>D. Shock Chlorination</u>			
1. Chlorine added to Blended brine feed, ppm	10	8 - 12	2 Hours, weekly.
2. Chlorine added to heat rejection cooling water, ppm	6	5 - 7	2 Hours, weekly.
<u>E. Condensate Transfer System</u>			
1. No. 1 Steam Chest Condensate	-	-	Flashdown to FT-2, P-43 off.
2. Hand Control valve in condensate line between flash-tanks 2 & 3	-	-	Closed enough to maintain level in FT-2.
3. Condensate to Module	-	-	On direct flashdown with all valves open. P-44 off.
4. High Temp. condensate withdrawal system	-	-	Off, valve from Flash tank 15 to product pump (P-41) is shut.
5. Heat Rejection condensate system	-	-	To product pump (P-41) suction. P-42 off.
6. Level in Flash Tank 17	As low as possible	-	Just above P-41 cavitation point.
<u>F. Vents</u>			
1. Configuration	-	-	All effect and exchanger vents except those noted below are cascade. Vent rate and configuration as needed to prevent non-condensable binding but start with same as last heat and material balance in Development Run 15.
a. 12 Effect Plant	-	-	
i) No. 4 Effect Chest 1b/hr	500	400-600	Steam Chest bottom vent only.
ii) HX-312 Bottom Vent	Wide Open	-	HX-312 vents through lower vent line that ties into 6" vent line from HX-318

TABLE B-2 (CONTD)

17-EFFECT EVAPORATOR

<u>F. Vents (CONTD)</u>	<u>Target</u>	<u>Control Limits</u>	<u>Remarks</u>
iii) No. 12 Effect Vent	Top & bottom Chest Vents Wide Open	-	No. 12 effect vents through upper vent that ties into D/A vapor line.
b. Module			
i) Vent tubes	All Open	-	
ii) External Evaporator and heat exchanger vents	All Closed	-	
iii) Heat Rejection Condenser Vent	Wide Open	-	
G. <u>Inter-Effect Brine Flows</u>			
1. Feed Forward	-	-	Effects 1 through 10. Effects 15 through 17. All feed forward effects on maximum recycle.
2. Sump-to-sump feed	-	-	Effects 11 through 14. All sump-to-sump transfer line valves wide open.
a. Tube Feed rate, lb/hr/tube			
No. 11 Effect	1000	900 to 1000	-
No. 12 Effect	1000	900 to 1100	-
No. 13 Effect	1000	900 to 1100	-
No. 14 Effect	Wide Open	-	-
H. <u>Freeport Product Water Condition, pH</u>	9.5	8.5 to 10.5	Regulate with N-Silicate addition when delivering to the City of Freeport.

AUXILIARY TEST UNIT

The operation of the ATU should duplicate the operation of the First Effect as closely as possible; equal quantities of flow per tube, of preheating, and of sea water concentration. The following table summarizes the desired operating conditions.

	A	B	C
Sump Brine Temp., °F	267.5-268.5	271-273	276-278
Inlet Sea Water Temp., °F	265-267	269-271	274-276
Sea Water Feed Rate, lb/hr	As required for tube loading equivalent of Effect 1		
Steam Supply, lb/hr	As required for concentration equivalent to Effect 1		
Degree of Concentration	As required for concentration equivalent to Effect 1		
Inlet Concentration Factor	0.99-1.04	0.99-1.04	0.99-1.04
Operating Time, Hours	500	500	500

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C. CONCLUSIONS AND RECOMMENDATIONS

On the basis of the work done during Development Runs 15 and 16, at the Freeport Test Facility, the following conclusions are made:

1. Overall Plant Performance.

- a. The plant is now and has been kept in a safe operating condition. A thorough metal inspection has shown no unusual or significant flaws in the operating process equipment. Operation of both the module and the old 12 effect plant can be continued safely for several years as long as the past routine preventative maintenance procedures are followed.
- b. The overall performance of the 17 effect plant did not decline noticeably between development runs 15 and 16. However, there was what appeared to be a significant decline in performance of some of the individual evaporators. Tube fouling is a long range phenomenon and continues over the entire life of a plant.
- c. Sump-to-sump brine transfer is a satisfactory mode of operation. Level control in this mode is easy. Under proper design conditions stratification or over concentration does not occur.

2. Performance of Experimental Heat Transfer Equipment.

- a. The double fluted tube bundle performance declined significantly between Development Runs 15 and 16.
- b. The titanium tube bundle performance was excellent and showed no decline between Development Runs. LTV evaporator tubes made of thin walled titanium tubing appear to perform and last well in a hot concentrated brine environment.

3. Performance of Process Equipment.

- a. The condensate transfer system in the module is still inadequate. Further enlargement of the condensate transfer piping is indicated.
- b. The performance of the module is significantly affected by small amounts of non-condensibles. The venting system is improperly sized.
- c. The performance of the heat rejection condenser is unsatisfactory; the heat rejection condenser performance declines significantly at low seawater temperatures.

- d. The porcelain spray nozzles do a better job of brine distribution than the SVL weirs. There is no difference in heat transfer performance between tubes equipped with porcelain spray nozzles and tubes equipped with SVL weirs.
- e. An acid mix chamber is not required to obtain satisfactory mixing of the acid and the sea water feed to the deaerator.
- f. The deaerator effluent is a satisfactory sensing point for pH control. Controlling the deaerator effluent between a pH of 5.6 and 6.0 does not cause any scale or corrosion problems. Caustic neutralization can, under these conditions, be eliminated.
- g. Sodium sulfite is successful in scavenging residual oxygen from the deaerator brine. However, removal of the residual oxygen does not significantly affect externally measured plant corrosion rate.
- h. The deaerator effluent dissolved oxygen level runs between 50 - 100 ppb. The optimum deaerator operating conditions appear to be one pound of stripping steam per thousand pounds of brine and about 0.5°F. flash down at the deaerator top.
- i. When making potable water only, the low temperature effects in the module can be operated without mist eliminators.
- j. Shock chlorination is effective in preventing marine growth in circulating brine process streams.

4. Materials Evaluation.

- a. If properly fabricated, glass reinforced plastic pipe holds up well in hot brine service. There is no significant difference in performance between the products of the two main producers of glass reinforced plastic pipe.
- b. Cement liners can be used to protect carbon steel evaporator sumps.
- c. CDA 194 is a satisfactory material for use in evaporator falling film tubes. The test of CDA 194 in heat exchanger tubes is not deemed significant because of the low brine velocity.

5. Scaling.

- a. With normal sea water, 273°F. appears to be the incipient scaling point for LTV evaporator tubes.
- b. If operating conditions are controlled properly, calcium sulfate anhydrite scale will gradually wash out of higher temperature effects.

6. Data Gathering

- a. The performance of the data gathering instrumentation has declined with time. Many of the critical instruments require replacement or recalibration.
- b. The atomic absorption spectrophotometer is an exceptionally useful development engineering instrument. It can be used to perform large numbers of cation analyses rapidly and accurately.
- c. The concentration of iron and copper in the effluent brine fluctuates in a random fashion during normal operations. The copper and iron concentration in the effluent brine is extremely high just after start up and quickly drops down to a much lower equilibrium level.

On the basis of the work done during Development Runs 15 and 16, the following recommendations are made:

1. Overall Plant.

- a. Install a high temperature test module to determine the economics and performance of a multi-stage flash preheater and various enhanced surfaces at high temperatures.
- b. Automate the plant so that no operator attention or surveillance is required outside of normal working hours.
- c. The Auxiliary Test Unit objectives have been met with limited success. This unit required considerable operating and engineering attention. It should either be scrapped or shipped to Wrightsville Beach where it would receive the necessary engineering attention.

2. Pursue the Original Development Run 17 Objectives.

- a. The brine flash down device has been installed between effect 13 and 14. If this device works, considerable savings in external brine piping will be realized in future plants. The device should be thoroughly tested during operations.
- b. The equipment for automatic pH control has been received. Automatic acid addition will save considerable operating attention and allow complete automation of an LTV desalination plant. It should be tested out as soon as possible.
- c. Venting in the low temperature module is a problem. No. 15 effect was equipped to carry out venting studies. The mechanical work on no. 15 effect should be finished and the venting studies should be carried out.
- d. There is considerable evidence that brine distribution irregularities exist, especially in the domed effects. The brine catchment and measurement system should be moved to one of the domed effects so that brine distribution patterns under different operating conditions can be observed.

3. Continue the Evaluation of Materials in Actual Operating Process Equipment.

- a. Titanium has performed well in an evaporator tube bundle. Install a multi-stage flash preheater bundle with titanium to determine its effectiveness in preheater service.

- b. Tube out a preheater bundle with CDA 194 to expose the alloy to tube velocities between 7 and 10 feet per second on the brine side.
 - c. Use CDA 194 alloy evaporator tubes in one of the high temperature effects.
 - d. Install stainless steel test spools in hot brine piping exposed to considerable turbulence.
 - e. Put cement liners in high temperature effects.
4. Improved Level of Precision of Data Gathering Equipment.
- a. Replace main plant input and output fluid orifice meters with turbine meters.
 - b. Discard all existing high-accuracy thermocouples. Carefully and selectively reinstrument the entire plant with precision grade platinum resistance thermometers. The temperature measuring elements should be positioned in the effect steam chest rather than the steam transfer line. Attempts to measure brine temperature at the top and bottom of tube should be discarded. At least one effect should be instrumented in such a way as to determine the temperature profile in the steam chest.
 - c. Obtain the equipment necessary to enable the keeping of the temperature sensing elements in proper calibration.
 - d. Discard the absolute pressure manometers. Relocate the differential pressure manometer systems to the top deck so as to minimize problems with liquid in impulse lines.
 - e. Defer any further work on the data logger until the new instrumentation is installed, calibrated, and proved out.
5. Staff the Plant to Accomplish the Development Engineering Objectives in the Field.
- a. Provide the Development Engineer with competent field Process Engineering help.
 - b. Provide the Development Engineer with the services of technicians for the purpose of data gathering and routine calculations.
 - c. Only try to accomplish objectives that can be reasonably handled by the allotted staff.

6. Undertake Other Studies and Process Changes to Improve the Knowledge of the VTE Process.

- a. Study limitations to plant capacity during cold weather. Add a third-stage jet ejector so that the heat rejection condenser can be properly vented during the winter when the water temperature gets down to 50°F.
- b. Determine how much carbon dioxide is absorbed out of the deaerator vent gas in the barometric precondenser and intercondenser.
- c. The double fluted tube bundles performance has significantly declined. Pull out several of the tubes; have some sent to be tested and compared against new tubes in a single or double-tube evaporator. Section and inspect the other tubes.

Run the double fluted tube bundles in its present condition at the start of Development Run 17. Then, mechanically or chemically clean the bundle to determine if this brings the performance back to its original level.

Test out SVL weirs in the double fluted tube bundle.

- d. Install and test an internal bundle vent in the module preheater.
- e. By inspection, determine the extent of the tube fouling in the various evaporators. Chemically or mechanically clean an entire section of the plant to determine how much performance is improved.
- f. Make probalog and ultrasonic inspection of metal thicknesses throughout the plant. Develop average mil per year corrosion rates for various materials in various environments.

D. DURATION AND ENVIRONMENT

Development Run 15 started on April 17, 1968, and continued until September 20, 1968, for a total of 3,129 operating hours. This represents an on-stream factor of 81.5%. There were 9 shutdowns during the run, taking 715 hours to complete the necessary maintenance work.

Air temperature reached a high of 85°F on July 3, 1968, and dropped to a low of 55°F on April 30, 1968. There were 376 hours of rain. Sea water temperature reached a high of 89°F on July 2, 1968, and a low of 63°F on May 6, 1968.

Sea water feed concentration was a high of 0.97 during June, 1968. After a period of extremely heavy rains, an all-time record low of 0.09 was reached on June 26, 1968.

During September, 1968, low supply steam pressure of 125 PSIG, down from 150 PSIG, necessitated completing the run with the vacuum pump; the steam jets unable to operate at the low pressure.

Development Run 16 started on December 19, 1968, and continued until May 9, 1969, at which time the Office of Saline Water ordered the plant shutdown in the middle of testing work. The run was later termed complete as of the shutdown date. The run had a total of 3,105 operating hours for an on-stream factor of 91.8%. There were 12 shutdowns during the run, taking 279 hours to complete the necessary maintenance work.

Air temperature reached a high of 82°F on April 10, 1969, and dropped to a low of 31°F on January 1, 1969. There were 385 hours of rain. Hail fell once for a brief period in late spring. It was necessary to invoke full plant freeze protection for a total of five days during the run.

On February 14, 1969, high tides caused by 40 to 50 MPH winds caused in-plant flooding. The Ethyl-Dow Chemical Company, by lowering the sea water supply flume level, managed to avert an emergency shutdown.

Sea water temperature reached a high of 76°F on May 6, 1969, and a low of 50°F on January 4, 1969.

Sea water feed concentration was a high of 1.00 on May 7, 1969, and a low of 0.50 on April 17, 1969.

For a three week period during the run, a low fluctuating supply voltage gave concern about motor overheating. The voltage, at its worst, reached a low of 310 volts, down from 480 volts. The plant was shutdown on January 28, 1969, following this extremely low supply voltage.

E. EQUIPMENT AND PROCESS CHANGES

Equipment changes prior to Run 15 were as follows:

1. No. 9 Effect Admiralty tube bundle was removed and a new bundle with titanium tubes, raised tube sheet, dome, and steam corridors was installed.
2. No. 13 Effect 90/10 copper/nickel tube bundle was removed and a new bundle with CDA-194 double fluted tubes, porcelain spray nozzles, and a No. 13 to 14 dome to dome vent was installed.
3. The internal module condensate transfer piping was enlarged from 8" to 10".
4. Sump to sump brine transfer piping was installed in Effects 10 through 14, together with the necessary piping for recycle tube feed and flow meters.

Process changes for Run 15 were established at the beginning or during the Run. They were:

1. Sump to sump with recycle versus top waterbox forward feed.
2. 268°F set as maximum at No. 1 Effect sump.
3. Sodium silicate to be added only to City of Freeport water.
4. Shock chlorination duration decreased from 4 to 2 hours per each circuit.

Equipment changes prior to Run 16 were as follows:

1. A polyester tank replaced the original carbon steel flash tank No. 10.
2. Exchanger No. 200 was removed and by-pass spools for both brine and sea water were installed. Pump P-43 was left as is.
3. Vapor shear preheater HX306X was installed in place of HX306b.
4. A level control valve was installed at P-41 discharge at the split tank. It is controlled by the level in Flash Tank 17.

5. Porcelain spray nozzles were installed in No. 11 waterbox together with a dome to dome, 11 to 12, vent line.
6. A sulfite addition system was installed.

Process changes for Run 16 were established at the beginning or during the Run. They were:

1. Deaerator acid addition control point switched to brine effluent.
2. Acid control range changed to 6.0 -- 6.4 pH.
3. Caustic neutralization stopped.
4. Oxygen scavenging by catalyzed sodium sulfite injection into the deaerator brine effluent started.
5. 266°F maximum No. 1 sump temperature; no blending.
6. Shock chlorination duration decreased from 2 to 1 hour per each circuit.

A table of installed heat exchanger heat transfer area (outside) for Runs 15 and 16 can be found in Table E-1.

TABLE E-1

HEAT EXCHANGER OUTSIDE HEAT TRANSFER SURFACE

Heat Exchanger	<u>AREA (A_O) SUMMARY</u>	
	<u>DEVELOPMENT RUN 15</u>	<u>DEVELOPMENT RUN 16</u>
	Area (A _O) Ft ²	Area (A _O) Ft ²
200	520	0
302a	1,800	1,800
302b	1,560	1,560
303a	1,380	1,380
303b	1,440	1,440
304	2,290	2,290
305	3,010	3,010
306a	1,700	1,700
306b	1,790	220
307	1,790	1,790
308	1,790	1,790
309	1,800	1,800
310	1,770	1,770
311	1,740	1,740
312	1,980	1,980
313	1,460	1,460
314	1,460	1,460
315	1,770	1,770
316	2,420	2,420
317	2,420	2,420
318	2,220	2,220
212	1,850	1,850
213	2,140	2,140
	<hr/>	<hr/>
	42,100 Ft ²	40,010 Ft ²

F. LIMITS AND ACCURACY

1. FLOW MEASUREMENT

One of the main data gathering limitations at the Freeport Test Facility is the flow measuring devices. In Development Run 16, material balances could be closed only to about 3%.

There are several things that act to keep the basic material balance accuracy at about ± 2 wt. %. First, many of the flow rates pulsate. In most cases, this is due to the fluctuating back pressure that develops when superheated liquid flashes when passing through a nozzle. The pulsation amplitude can approach 5% of the flow rate.

Another factor that contributes to the basic uncertainty is the lack of precision in the orifice calculations. Although orifice plate correction factors can be calculated to as many as four digits, there is a basic uncertainty in the factors. The best way to obtain the maximum accuracy out of the flow meter system is to perform a volumetric calibration in the field under actual operating conditions. The basic repeatability of the calibration data should also be determined.

The transducers in the instrumentation circuit also cause some error. The output from a orifice plate is measured by a D/P cell where it is converted to a pneumatic signal that is transmitted to the board and converted through a mechanical linkage to a chart reading. Even after careful calibration, the error in each part of the system combines to cause a significant uncertainty in the flow reading.

For the best instantaneous orifice plate drop pressure readings, locally mounted manometers should be used. Great care must be taken to make sure that the impulse lines are set up properly. One way to improve accuracy is to use turbine type meters; these are usually a great deal more accurate than the orifice meters. Unfortunately, turbine meters are expensive and difficult to maintain.

2. ACCURACY OF TEMPERATURE MEASUREMENT EQUIPMENT

The temperature measuring sensors at the Freeport Test Facility fall into two general categories: operating and data gathering. Regarding the operating sensors, there are two multi-point devices that indicate temperature and one multi-point device that records temperature; all of these are in the control room. The multi-point recorder and the newest indicator work

from thermocouples and have a smallest scale division of 2°F. The other indicator was supplied with the original plant; it utilizes resistance thermometers and has a smallest scale division of 5°F. These operating temperature measuring devices are used only for operating purposes; they are never used for any data gathering or process evaluation purposes. The data that are read from these devices are recorded on the operator data sheets and are used only for monitoring the gross behavior of the process. The sensing elements have neither been calibrated or rechecked.

The data gathering temperature measuring devices fall into two major classes:

a. Glass Thermometers

Most of the material balance work is done with mercury filled, glass, total immersion thermometers.

The first problem with the glass thermometer might be called "positioning error." Attempts have been made to standardize the thermowell installations and subsequent thermometer insertions. The thermowells are for the most part vertical so that a heat transfer fluid can be used. Most of the thermowells stick out from the pipe the same distance; the heat loss from the thermowell head is minimized by uniformly applied insulation. The thermometer is inserted in such a way as to minimize stem correction; a one-hole rubber stopper is used to prevent heat loss from the top of the thermowell. Reading accuracy is minimized by use of a magnifying glass and sighting techniques to minimize any errors of parallax.

Some of the thermometers have been calibrated at single temperatures close to their normal operating temperatures; some of the thermometers are certified to the nearest 0.05°F over their whole range. Unfortunately, glass thermometers have a tendency to anneal with time and temperature. That the temperature characteristics of these glass thermometer changes is a certainty; how much they have changed since calibration or certification is unknown.

There are in general three groups of mercury filled, glass, total immersion thermometers in use in the plant today:

1) Yellow Back

These are laboratory grade thermometers. Uncalibrated, these thermometers can be assumed good to no better than $\pm 1/2$ of the smallest scale division (1°F). These thermometers are in general positioned around the preheat exchanger train.

2) Calibrated Thermometers

These precision grade thermometers are used mainly to determine the steam chest temperatures. They have been calibrated at the ice point and a temperature close to their normal operating temperature. These thermometers were probably good to $\pm 0.1^{\circ}\text{F}$ during Development Runs 15 and 16. It will, however, soon be necessary to recheck the calibration of these thermometers. Note that one of the problems with calibrating a thermometer is that the correction factor is frequently non-linear and the deviation from true temperature may be significantly different 5 or 10° away from the calibration point.

3) Certified Thermometers

These are short range (approximately 40°F) total immersion thermometers which have been certified against a Bureau of Standards thermometer in a calibration bath at several different temperatures by the manufacturer. Unannealed, their accuracy is well within $\pm 0.1^{\circ}\text{F}$. These have been used long enough that they should be rechecked at their normal operating temperatures to make sure that there has been no significant drift with time.

b. Thermocouples

Thermocouples are mainly used in the high accuracy instrumentation systems. These thermocouples were all calibrated at several points in their operating range when they were purchased. The present accuracy of these thermocouples is about $\pm 0.5^{\circ}\text{F}$. There are several potential error sources:

1) Potentiometer Reading

The Catalog No. 8686 Millivolt Leeds & Northrup Potentiometer is good to ± 3 microvolts or for an iron constantan thermocouple $\pm 0.1^{\circ}\text{F}$. This is consistent with the read-out which has a smallest scale division of 5 microvolts.

2) Cold Junction

The present technique for ascertaining cold junction temperature is to position a certified thermometer close to the thermocouple leads on the terminal board. This procedure is satisfactory to achieve precisions in the range of $\pm 1^{\circ}\text{F}$. The problem is that the system--thermometer, air, and thermocouple junction--is not totally insulated and does not get a chance to reach true thermal equili-

brium. Disequilibrium is generally caused by wind currents in and around the high accuracy cabinets and thermal currents in the high accuracy cabinet itself.

Fortunately, there are several different ways to solve the problem. The easiest, is to use the built in reference junction in the No. 8686 Leeds & Northrup Millivolt Potentiometer. This would bring the error down to within ± 6 microvolts or $\pm 0.2^{\circ}\text{F}$. Another solution would be to buy a commercial multi-point cold junction generator. These are generally good to well within $\pm 0.1^{\circ}\text{F}$, irrespective of external ambient conditions.

Probably the most practical solution to the cold junction problem, is the use of an ice bath. Getting an ice bath that is good to $\pm 0.01^{\circ}\text{F}$ is an extremely difficult job and probably requires commercial equipment. However, our requirements are not so stringent; the achievement of $\pm 0.1^{\circ}\text{F}$ accuracy at the cold junction is probably satisfactory. This can be done by using chipped ice with distilled water in an insulated container with some agitation.

3) Thermocouple Calibration

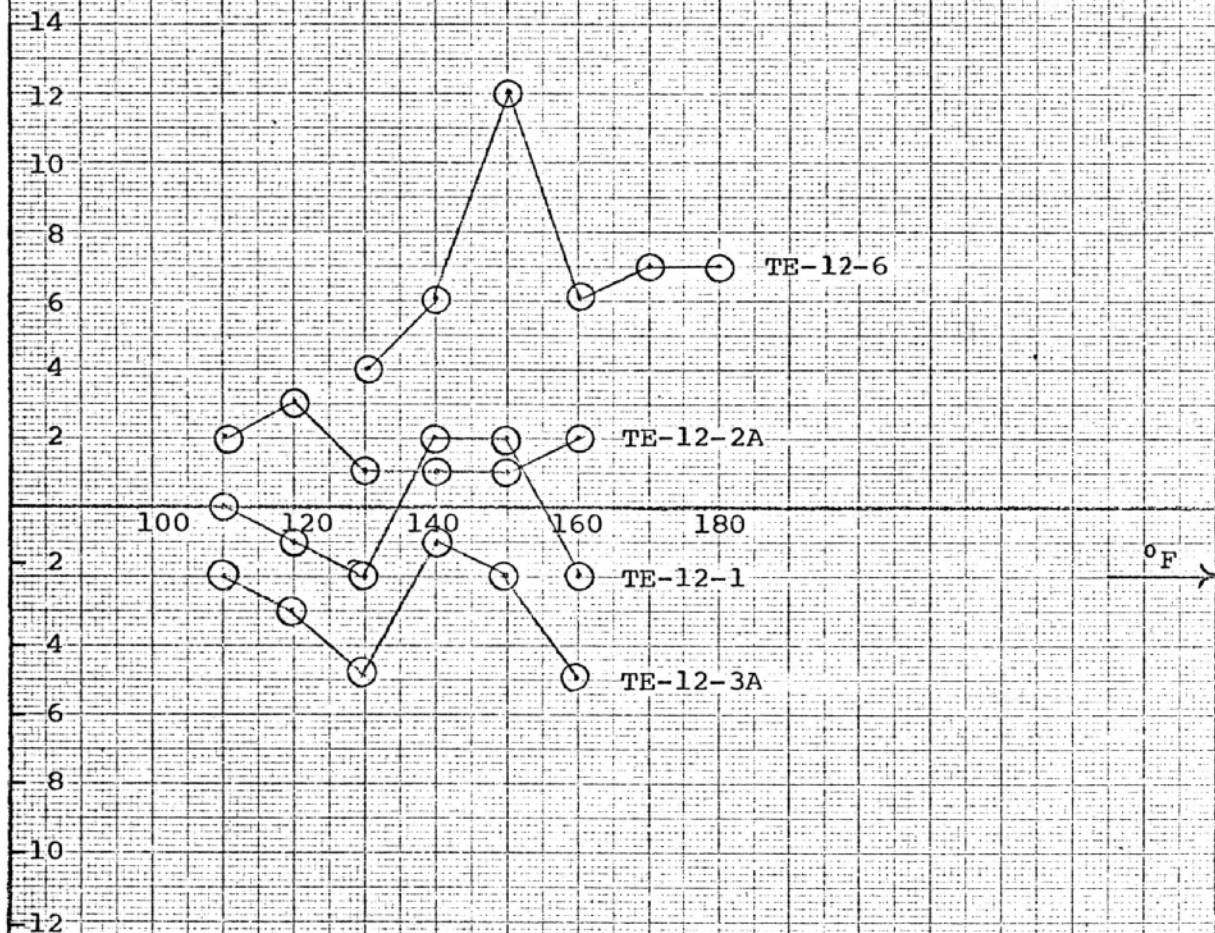
All of the high accuracy thermocouples were calibrated by the manufacturer at 10°F intervals over their normal operating range. Calibration, of course, brings the accuracy of the thermocouples down from the $\pm 2^{\circ}\text{F}$ normally associated with commercial thermocouples down to $\pm 0.5^{\circ}\text{F}$ for the calibrated thermocouples. To achieve better accuracy, some more data is needed.

The calibration curve for a typical set of high accuracy thermocouples is shown in the attached Figure F-1. It can be seen that there are some discontinuities in this calibration curve; these discontinuities are significant in that they amount to as much as $\pm 0.2^{\circ}\text{F}$ correction to the observed temperature. What is needed is for several of the thermocouples to be recalibrated the same way at the same temperature points several different times. From this, we would have an idea of the magnitude of the basic calibration uncertainty and also a better feel for how to treat the discontinuous areas.

4) Variability of Cold Junction Temperature

Most of the high accuracy thermocouples were calibrated with a cold junction of approximately 80°F .

Figure F-1
 Thermocouple Calibrations
 Effect No. 12
 Tested Thermocouple Microvolts
 (+) or (-) Error



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This is probably satisfactory for thermocouples that are used in control houses. However, the temperature in the high accuracy cabinets range from 35 to 90°F.

Probably the only way to eliminate the problem of variable cold junction is to either use an ice bath in all cases and recalibrate the thermocouples with an ice bath cold junction or stick with the present calibrations and buy an accurate 80°F constant cold junction controller.

5) Drift With Time

There is some evidence that the thermocouple calibration does drift with time. We have no data regarding what the magnitude of this drift is. The only way that we can determine the order of magnitude of this drift is to periodically recalibrate selected thermocouples.

The present accuracy of the high accuracy thermocouples is probably about $\pm 0.5^\circ\text{F}$. This uncertainty could be reduced to about $\pm 0.2^\circ\text{F}$ by undertaking a program of recalibration and using an ice bath.

3. DATA ACQUISITION SYSTEM

At some time in the future a data acquisition will be installed at the Freeport Test Facility. There are some limitations to these systems. A data logger simply reads the millivolt output from a variety of sensors and prints the information out; with respect to accuracy, the data logger usually adds a little uncertainty to the output signal that is being read.

The data logger gives instantaneous readings. In the case of flow meters in pulsing flow service, the data logger instantaneous reading can be considerably different than the true average reading. The data logger cannot improve the precision of the data gathering; a data logger installed at Freeport would not in any way improve the accuracy of the development data.

When planning a data acquisition system, it should be remembered that a great deal of money will have to be spent before any information is produced. First, the data logger and the sensors must be obtained and installed. Then, each sensor must be adequately loop-calibrated with the data logger, and a computer program must be written to convert the data logger output into meaningful engineering type information. Then the entire system must be tested out.

G. RESULTS AND COMPARISONS (CONTD)

1. OVERALL PLANT PERFORMANCE

a. Evaporative Heat Transfer Coefficients

Evaporator overall evaporative heat transfer coefficients determined on a salt balance basis have been calculated for selected heat material balance tests made during Development Runs 15 and 16.

These calculations were made on the basis of evaporator tube outside area. The heat transfer surface data that were used in these calculations are shown on Table G-1-a. The only real significant difference in heat transfer area between Development Runs 15 and 16 was the plugging of vent tubes for the latter run. Also shown on Table G-1-a are other data regarding the physical dimensions and material makeup of the evaporator bundles.

Shown on Table G-1-b are the overall evaporative heat transfer coefficients that were calculated during Development Runs 15 and 16 for comparison purposes. The individual coefficients calculated for each heat and material balance are averaged on a development run basis. These heat transfer coefficients are also calculated on a salt balance basis. The chlorinity data were obtained from analysis of composite samples collected during each heat and material balance run.

The evaporative heat transfer coefficients are for the most part calculated from raw, unadjusted data. In some cases it was necessary to adjust the brine sump temperature because of serious irregularities in the observed boiling point rise. It also appeared necessary in several cases to adjust the evaporator ΔT in the low temperature effects (16 and 17) to compensate for a lower temperature driving force at the top of the tubes due to the pressure drop across the brine side of the tube bundle. Note also that the ΔT for the 1st Effect was based on totally evaporating 1st Effect service. It was not felt that enough of the tube surface was used for preheating service to warrant the downward adjustment of the 1st Effect ΔT .

It is estimated that these coefficients are good at the 90% confidence level to about $\pm 5\%$. Increased precision would require more accurate data gathering equipment than was available during the runs.

TABLE G-1-b

DEVELOPMENT RUN 15 & 16

OVERALL EVAPORATIVE HEAT TRANSFER COEFFICIENTS

Btu/hr., °F, Ft.²

DETERMINED ON A SALT BALANCE BASIS

Heat & Material Balance No.	Average Development Run 14	15-1 5/2/68	15-3 5/28/68	15-4 6/11/68	15-5 7/1/68	15-6 7/22/68	15-7 8/7/68	15-8 9/12/68	Average Development Run 15	16-1 1/8/69	16-2 3/7/69	16-3 5/2/69	Average Development Run 16
Heat & Material Balance Date													
Evaporator													
1	730	-(1)	710	730	-(1)	-(1)	650	590	670	540	680	630	620
2	600	-(1)	640	630	500	600	540	750	610	640	560	670	620
3	530	530	610	570	610	600	670	680	610	640	-(1)	680	660
4	750	510	610	660	730	650	660	700	650	670	600	560	610
5	500	510	660	630	690	550	620	680	620	470	650	580	570
6	660	600	670	670	640	610	690	690	650	490	600	580	560
7	600	590	620	-(2)	660	600	540	650	610	500	480	540	510
8	700	500	550	-(2)	540	500	540	510	520	580	520	520	540
9	430	560	590	-(2)	640	660	550	660	610	610	590	550	580
10	360	440	380	420	460	320	380	420	400	480	340	360	390
11	390	-(2)	360	400	350	300	390	340	360	360	-(2)	400	380
12	350	-(2)	-(2)	330	270	260	460	290	320	530	-(2)	520	530
13	530	-(2)	-(2)	1300	1380	1080	1010	1140	1180	1050	1060	1020	1040
14	340	-(2)	380	400	450	410	400	430	410	260	320	310	300
15	330	420	350	330	340	320	350	380	360	200	310	290	270
16	430	330	410	400	310	320	340	420	360	230	350	310	300
17	350	310	390	360	310	310	290	330	330	230	290	280	270

(1) Out of line value discarded

(2) No salt balance data available

On Table G-1-c are shown background data for the heat and material balance tests for which overall heat transfer coefficients have been calculated. Note that during Development Run 16, the raw sea water was not blended and the first effect sump temperature was kept about 2°F lower than during Development Run 15. This was done to avoid further scaling of the tubes in effect No. 1 and to fix the operations of the ATU (Auxiliary Test Unit).

The heat transfer coefficients shown on Table G-1-b have been averaged and plotted against brine sump temperatures on Figure G-1-a. Shown on this figure are average heat transfer coefficients for Development Runs 14, 15, and 16. This represents operations at the Freeport Test Facility from September 1967 through May 1968. Note that the heat transfer coefficients for the No. 13 effect double fluted tube bundle are not shown on this graph.

On Figure G-1-a, the average data points for the even numbered evaporators have been identified. Note that Development Runs 14 and 16 represent, in general, cold weather operations; Development Run 15 represents mostly summer operations. Regarding the individual plots:

- 1) As expected the overall heat transfer coefficient trends up with temperature. The extreme variability of the heat transfer coefficients for the first four effects in Development Run 14 is probably due to sampling and analytical problems rather than erratic evaporator performance.
- 2) In the first four, highest temperature, evaporators the heat transfer versus brine boiling point trend seems to fall off. This could be in part due to the more open tube configuration in these bundles since some tubes were removed when the module was installed. This open tube configuration would have the most influence in the first few effects where the steam specific volumes and the steam velocities are the lowest. Also contributing to the lower evaporative heat transfer coefficients in the hotter effects is the inside tube physical fouling probably due to poor distribution that has been observed in these effects.
- 3) It is difficult to base any generalizations about venting on these data. Most of the time large changes in vent rate and position did not significantly affect the evaporative heat transfer coefficient.

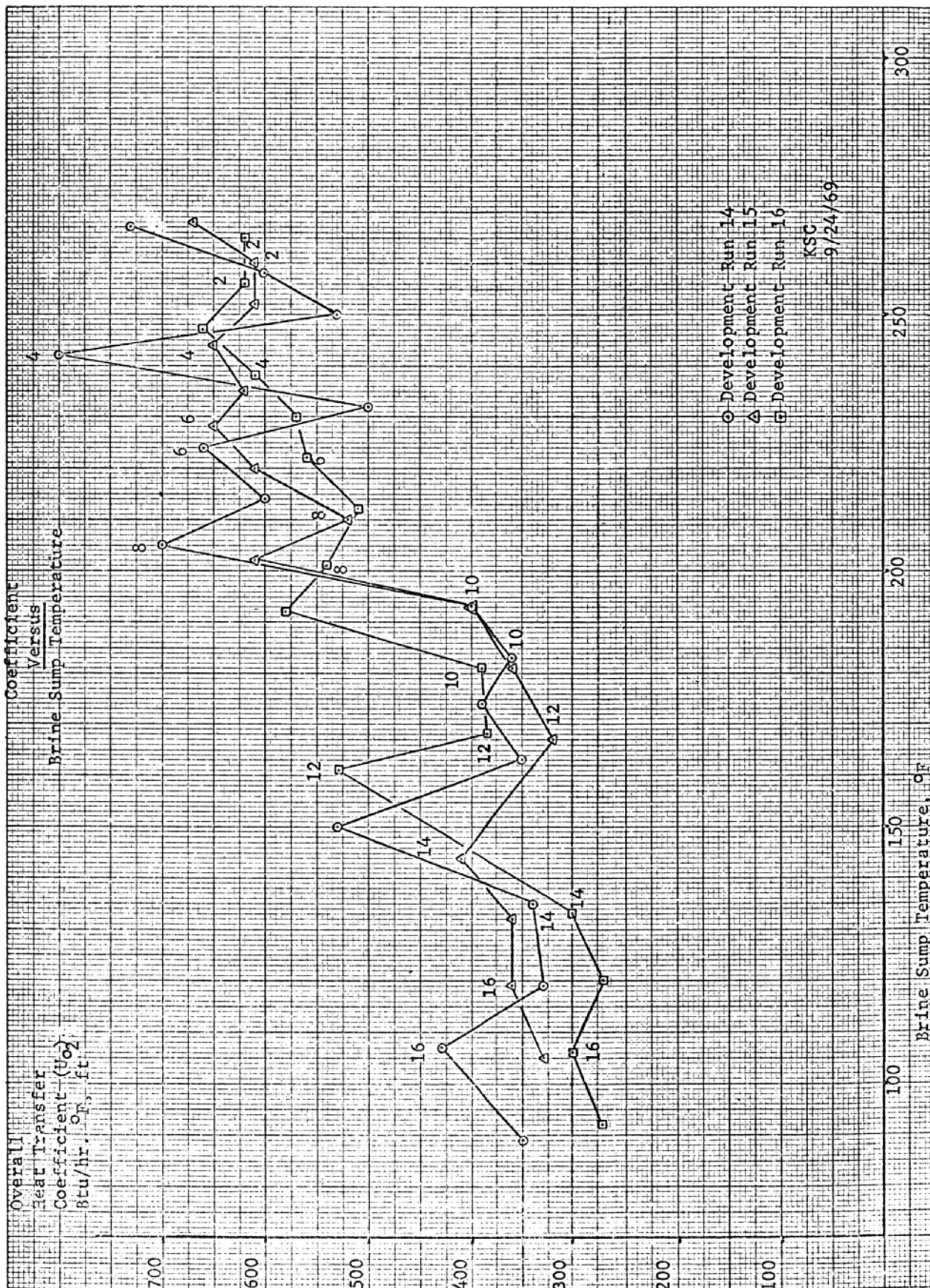
Table G-1-c

Heat and Material Balance Background Data

	Date	Temperatures, °F #1 Sump	#17 Sump	ΔT	Salinity WT% Total		#1 Effect	#17 Effect	Brine into #1 Effect	Brine into #17 Effect	#1 Effect #/hr.	Calculated Values	
					Dissolved Solids							Net Product Rate by Extraction MMGPD	GPM Water of Flashing Range
					Brine into #1 Effect	Brine into #17 Effect							
Development Run 15													
15-1	5/2/68	267.1	99.4	167.7	3.51			10.0	479			0.90	3.7
15-3	5/28/68	268.2	105.3	162.9	3.60			10.5	471			0.89	3.8
15-4	6/11/68	267.8	108.0	159.8	3.59			10.4	483			0.91	4.0
15-5	7/1/68	267.8	107.1	160.7	3.32			10.1	449			0.86	3.7
15-6	7/22/68	267.9	108.9	159.0	3.66			10.0	479			0.87	3.8
15-7	8/7/68	267.7	112.9	154.8	3.63			9.7	479			0.86	3.9
15-8	9/23/68	267.2	105.5	161.7	3.51			10.6	495			0.95	4.1
Development Run 16													
16-1	1/8/69	265.7	91.9	173.8	2.79			8.9	470			0.99	3.9
16-2	3/7/69	265.1	91.2	173.9	2.36			7.8	488			0.98	3.9
16-3	5/2/69	266.0	98.3	167.7	2.26			7.4	472			0.94	3.9

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FIGURE G-1-a
Evaporator Average Overall Heat Transfer



- 4) It can be seen that the average coefficients for Effects 10 through 12 are considerably out of line on the low side relative to the coefficients for the other effects. This is most probably due to the fact that the efficiency of these effects was significantly lowered by air leakage. This was demonstrated during the run by the fact that the condensate outlet temperatures of these three effects were significantly lower than the steam chest inlet temperatures for the same effects. This condensing temperature depression is generally attributed to the reduced partial pressure of the steam caused by the presence of significant quantities of noncondensables. The heat transfer coefficients for these three effects were low in both the feed forward and the sump-to-sump modes of operation. For some unknown cause, the apparent performance of the No. 12 Effect bundle improved significantly during Development Run 16; at the same time the apparent performance of the adjacent No. 13 effect double fluted tube bundle declined.
- 5) The performance of the modular low temperature effects declined significantly during Development Run 16.

There are two possible causes for the decline in performance for Effect No. 14. First, a reddish brown deposit has accumulated on the inside of these stainless steel tubes. Secondly, at the end of the run, many of the tubes in this bundle were found to contain gypsum scale. The most probable cause of this scale is over-concentration due to insufficient feed. The combination of these two factors is enough to drive down the apparent performance of the No. 14 Effect bundle.

The reasons for the decline in performance of effects 15 through 17 are not known. There was no obvious sign of tube fouling on either the steam or brine side. Preliminary experiments did, however, show that the module performance could be seriously impaired by fairly low noncondensable inleakage. It is probable that the poor performance was caused by a combination of high noncondensable loading and inadequate venting.

- 6) Although the performance of individual evaporators changed from run to run, there does not seem to be any overall or average decline in total plant performance between Development Run 15 and Development Run 16.
- 7) Attached in Table G-1-d are data regarding the difference

TABLE G-1-d
OBSERVED VERSUS EQUILIBRIUM BOILING POINT RISE

HEAT & MATERIAL BALANCE

16-3

Effect	Observed Brine Sump Temperature (1) °F	Observed Sump Vapor Temperature (1) °F	Apparent Boiling Point Rise °F	Brine Total Dissolved Solids Wt%	Equilibrium Boiling Point Rise From Correlation (2) °F
1	266.0	246.6	1.4	2.39	0.8
2	256.9	255.7	1.2	2.52	0.8
3	248.7	247.1	1.6	2.64	0.8
4	239.6	238.6	1.0	2.80	0.8
5	231.4	230.6	0.8	2.96	0.9
6	223.4	222.1	1.3	3.13	0.9
7	214.3	213.5	0.8	3.34	0.9
8	204.7	203.9	0.8	3.55	1.0
9	195.7	194.2	1.5	3.79	1.0
10	184.3	183.4	0.9	4.05	1.0
11	172.8	171.5	1.3	4.38	1.1
12	165.4	161.2	4.2	4.02	1.1
13	153.3	151.5	1.8	5.08	1.2
14	140.4	138.6	1.8	5.53	1.2
15	126.5	124.9	1.6	6.07	1.3
16	113.1	110.7	2.4	6.65	1.4
17	98.3	97.6	0.7	7.38	1.4

(1) Certified mercury filled glass thermometer, I 0.2°F

(2) OSW 11.51, Distillation Digest: Volume 2, Number 1, Page 64.

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in temperature between the sump vapor and the sump liquid that were observed during heat and material balance 16-3. Also shown are the equilibrium boiling point rise values determined from the correlation at the indicated brine temperature and brine salinity. It should be noted that there is some additional loss in temperature level over and above the calculated equilibrium boiling point rise. The heat transfer coefficients that are shown in this report are all calculated on the basis of observed sump temperatures and observed chest temperatures. The designer should be aware that in some cases there might be additional loss in temperature level between effects over and above the equilibrium boiling point rise loss.

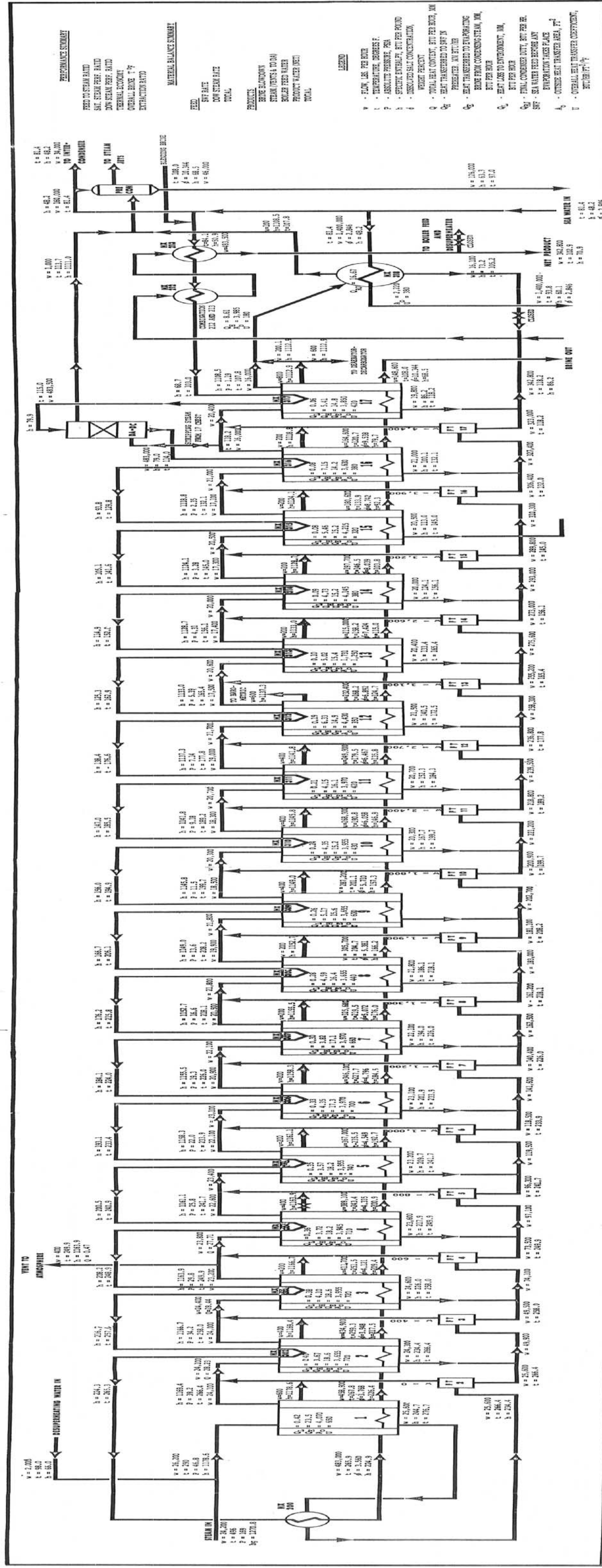
b. Detailed Heat & Material Balances

The performance of the process is most clearly indicated by reference to the complete, detailed heat and material balance data on Figures G-1-b and for heat and material balance 15-4 in Development Run 15 and Figure G-1-c for heat material balance 16-3 in Development Run 16.

These heat material balance flow sheets were generated from the raw data taken during the heat and material balance runs. The plant was balanced on an effect by effect basis starting at the hot end and working backward to the product end. Heat losses were estimated as a function of evaporator operating temperature and insulation competency. Heat losses were recalculated and the balance calculations repeated until a compatibility of $\pm 1\%$ on total water production was obtained.

These calculations were made using standard steam and brine enthalpy values. Actual observed temperatures were used. For balancing purposes, the brine enthalpy values were carried out to the nearest 0.1 Btu/lb; this is one digit beyond the actual accuracy of the enthalpy data.

On occasion, attempts have been made to determine the evaporator heat loss by difference between the heat in and the heat out. It is impossible to accurately determine the heat loss by difference;



PERFORMANCE SUMMARY

FEED TO STEAM RATIO
SAL. STEAM FEED RATIO
DOW STEAM FEED RATIO
THERMAL EFFICIENCY
OVERALL HEUR T²/°
EXTRACTION RATIO

MATERIAL BALANCE SUMMARY

FEED
STEAM RATIO
DOW STEAM RATIO
TOTAL
PRODUCTS
STEAM RATIO
DOW STEAM RATIO
TOTAL

HEAT BALANCE SUMMARY

FEED
STEAM RATIO
DOW STEAM RATIO
TOTAL
PRODUCTS
STEAM RATIO
DOW STEAM RATIO
TOTAL

- 1 - FEED, LBS PER HOUR
- 2 - TEMPERATURE, DEGREES F.
- 3 - ABSOLUTE PRESSURE, PSIA
- 4 - SPECIFIC ENTHALPY, BTU PER POUND
- 5 - WEIGHT PERCENT
- 6 - INSULATED HEAT CONDUCTION
- 7 - TOTAL HEAT CONTENT, BTU PER HOUR, MM
- 8 - HEAT TRANSFERRED TO STEAM IN REBOILER, MM BTU PER HOUR
- 9 - HEAT TRANSFERRED TO CONDENSING STEAM, MM, BTU PER HOUR
- 10 - HEAT TRANSFERRED TO STEAM IN REBOILER, MM BTU PER HOUR
- 11 - FINAL CONDENSING DUTY, BTU PER HOUR
- 12 - STEAM FEED BEFORE ANY EVAPORATION TAKES PLACE
- 13 - OVERALL HEAT TRANSFER COEFFICIENT, BTU PER HOUR PER SQUARE FOOT

the heat loss is simply too small relative to the total quantities of heat going into and out of the system.

The calculation technique was changed slightly for heat and material balance 16-3. During the actual data gathering, it was noted that as much as 15,000 lbs/hr of purge and seal water was used but could not be accounted for. Experiments showed that most of this was going into faulty mechanical seals on the brine sump pumps. The heat and material balance calculations were adjusted to show the influence of additional water being added through the pump seals and the level control purge systems. More water was added to the effects with pumps that contained mechanical seals.

c. Preheater Performance

The attached Tables G-1-e & f show heat transfer coefficients that have been calculated for the preheat exchangers using data taken at various times during Development Runs 15 and 16. The surface area data used to make these calculations are shown in Table E-1.

Some discussion of the precision of these coefficients is in order. These coefficients are calculated using the standard formula:

$$U_o = \frac{(C_p)(W_b)(T_b)}{A_o \frac{GTD-LTD}{\ln \frac{GTD}{LTD}}}$$

Where: U_o = Overall Heat Transfer Coefficient Btu/hr.,
ft.², °F
 C_p = Brine Specific Heat, Btu/lb., °F
 W_b = Brine Rate, lbs./hr.
 T_b = Brine Temperature Increase Across the
Exchanger, °F

Preheat Exchanger - Heat Transfer Coefficients (U_o) Btu/hr, oF , ft^2

LEGEND:

N = LTD Negative.

Table G-1-f

Preheater Exchangers - Heat Transfer Coefficients (U_o), BTU/hr, °F, ft²

Development Run 16

H&MB 1 1/8/69	H&MB 2 3/2/69	H&MB 3 5/2/69	Tube Brine Velocity ft/sec	Ave. Run 14	Ave. Run 15	Ave. Run 16
HX-302 (a)	360*	420*	2.7	400*	290*	390*
HX-302 (b)	430	420	3.1	530*	480*	470*
HX-303 (a)	490*	310*	3.6	860*	540*	570*
HX-303 (b)	350	480	3.4	490	630*	450
HX-304	150	130	2.1	310*	370*	220*
HX-305	250	260*	1.6	250	340*	270*
HX-306 (a)	460*	430*	2.9	475*	370*	460*
HX-306 (x)	780	930	8.6	400	330	880(1)
HX-307	320	270	2.7	420	400	310
HX-308	270	300	2.7	390	410	300
HX-309	330	280	2.7	355	410	320
HX-310	270	300	2.8	330	280	310
HX-311	790	260	2.8	380	290	380
HX-312	540*	540*	6.5	430	410*	540
HX-313	380	360	9.9	725	460	380
HX-314	310	280	9.9	445	380	330
HX-315	390	350	8.3	440	350	390
HX-316	240	230	4.8	350	330	300
HX-317	320	320	4.8	435*	340	320

* = LTD Less than 2°F.

N = LTD Negative

(1) Changed to vapor sheer preheater.

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A_o = Heat Transfer Surface (based on tube outside area), ft.²
 GTD = Greatest Temperature Difference Between Condensing Steam and Brine, °F
 LTD = Least Temperature Difference Between Condensing Steam and Brine, °F

It can be seen that in this application the ΔT_b and (GTD-LTD) terms cancel out. C_p , W_b , and A_o can all be determined with second place accuracy. Thus it can be seen that the uncertainty in the heat transfer coefficient value is going to be very sensitive to any uncertainty in GTD and LTD. Most of the brine temperatures were determined using "yellow back" thermometers with a stated accuracy of $\pm 1^\circ\text{F}$ and a reading accuracy of $\pm 0.2^\circ\text{F}$. The steam temperatures were taken with calibrated thermometers with a stated accuracy of either $\pm 0.25^\circ\text{F}$ or $\pm 0.1^\circ\text{F}$. It can be demonstrated that when the LTD value gets below 2.0°F , there is considerable uncertainty in the calculated heat transfer coefficient. On the attached table we have put an asterisk where the LTD is 2.0 or less to warn the reader that considerable uncertainty exists.

For comparison purposes, on Table G-1-f the average overall heat transfer coefficients from Development Run 14 and the average brine velocities are also shown on the table.

Regarding the data:

- (1) The average overall heat transfer coefficient for HX-313 remains high. Because of its position in the path of the steam coming to the module from Effect 12, this exchanger sees a lot more turbulence on the steam side than the other exchangers.
- (2) The performance of HX-312 improved drastically between Heat & Material Balance Runs 15-3 and 15-4. Prior to Run 4, this exchanger's bottom vent had been valved in to the No. 12 Effect chest vent. Just before Heat & Material Balance 4, this was changed and the HX-312 bottom chest vent was tied into the vent line from HX-318 and was isolated from the chest vent from the No. 12 Effect. Obviously the No. 12 Effect chest vent had been back-pressuring the HX-312 vent and preventing adequate flow.

After the vent change, HX-312 showed a coefficient in excess of 500 Btu/hr., °F, ft.², which was greater than most of the other exchangers. This superior performance can probably be attributed to the fact that HX-312 is a four-pass exchanger and the increased turbulence on the brine side has served significantly to increase the overall heat transfer coefficient.

- (3) The heat exchanger vents on the module were opened wide for Heat & Material Balance No. 15-5; there was no significant or apparent increase in heat transfer coefficient that could be attributed to the opening of these external vents.

Actually, positioned as they are in the path of the steam flow to the main effect bundle, these module preheat exchangers probably are not helped by external vents. The steam probably flows across these heat exchanger tube bundles driving the non-condensibles into the center of the bundle. An internal "vent tube" could be installed to remove non-condensibles from the center of the tube bundle.

The vents on exchangers HX-306b through HX-311 were gradually opened more and more during the run. There is no indication that the change in the vent setting caused any improvement in the operation of any of the heat exchangers.

- (4) Heat exchangers HX-310 and HX-311 performed quite poorly during most of Development Run 15. This is most probably due to the fact that there were serious vacuum leaks in this area all during the run and the non-condensibles that were introduced through these leaks served to vapor bind the heat exchangers and consequently lower their performance.
- (5) HX-307 has CDA-194 tubes. The heat transfer performance of this bundle does not differ (high or low) from exchangers HX-308 or HX-309. There is no reason to expect the CDA-194 tubes to perform significantly better from a heat transfer stand point than aluminum brass.
- (6) The brine velocities in the heat exchangers in the old plant are, for the most part, quite low. Higher brine velocities would undoubtedly result in better exchanger performance and in more efficient plant operation.
- (7) The performance of the heat exchangers in the module seems to be deteriorating with time. This is most probably due to tube fouling. These tubes should be chemically cleaned to determine if performance improves up to the original level.

d. Performance of Heat Rejection Condenser (HX-318)

Attached as Table G-1-g is a summary of the data that were taken and the calculations that were made on said data regarding the operation of the heat rejection condenser (HX-318) during Development Runs 15 & 16.

Table G-1-g

Heat Rejection Condenser Performance (HX-318)

Heat & Material Balance No. →		Development Run 15			Development Run 16			Approximate Design					
Date		1 1968 → 5/ 2	2 5/28	3 6/11	4 7/ 1	5 7/22	6 8/ 7		7 9/12	1 1/ 8	2 3/ 7	3 5/ 2	
I.	DATA												
II.	CALCULATED VALUES												

The overall heat transfer coefficient ranged from a low of 160 to a high of 450 Btu/hr., °F, ft.². The estimated accuracy of these heat transfer coefficients is $\pm 5\%$. All observed coefficients are below the design coefficient.

According to the manufacturer, the non-condensable cooling section of the heat rejection condenser should see the coldest cooling brine. This exchanger (HX-318) is designed for top inlet; changing the inlet point from top to bottom at various times during Development Run 15 did not seem to significantly affect the overall heat transfer coefficient.

The steam approach temperature averaged about 25°F to the inlet cooling brine. This approach remained constant over an almost 15°F change in inlet cooling water temperature.

The performance of the condenser fell off significantly during cold weather operations; this severely limited the ability of the plant to produce more water when the seawater temperature dropped down in the winter. The vacuum producing equipment was not to blame, it pulled down to an extremely low backpressure (0.7 m.Hg) during these low temperature operations.

G. RESULTS AND COMPARISONS (CONTD)

2. SPECIAL HEAT TRANSFER SURFACE PERFORMANCE

a. Effect 13, Double Fluted Tube Bundle

The first enhanced surface tubes to be used at the Freeport Test Facility were installed, as a complete bundle, in the 13th Effect during the shutdown between Development Runs 14 and 15. The tubes used had a General Electric Company double-fluted surface configuration. As in the 9th Effect, this bundle replaced one of a more conventional design. The new and old 13th Effect tube bundles are compared in Table G-2-a, below.

Table G-2-a

Comparison of Effect 13 Tube Bundles

	<u>Original</u>	<u>New</u>
Tube Surface	Smooth	Double-Fluted
Tube Material	90-10 Cu-Ni	CDA-194
Tube Length	22'-0"	11'-1"
Tube O.D.	2"	3 1/8"
Tube Wall	0.049"	0.065"
Number of Tubes	360	150 (1)
Bundle Area	4,140 ft. ²	1,722 ft. ²
Brine Distribution	SVL Weirs	Spray Nozzles
Water Box Style	Raised	Raised

(1) Plus one vent tube that was open at the top during Development Run 15.

The performance factors that were calculated for the double-fluted tube bundle during Development Runs 15 and 16 are shown graphically on Figure G-2-a and in tabular form on Table G-2-b. The calculation procedure that was used to calculate the overall heat transfer coefficients is shown in the Appendix, Section 2.

From the data, it can be seen that there was an initial 250% improvement in performance due to the enhanced surface tube installation. The heat transfer coefficient during Run 15 was fairly steady with no trend to decrease as the operation hours increased. During Run 16 the tube performance was generally lower than during Run. 15, by about 15%, and also a slight but gradual decrease was evident. To date, the bundle has seen over 6,200 operation hours in evaporation

TABLE G-2-b

NO. 13 EFFECT DOUBLE FLUTED TUBE PERFORMANCE

<u>Date</u>	<u>Plant Feed M lb/hr</u>	<u>No. 13 Chest °F</u>	<u>Q_t 10⁶ Btu/hr</u>	<u>Δ^T °F</u>	<u>U_o Btu/hr-ft²-°F</u>
<u>Run 14 Average</u>					
<u>Smooth Tube:</u>	485	158.5	16.2	7.8	500
<u>Run 15:</u>					
5-2-68	478	159.1	15.7	6.2	1,470
5-8-68	468	166.6	15.7	6.9	1,320
5-17-68	455	161.8	16.9	7.3	1,350
5-22-68	467	162.1	16.1	7.9	1,190
5-28-68	472	162.5	16.2	7.1	1,330
6-11-68	484	165.4	17.2	7.2	1,390
6-25-68	460	164.2	17.8	6.9	1,500
6-28-68	470	166.2	15.6	6.3	1,440
7-1-68	450	164.8	17.8	6.5	1,590
7-5-68	472	166.3	15.3	6.9	1,290
7-10-68	457	166.8	17.0	7.2	1,370
7-22-68	480	164.1	16.6	7.2	1,340
7-30-68	481	165.4	16.2	6.8	1,380
8-3-68	480	166.1	17.1	6.7	1,480
8-7-68	480	167.6	17.8	6.7	1,540
8-12-68	481	166.3	17.4	6.7	1,510
8-15-68	479	168.4	16.8	6.5	1,500
8-27-68	482	169.0	13.5	6.4	1,220
9-12-68	496	163.2	15.9	7.4	1,250
9-20-68	409	164.4	15.0	6.5	1,340
<u>Run 15 Average: 470</u>					
		165.0	16.4	6.9	1,380

TABLE G-2-b (CONTD)

No. 13 EFFECT DOUBLE FLUTED TUBE PERFORMANCE

<u>Date</u>	<u>Plant Feed 10³ lb/hr</u>	<u>No. 13 Chest °F</u>	<u>Q_t 10⁶ Btu/hr</u>	<u>ΔT °F</u>	<u>U_o Btu/hr-ft²-°F</u>
<u>Run 16:</u>					
1-8-69	441	158.3	16.9	8.4	1,170
2-18-69	492	155.2	17.4	8.4	1,200
2-24-69	492	156.3	17.6	8.0	1,280
2-27-69	492	155.6	18.5	8.1	1,330
3-7-69	489	156.6	17.6	8.5	1,200
3-18-69	489	154.9	14.4	8.6	970
3-27-69	491	155.8	14.5	8.2	1,030
4-17-69	480	160.5	15.5	7.8	1,150
5-2-69	473	161.1	13.5	7.8	1,010
<u>Run 16 Average:</u>	<u>482</u>	<u>157.1</u>	<u>16.2</u>	<u>8.2</u>	<u>1,150</u>

service and has a 2:1 performance ratio compared to the original smooth tube bundle. The initial performance of the double-fluted tube bundle was reported⁽¹⁾ at the recent OSW enhanced surface symposium.

During Development Run 15, two major changes were made to the operations of Effect 13; neither one of them seemed significantly to affect the evaporative heat transfer coefficient. Looking at Figure G-2-a, from the end of May to the beginning of July, the external vent pipe was fully opened allowing an additional 300 lb/hr of steam to vent from the chest. There was no significant change in the average heat transfer coefficient during this time. For most of July, the tube feed rate was cut down to about 900 lb/hr per tube. Again there was no significant change in the average heat transfer coefficient.

Operations of the 13th Effect were not uniformly smooth during Development Run 15. Because of pump trouble there were several occasions on which tube feed was lost for a significant amount of time. At the beginning of the run, several of the distribution nozzles popped out of place; a shutdown was taken so they could be replaced and restrained with a wire mesh grid. Early in the run a small continuously operated vent line was installed between the dome of Effects 13 and 14. The purpose of this was continuously to bleed non-condensables out of the flooded water box of the 13th Effect.

During Run 16, operations were generally constant, with tube feed provided at 1,100 lbs/hr on recycle, evaporator feed sump-to-sump from the 12th Effect while venting full from the chest at about 500 lbs/hr with the dome vent open.

The double-fluted tube bundle was probably the single most important piece of equipment tested in Development Runs 15 and 16. Regarding future runs, it is suggested that:

- 1) The tubes be inspected carefully for any sign of fouling on either the evaporating or condensing side.
- 2) Specimen tubes be pulled and tested in laboratory evaporators for a direct comparison with the performance of clean or reference tubes.
- 3) The #13 double-fluted tube bundle be run with open top S.V.L. weirs at different tube feed rates to determine the relative performance between nozzle type distributors and the S.V.L. Distributors.

(1) "Large Plant Performance Testing of a Commercial Sized Double-Fluted Tube Bundle"; Keith S. Campbell and J. P. Lennox, March, 1969.

FIGURE G-2-a

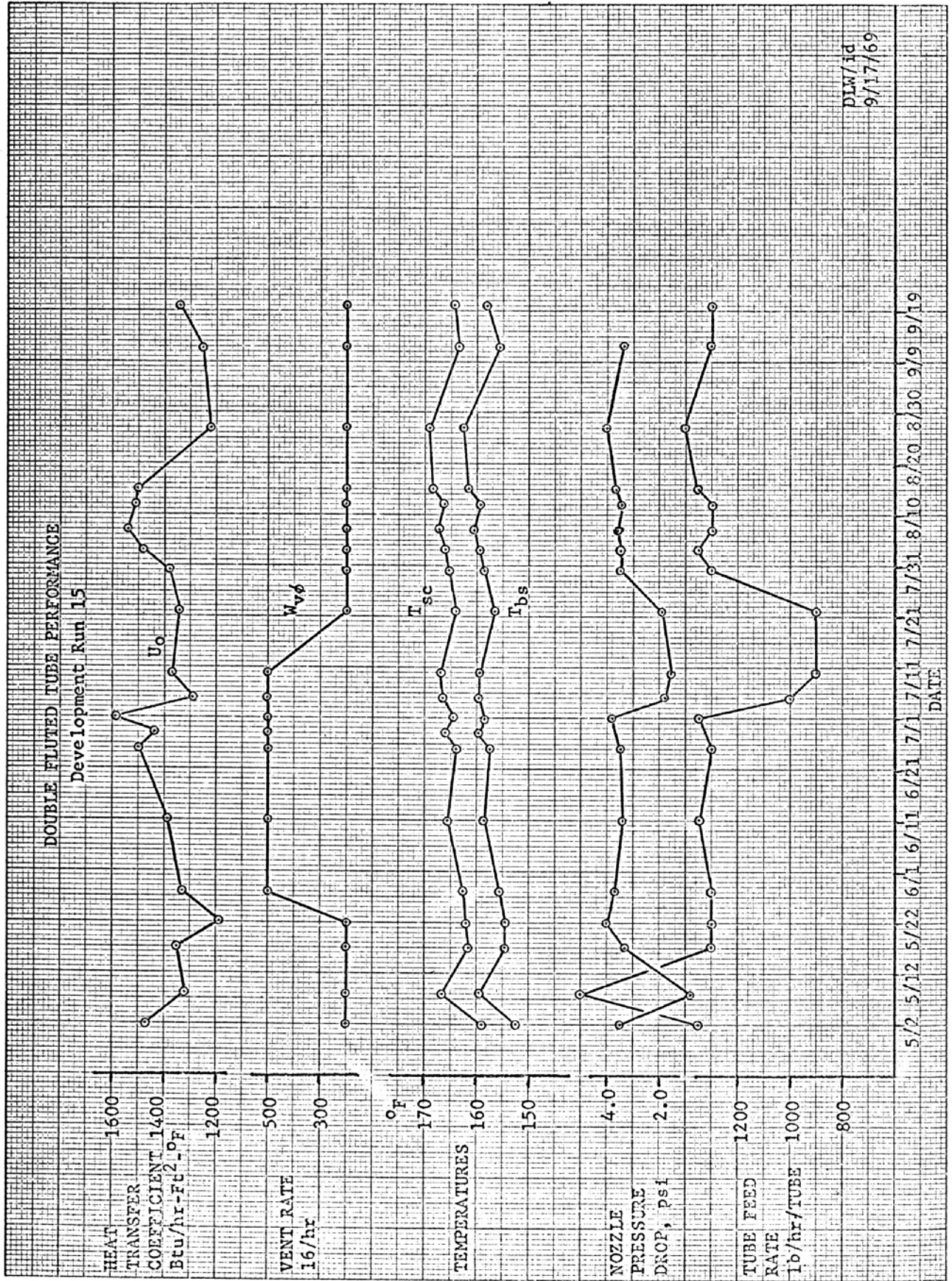
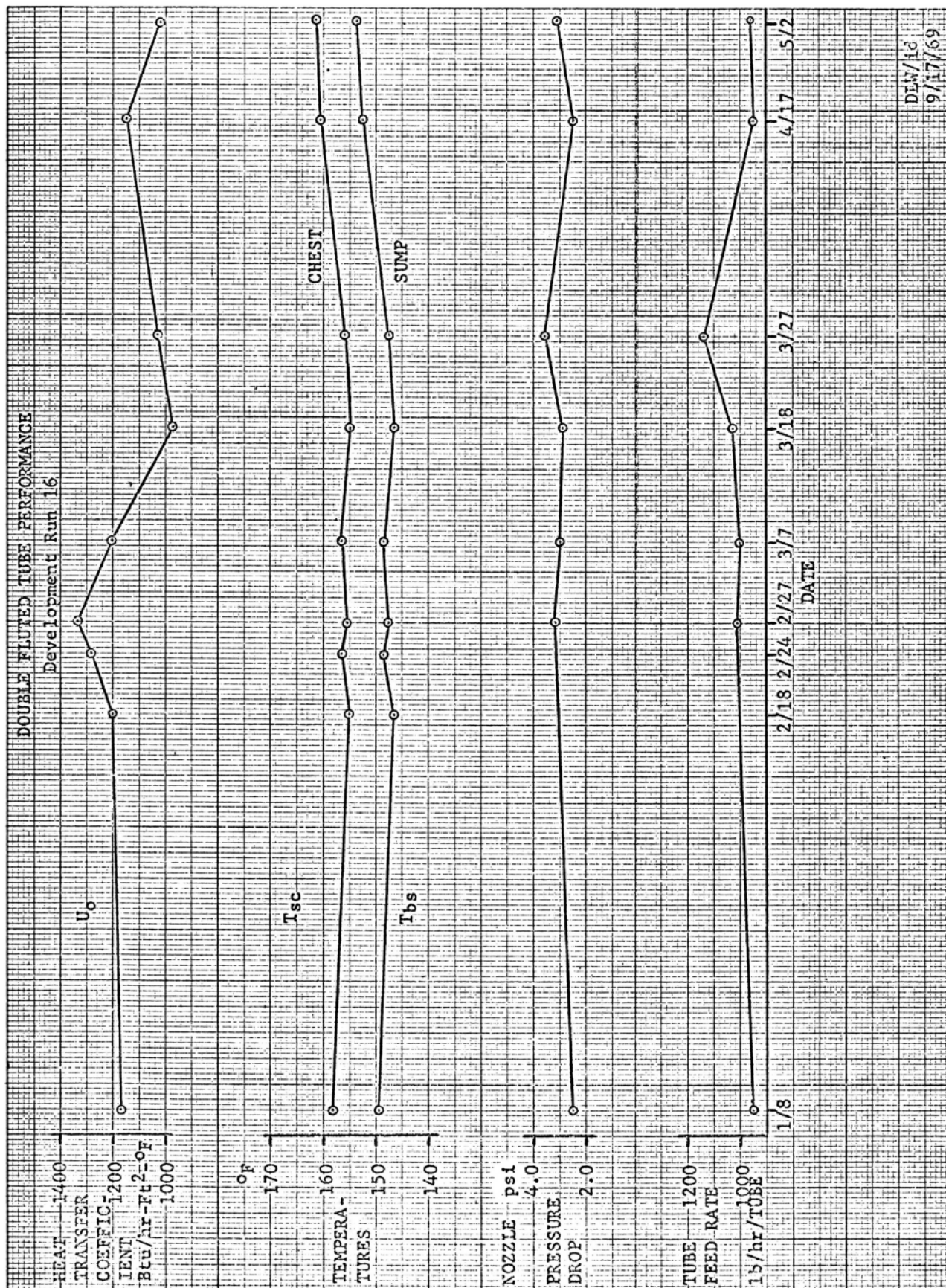


FIGURE G-2-a (cont'd)



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- 4) The bundle be chemically or mechanically cleaned to see if it can be restored to its original performance level.
- 5) That the existing #13 Effect temperature measurement equipment either be recalibrated or replaced with resistance thermometers to achieve an accuracy of $\pm 0.1^{\circ}\text{F}$ compared to the present estimated accuracy of $\pm 0.5^{\circ}\text{F}$.

b. Effect 9, Titanium Tube Bundle

During the modification period between Development Runs 14 and 15, the water box and tube bundle in Effect No. 9 was replaced with a new modified assembly. This consisted of a raised water box and polyester dome for improved brine distribution, and thin wall commercially pure titanium tubes for new heat transfer area. The tubes were laid out to form steam corridors into the bundle and the center tube was a vent tube. A comparison of the new equipment with the original is shown in Table G-2-c below:

Table G-2-c

Comparison of #9 Effect Tube Bundles

	<u>Original</u>	<u>New</u>
Tube Material	Admiralty	Titanium
Tube Length	24'	24'
Tube O.D.	2"	2"
Tube Wall	0.065"	0.020"
Number of Tubes	303	290
Bundle Area (Ao)	3,810 ft. ²	3,640 ft. ²
Brine Distribution	V-Notch Weirs	SVL Weirs
Water Box Style	Flooded	Raised

Data was taken on this unit to evaluate its heat transfer performance approximately every two weeks during its installed life. After 6,200 operation hours, the tubes were inspected, with special care taken to note any sign of stress corrosion cracking at the tube roller expanded area. At this time, no corrosion of any type was evident and the brine side of the tubes appeared in an "as new" condition. It is possible that the carbon steel tube sheet provides some "cathodic protection" for the titanium tubes.

Data taken around #9 Effect during Development Runs 15 and 16 are shown along with some calculated values on Table G-2-d. The overall evaporative heat transfer coefficients (U_o) were calculated on a salt balance basis.

TABLE G-2-d

DATA AND CALCULATED VALUES FOR #9 EFFECT (TITANIUM TUBES)

<u>Date</u>	<u>Plant Feed MPPH</u>	<u>No. 9 Feed MPPH</u>	<u>No. 9 Product MPPH</u>	<u>No. 9 Chest °F</u>	<u>Qt 106 Btu/hr</u>	<u>ΔT °F</u>	<u>Uo Btu/hr-ft²-°F</u>
<u>Development Run 14 (Admiralty Tubes)</u>							
Average	---	---	---	---	---	---	470
<u>Development Run 15 (Titanium Tubes)</u>							
5/2/68	479	322	18.8	203.2	15.9	7.7	570
5/28/68	471	308	17.7	208.7	15.1	7.1	580
6/11/68	483	319	18.4	208.2	15.7	7.1	600
7/1/68	449	306	18.5	210.1	15.9	6.8	640
7/22/68	479	317	17.7	209.4	15.1	6.3	660
8/7/68	479	323	16.5	210.0	13.9	6.9	550
9/12/68	495	327	19.7	210.6	16.9	7.0	660
<u>Development Run 16 (Titanium Tubes)</u>							
1/8/69	470	293	19.4	200.3	16.8	7.2	640
2/12/69	492	329	26.7	201.8	23.6	8.4	770
3/7/69	489	314	20.2	200.9	17.3	8.6	550
3/14/69	490	315	19.8	198.8	16.9	8.0	580
3/28/69	490	327	18.6	201.1	15.7	8.3	520
4/16/69	480	312	18.5	205.0	15.7	8.1	530
5/2/69	472	300	18.9	203.9	16.1	8.2	540

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9/16/69

The new titanium bundle has performed significantly better than the old admiralty bundle; the overall heat transfer coefficient appears to be slowly decreasing with time.

c. Effect 14, 316 Stainless Steel Tube Bundle

The number 14 Effect in the five effect module was equipped with thin wall 316 Stainless Steel Tubes as an original modular component. The purpose of this installation was to expose 316 Stainless Steel to actual saline water evaporator environment, and evaluate this material as a heat transfer medium. Special emphasis was placed on the post development run inspections to determine the condition of the alloy. The evaporator dome flange and tube sheet were 316 SS clad; the tube sheet hub, suction and discharge piping and SVL distributors were solid 316 SS. Bundle information is:

Tube Surface	Smooth
Tube Material	316 SS (Welded)
Tube Length	22'
Tube O.D.	2"
Tube Wall	0.035"
Number of Tubes	359
Bundle Area (Ao)	4,136 ft. ²

Data to evaluate the performance of the bundle was taken as a part of total plant heat and material balance information and reported with that data. A summary of performance is shown in Table G-2-e.

Table G-2-e

Effect 14 - 316 Stainless Steel Tube Performance

	<u>U (Btu/hr-ft²-°F)</u>	<u>Cumulative Operating Hours</u>	<u>Remarks</u>
<u>Run 14</u>			
12-5-67	360		Low performance due to condensate level in chest
1-5-68	350		
1-11-68	360		
Average	<u>350</u>	2,900	
<u>Run 15</u>			
5-28-68	380		Flash tank level glass installed
6-11-68	390		
7-1-68	450		
7-22-68	410		
8-7-68	400		
9-12-68	430		
Average	<u>410</u>	6,000	

Table G-2-e (cont'd)

Effect 14 - 316 Stainless Steel Tube Performance

	<u>U (Btu/hr-ft²-°F)</u>	<u>Cumulative Operating Hours</u>	<u>Remarks</u>
<u>Run 16</u>			
1-8-69	260		Over 6% tubes scaled. Flash tank level glass leak.
3-7-69	320		
5-2-69	310		
Average	300	9,100	

The performance improvement shown during Run 15 over Run 14 was due to the condensate transfer pipe line size increase from 8 to 10 inches. This allowed a lower operating level in the flash tanks and consequently allowed Effect 14 to operate with a completely drained steam chest.

Following Run 16, six percent of the tubes in No. 14 were found scaled with calcium sulfate (gypsum); also, a large air leak was discovered in flash tank 14. The poor performance during Run 16 is attributed to a combination of these two items.

At each inspection, a continuous reddish-brown deposit has been noted in these tubes. This could be another cause of the worsening performance of the stainless steel tube bundle.

The cause of the gypsum scale that deposited during Development Run 16 has not been exactly determined. It could be due to insufficient top water box feed in this sump-to-sump effect; the evaporator bottoms pump P-24 is set up to provide feed forward brine to Effect 15 and total recycle to Effect 14.

An inspection of the tubes was performed on June 25, 1969. The inspection was made in the evaporator steam chest and sump. The condensing side tube wall, tube extension below the lower tube sheet, and the stainless steel-carbon steel tube sheet interface were examined. No evidence of tube failures or stainless steel corrosion were noticed. However, there was some discoloration due to iron oxide deposits on the condensing side.

d. Preheater HX-306(x), Experimental Horizontal Enhanced Surface Vapor Shear Tube Bundle

During the shutdown between Development Runs 15 and 16, a new experimental preheat exchanger was installed in the plant. It was designed and furnished to the plant by ORNL, and replaced an original conven-

tional vertical tube preheater in the main plant preheating circuit. A comparison of the two preheaters is shown in Table G-2-f.

Table G-2-f
Comparison of Preheater Data

	<u>Conventional</u>	<u>Experimental</u>
Vessel Number	HX306b	HX306X
Surface Area	1,791 ft. ²	220 ft. ²
Tube Surface	Smooth	Spiral Grooved (Rope)
Number of Tubes	342	104
Tube O.D.	1"	1"
Tube Wall	0.049"	0.035"
Tube Length	20'	8'
Tube Material	Al-Brass	Cu-Nl, 90-10
Vessel Orientation	Vertical	Horizontal
Passes Brine Side	2	1
Passes Steam Side	N/A	3

The vessel installation was designed to allow two modes of operation, full condensing and flow thru - both being conditions of the steam side.

From the data presented in Table G-2-g, it can be seen that the heat transfer coefficient has averaged 970 Btu/hr-ft.²-°F for 3,100 hours of operation, and has not decreased with time. It has not performed as well as the design, but considerably above the 370 Btu/hr-ft.²-°F coefficient of the conventional preheater which it replaced. This is due to:

- 1) The higher velocity on the brine side.
- 2) The enhanced surface tubes.
- 3) The horizontal tube configuration.
- 4) The steam side baffling.

TABLE G-2-g

HX 306(X) PERFORMANCE

<u>Date</u>	<u>SWF MPPH</u>	<u>SWF In °F</u>	<u>SWF Out °F</u>	<u>Stm. Chest °F</u>	<u>LMTD °F</u>	<u>Qt MMBtu/hr</u>	<u>Uo Btu/hr-ft²-°F</u>
Design (2)	500	218	224	230	8.7	3.0	1,100
1-1-69 (1)	450	216.5	221.5	229.0	9.8	2.18	1,010
1-7-69 (1)	440	217.3	222.5	230.3	10.2	2.23	990
1-8-69 (1)(3)	441	216.2	221.4	230.3	11.2	2.23	910
2-12-69 (2)	492	214.9	218.8	226.9	8.3	1.87	1,000
3-7-69 (1)(3)	489	216.0	219.9	228.8	10.7	1.87	790
3-12-69 (1)	489	213.1	217.8	227.0	11.2	2.25	910
3-25-69 (1)	491	214.8	219.8	227.4	9.9	2.39	1,100
4-16-69 (1)	480	219.2	223.7	230.5	8.9	2.12	1,080
5-2-69 (1)(3)	472	217.8	222.4	230.5	10.2	2.12	950

Notes: (1) Full Condensing Mode
 (2) Flow Thru Mode
 (3) Temperatures Taken with Laboratory Grade Thermometers

DLW:alb
 9/16/69

G. RESULTS AND COMPARISONS (CONTD)

3. REVISED METHODS OF FLUID TRANSFER

a. Sump to Sump Brine Transfer

The two most generally favored methods of transferring brine from evaporator to evaporator in a VTE Plant are:

- (1) Feed Forward. In this mode of operation, brine is pumped from the sump of the evaporator to the top water box of the next evaporator. In the top water box, the brine flashes down to the lower temperature and pressure. A level controller is required to regulate the brine level in the previous sump.
- (2) Sump-to-Sump. In this mode of operation, the effect pressure differential is used to force brine from the sump of one effect into the sump of the next effect. The brine being transferred flashes down to the pressure and temperature existing in the next sump. To regulate the level and protect the pump, a standpipe is usually put inside the sump of the evaporator from which the brine is being transferred. To avoid turbulence and direct the flashing brine, a flash cap is usually placed on the outlet of the sump-to-sump transfer line inside of the lower temperature effect. Equilibrium brine is pumped from the sump up to the upper water box of the same effect. No flashing occurs in the upper water box. No level control is needed on the brine pump discharge.

Elimination of flashing in the top water box serves to improve brine distribution to the tubes and lessen the pressure drop across the tube bundle. A control system is eliminated. The system is self-regulating. In modular effects, additional savings can be realized if it becomes possible to transfer the brine internally through the intereffect bulkhead.

To further test out the process dynamics of sump-to-sump brine transfer, work was done during the Development Run 14 - Development Run 15 shutdown to equip evaporators 10 through 14 with sump-to-sump transfer piping. The sump-to-sump system was put into operation early in Development Run 15.

The sump-to-sump transfer piping operated satisfactorily. Data taken during Development Run 15 indicated that there was no flashing in the sump-to-sump line between effects 10 and 11; and between effects 11 and 12. Aside from minimum line loss and elevation loss, all the inter-effect pressure drop was taken across the inlet nozzle to the low pressure effect.

The top water boxes of the effects operating in the sump-to-sump mode were observed several times during operations. Under sump-to-sump operations with no flashing taking place in the top water box, very little turbulence was noted. Brine distribution seemed uniform. Considerable turbulence due to flashing was observed in other top water boxes that were operating in the feed forward mode.

The sump-to-sump system was studied to see if any by-passing was occurring. Composite samples of circulating brine from the adjacent evaporators and the sump-to-sump line were taken at the same time; the samples were analyzed for chlorinity. If the sump-to-sump sample has the same chlorinity as the higher temperature adjacent effect, then no by-passing is occurring. The results of these tests and analyses are shown on Table G-3-a.

There is an elevation difference between the sump of Effect 12 and the sump of Effect 13 (the first modular effect) such that when operating in the sump-to-sump mode, an unusually high level is required in the sump of Effect 12 to force the brine into Effect 13. Preliminary data indicated that the temperature of the brine in the 12-13 sump-to-sump line was higher than the temperature of the brine at the discharge of No. 12 Effect recycle pump (p-22). Additionally, looking at the data in Table G-3-a, the brine in the 12-13 sump-to-sump line was more dilute than the brine being circulated to the top water box of effect 12. By-passing of brine from the 11-12 sump-to-sump line through the No. 12 Effect into the 12-13 sump-to-sump line was clearly indicated. This was possible because the sump-to-sump inlet and outlet nozzles in No. 12 sump are in close proximity and point at each other

TABLE G-3-a

SUMP-TO-SUMP CHLORINITY RESULTS
wt. % as Cl

<u>Date</u>	<u>No. 12 Sump</u>	<u>12 to 13 Sump-to-Sump</u>	<u>No. 13 Sump</u>	<u>No. 13 Sump</u>	<u>13 to 14 Sump-to-Sump</u>	<u>No. 14 Sump</u>
7-17-68				3.513	3.539	
7-17-68				3.771	3.806	
7-19-68	3.522	3.487	3.692			
7-24-68	3.206	3.184	3.418			
7-25-68				3.169	3.174	3.450
7-29-68				3.869	3.880	4.153
7-30-68	3.921	3.858	4.085			
7-31-68	3.762	3.694	3.942			
8-1-68				3.910	3.934	4.218
8-7-68	3.820	3.765	3.992			
8-8-68	3.706	3.669	3.867			
8-9-68	3.685	3.645	3.867			

TABLE G-3-a (CONTD)

SUMP-TO-SUMP CHLORINITY RESULTS
wt. % as Cl

<u>Date</u>	<u>No. 12 Sump</u>	<u>12 to 13 Sump-to-Sump</u>	<u>No. 13 Sump</u>	<u>No. 13 Sump</u>	<u>13 to 14 Sump-to-Sump</u>	<u>No. 14 Sump</u>
8-13-68				4.250	4.272	4.630
8-14-68*						
8-15-68	3.826	3.772	4.016			
8-27-68	3.471	3.482	3.584	3.584	3.623	3.876

*No. 11 Sump - 3.331%

No. 12 Sump - 3.606%

Sump-to-Sump -3.326%

b. Condensate Transfer

During Development Run 14 it became apparent that some of the modular evaporators were operating with condensate levels in their respective steam chests. This resulted in poor heat transfer performance and less than 100% plant rate operation. The major cause of this problem was attributed to undersized condensate transfer piping. During the Development Run 14/15 shutdown, several equipment changes were made in the condensate transfer system. The list below summerizes the modifications and reasons why they were made.

<u>Modification</u>	<u>Reason</u>
1. Internal modular condensate line size increase from 8" to 10".	Allow free condensate transter so flash tanks would operate at lower levels.
2. Flash tank 13 condensate inlet nozzle size increase from 8" to 10".	Decrease pressure drop to allow lower operating level in flash tank 12.
3. Flash tank 15 to 16 internal condensate transfer pipe.	Reduce heat loss by internal piping. Allow lower operating level in flash tank 15 by providing shorter, larger piping system.
4. Condensate inlet baffles in flash tanks 13 thru 17.	Provide water seal to avoid steam blow-by. Insure complete equilibrium condensate flashdown.

<u>Modification</u>	<u>Reason</u>
5. Large, rectangular, slotted effect condensate drains in Nos. 13 thru 17.	Minimize possibility of plugging 2 inch pipe chest drains due to large rust flakes.

During Run 15, the plant operated with the above mentioned changes. The modular evaporator performance was significantly better; the condensate levels were lower, flash tank 12 included; total plant production was increased. It was still, however, necessary to divert water from #15 flash tank to the suction of product pump P-41 in order to get maximum plant output.

c. Brine Distribution

The 11th effect at Freeport is equipped with a device that can be used to measure the liquid rate out of a number of evaporator tubes. Figure G-3-a is a schematic drawing of the device. The brine falls out of the tube into a collection cup. The collection cup has been centered under the specific tube by an external push rod arrangement. The brine flows through a flexible hose through the evaporator shell, a valve manifold, and through the rotometer to the suction of P-21. The tubes through which the brine flow can be measured are shown numbered on Figure G-3-b. The blacked in circles on this Figure represent tubes that are blanked off. It can be seen that the flow through a large percentage of the tubes in Effect 11 can be studied.

This device is used to study such things as: the effectiveness of different distribution weirs; the effectiveness of the brine distribution plate; distribution under flashing and non-flashing conditions; the effect of water box feed rate on brine distribution; and the effectiveness of various tube feed devices.

Normal procedure has been to position the collection device and then open the valve on the discharge of the rotometer wide and allow the system to come to equilibrium. When this is done the rotometer bob cycles vigorously up and down almost half the total flow span of the instrument. It was found that if the rotometer discharge valve was throttled, the rotometer reading would gradually drop until it became relatively stable and constant with time. Presumably a liquid level is established in the flexible hose that is just high enough to offset the pressure drop across the rotometer discharge valve at a constant flow.

Using this technique the results from these tests were quite reproducible. That is, the same point would give the same reading at different times during the individual test period, and the same general flow pattern seemed to repeat itself from test to test.

Tests were conducted using two types of brine distributors; polypropylene swirl promoting weirs (SVL), and porcelain spray nozzles. Also, each distributor was tested with flashing feed, and recycle feed.

Several sets of data were taken at each of the four combinations of feed mode - distributor type. One typical brine distribution pattern is shown for each distributor tested in Figures G-3-c & d. The numbers in the tubes tested represent the volume fraction of the average tube feed rate. The distribution patterns are summarized on Table G-3-b. The spray nozzle distribution is much more even. It is not known whether this is due to the SVL weirs or the action of the distribution plate.

Looking at the data in Table G-3-c it can be seen that the average of the measured tube feed rates is about 65% of the average tube feed that was calculated by dividing the unit top water box feed by the number of tubes. It could be that some of the tubes--not in the measurement system--are heavily loaded; there might be some brine coming out of the tubes that is not collected. It should be pointed out, that there is excellent repeatability between data taken under the same conditions but at different times with the flow measurement device. There is a considerable variation in tube feed rate; those tubes receiving less than 40% of the average tube feed could, on occasion, dry up and encounter scale problems.

When the brine falls out of the end of the tube, it bells out because of edge effects. This coupled with the cross-flowing steam probably diverts some of the brine out of the collection device.

From the tests on these two distributors the following statements can be made:

Polypropylene SVL Weirs

Advantages:

1. Unit cost very low
2. Will not plug easily
3. Small shipping breakage

Disadvantages:

1. Must be held in place
2. Material deforms and becomes brittle
3. Does not put all brine on tube wall

Porcelain Spray Nozzles

Advantages:

1. Unit costs low
2. Material unaffected by brine
3. Performs best distribution

Disadvantages:

1. Must be held in place
2. Orifice can plug with rust flake
3. Needs special packaging to avoid high shipping breakage
4. Requires a pressure drop for spray action

TABLE G-3-b

#11 EFFECT DISTRIBUTION PATTERN

SUMMARY

<u>Type Nozzles</u>	<u>SVL</u>	<u>Porcelain Spray</u>
Date	1/10/68	1/31/69
Calculated Average Tube Feed Rate, #/Hr./Tube	850	920
Mode of Brine Transfer	Feed Forward	Sump-to-Sump
Tube Rate, Volume Fraction of Average Tube Rate	<u>% of Tubes in Each Class</u>	
Less Than 0.2	5	2
0.3-0.4	17	0
0.5-0.6	30	4
0.7-0.8	17	24
0.9-1.0	19	67
1.1-1.2	14	3
More Than 1.3	8	0

KSC :cmb
9/22/69

TABLE G-3-c

BRINE DISTRIBUTION IN EFFECT NO. 11

DEVELOPMENT RUN 16

Date	Water Box Feed Rate, lbs/hr	Water Box Feed Mode	Tube Feed Measured lb/hr	Tube Feed Calculated, lb/hr
1-3-69	324,000	Recycle	(Average) 640	950
1-10-69	328,000	Recycle	620	980
2-12-69	349,000	Recycle	700	1,040
2-14-69	264,000	Forward (Flashing)	480	770
3-7-69	276,000	Forward (Flashing)	450	810

FIGURE G-3-a

SKETCH OF SYSTEM USED TO COLLECT AND
MEASURE BRINE FLOW OUT OF TUBES IN EFFECT 11

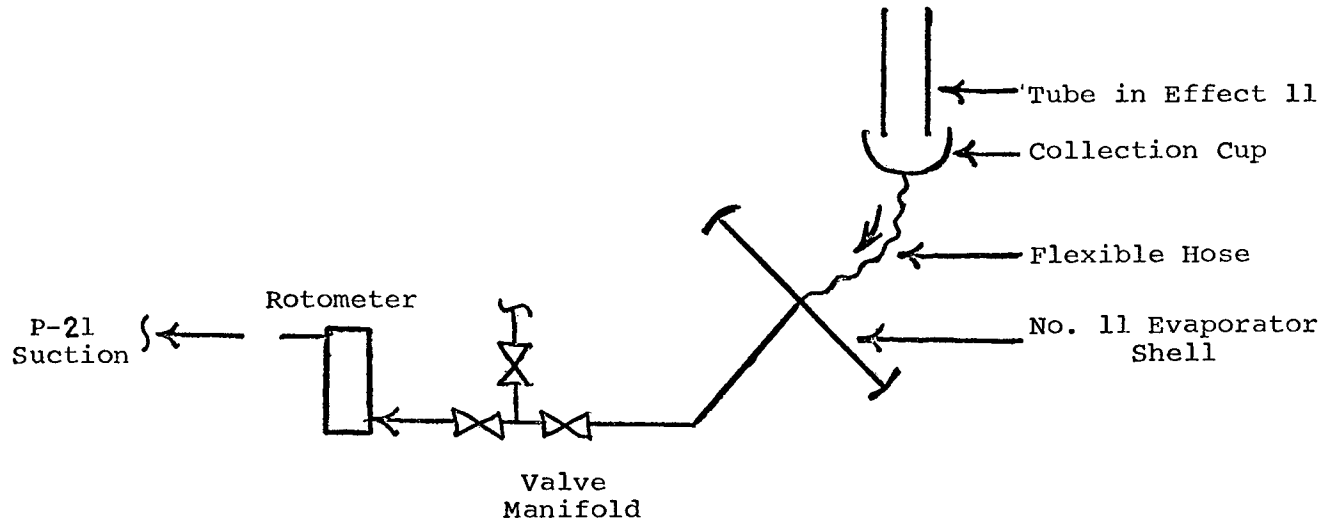


FIGURE G-3-b

TUBES COVERED BY TUBE FLOW
MEASURING DEVICE

EVAPORATOR 11

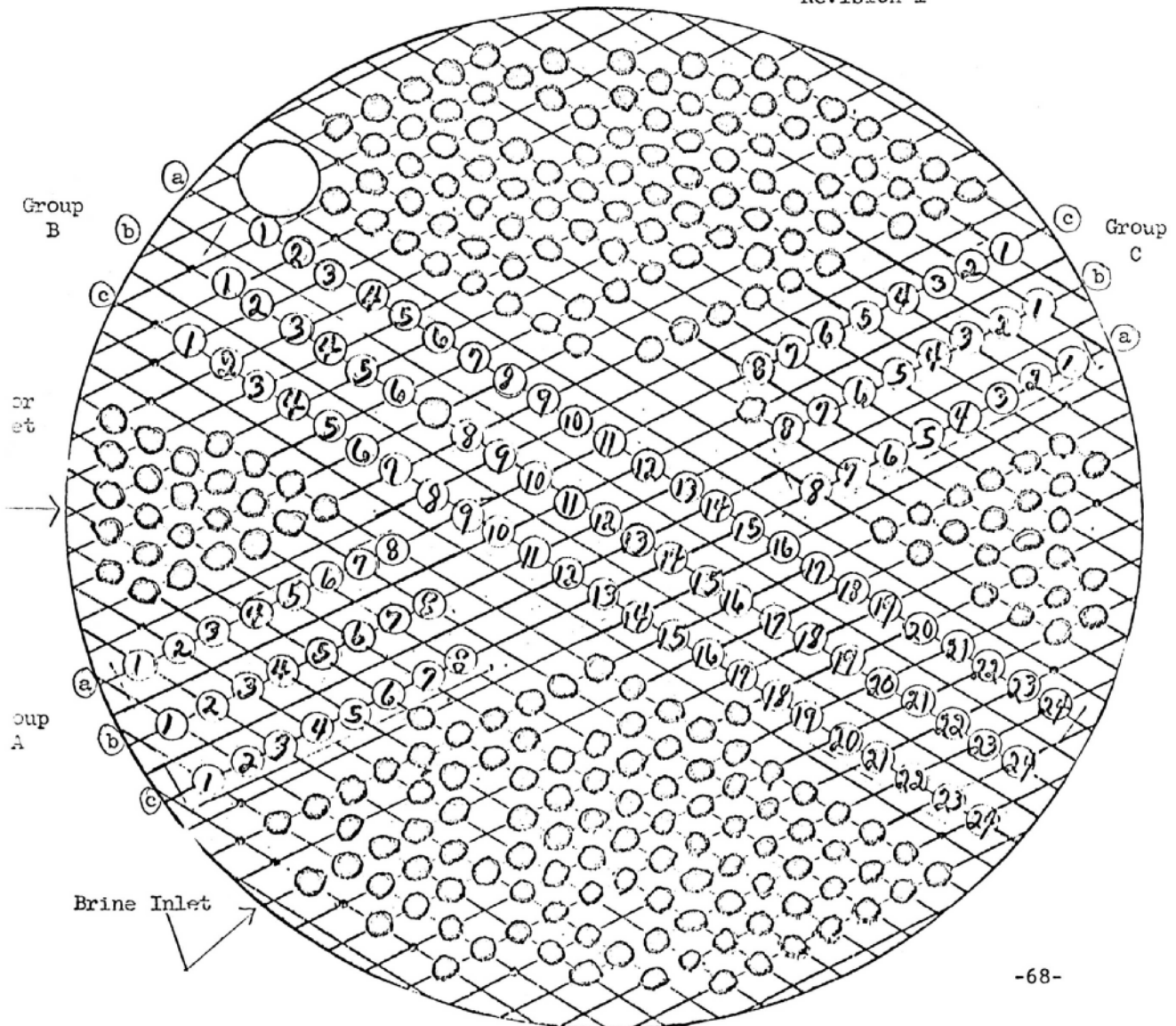
May 1, 1967

Evaporator Heating Element
Tube Configuration

EFFECT NO. 11

Total Open Tubes	316
Total Plugs	244

Revision 1



May 1, 1967

Evaporator Heating Element
Tube Configuration

EFFECT NO. 11

Total Open Tubes 316
Total Plugs 244

Revision 1

FIGURE G-3-c

BRINE DISTRIBUTION IN NO. 11 EFFECT
2" PORCELAIN NOZZLES

Date: 1/31/69

BASIS:

1. Average Tube Feed Rate
1.87 GPM/tube
2. Sump to Sump Mode
3. 2" Porcelain Nozzles
4. Data are in terms of
fraction of average
tube feed rate.

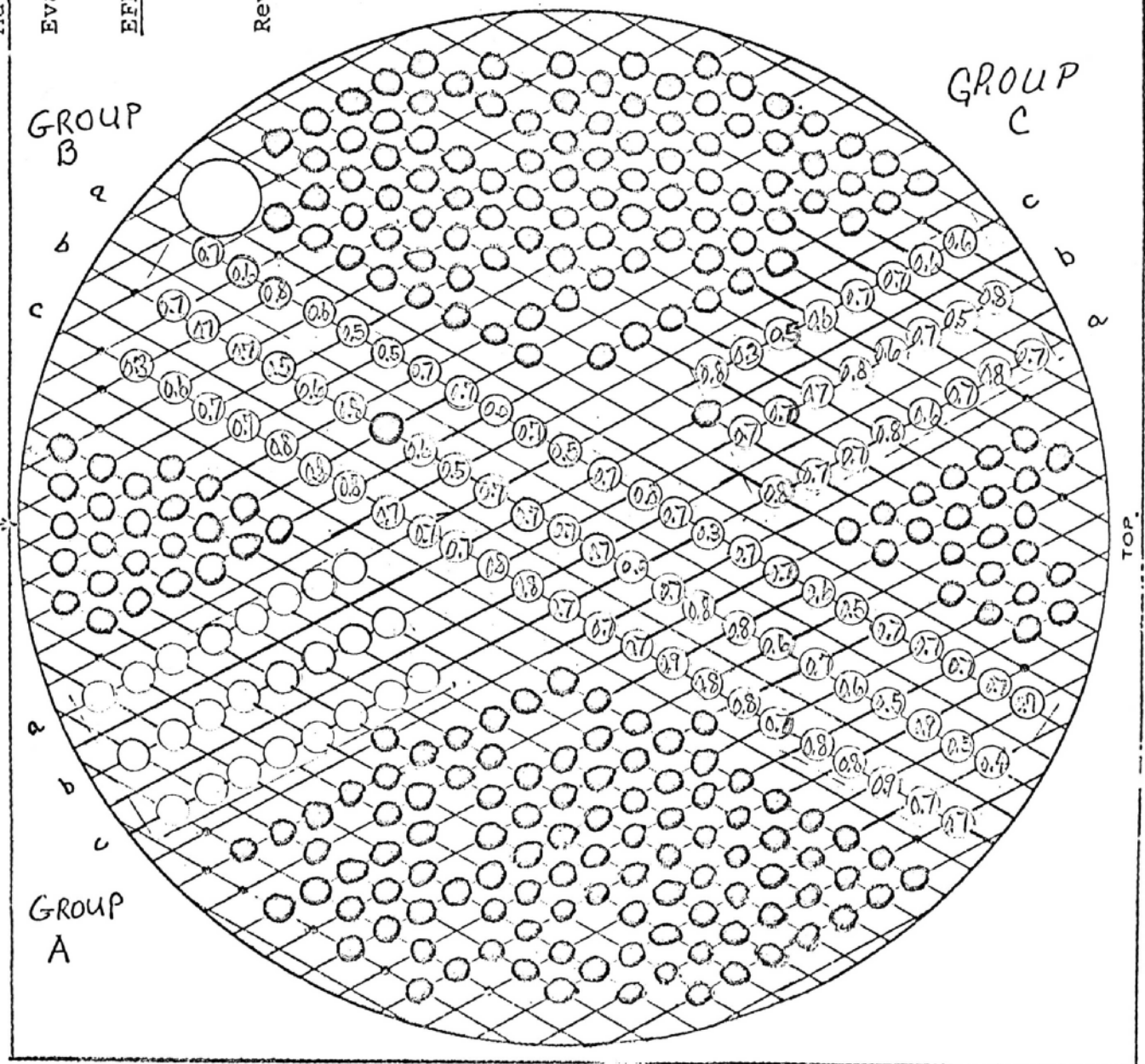


FIGURE G-3-d

BRINE DISTRIBUTION IN NO. 11 EFFECT

SVL WEIRS

CONDITIONS:

- (1) Feed forward (Flashing) with maximum recycle
- (2) Average Calculated tube feed 850 #/hr/tube
- (3) Figures in circles equal observed tube flow rate divided by average tube Feed Rate
- (4) Polypropylene swirl promoting weirs in service

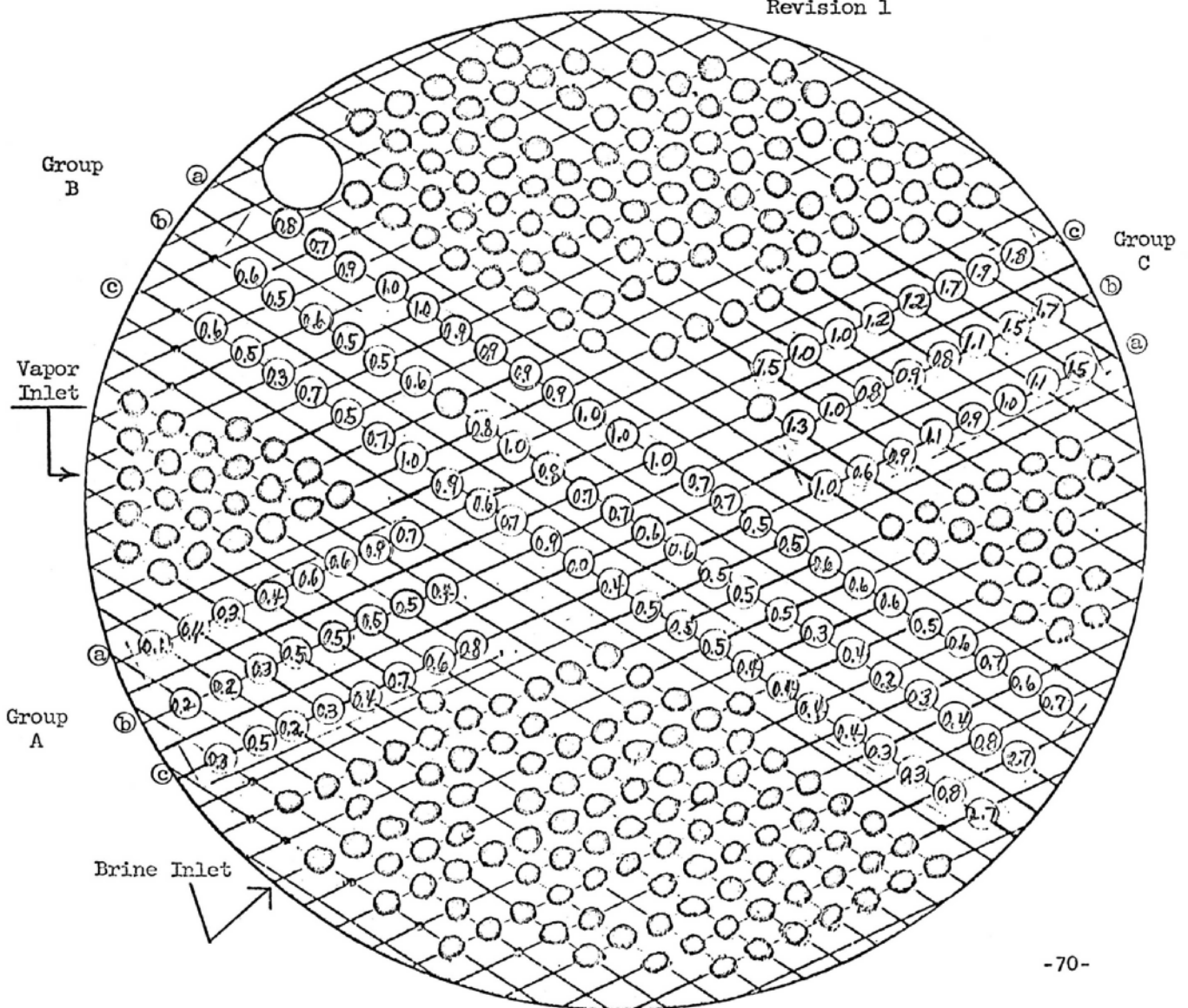
May 1, 1967

Evaporator Heating Element
Tube Configuration

EFFECT NO. 11

Total Open Tubes	316
Total Plugs	244

Revision 1



G. RESULTS AND COMPARISONS (CONTD)

4. DEAERATOR-DECARBONATOR PERFORMANCE

a. Acid Mix Chamber

At the shutdown between Development Run Nos. 14 and 15, an acid mix chamber, constructed entirely of glass-fiber reinforced polyester and filled with the same packing as the deaerator, was inserted between the acid injection point and the deaerator-decarbonator. The presence of this mixing chamber did not improve the erratic trace of the continuous pH meter monitoring the pH of acidified brine into the deaerator-decarbonator. Neither was any improvement in acid mixing noted. The packing was removed from this device to reduce system pressure drop to compensate for the reduced output from brine feed pump P-3. The chamber remained in the line through Development Run 16; the polyester material was unaffected by the acidified sea water.

Considerable time and effort has been expended during the last few development runs to determine an optimum acid mixing and brine sampling strategy. The inlet to the deaerator, just after acid addition, is a much favored pH sensing point because of the rapid response. However, because of the mixing and the change in pH that both occur in the deaerator-decarbonator, many prefer to sense the desalination plant acid-treated feed pH at the deaerator discharge.

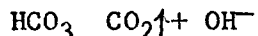
If mixing time is considered critical, then the acid injection point should be positioned in the brine line further ahead of the deaerator-decarbonator. If uniform acid addition is necessary, then the acid injection pumps should be equipped with pulsation dampeners. Certainly a pulsed addition of the acid with the existing single acting positive displacement pump does not yield a constant acid concentration in the brine stream. However, experiments with the deaerator have indicated that the effluent brine pH is quite stable in spite of considerable fluctuations in the pH of the incoming brine.

No further work on optimizing the acid addition point for sampling or sensing the brine pH before the deaerator-decarbonator is recommended. However, if work in this area is carried out, it should be remembered that the first problem is to get a pH meter that can handle acidified inlet brine. One of the problems that is encountered is that the CO₂, which tends to flash off immediately after acid addition, vapor binds the pH meter electrode chamber and causes erroneous and fluctuating readings. No meaningful inlet pH readings can be made until this problem is solved.

b. Deaerator Effluent pH Control

In Development Run 15, the control sample point for acid addition was switched to the deaerator-decarbonator outlet brine before the caustic injection point. The acid addition was regulated to control the pH between 5.0 to 5.5.

Analysis of deaerator-decarbonator data taken during Development Run 14 indicated that when less than the so called "stoichiometric" amount of acid was added to the brine, the pH of the deaerator outlet brine went up. It was decided to try some experiments to see if the acid addition requirements could be minimized and the caustic neutralization step could be eliminated. It was reasoned that if the bicarbonate alkalinity of the deaerator effluent were kept around 10, as ppm Ca CO₃ and the pH were kept between 5.6 and 6.0, then as the brine was heated up the bicarbonate ions would break down forming free carbon dioxide and hydroxyl ions as per the equation:



If this reaction occurred early in the preheat train, the pH of the brine would be raised to the neutral range eliminating the necessity of caustic neutralization. This would save equipment, chemicals, and operator attention. If the degree of decarbonation were carefully controlled, the pH would never get high enough to cause magnesium hydroxide or calcium carbonate scale problems. The only real cause of concern was that carbon dioxide would be given off in the first effect and cause corrosion and venting problems.

Toward the end of Fiscal Year 1968, the control range was raised to 5.5 to 5.8 pH units. During this time, the deaerator effluent was checked for total alkalinity by the plant operators. An upper allowable limit for proper decarbonation was arbitrarily set at 25 ppm as Ca CO₃. At the end of the run, the preheaters were examined for any alkaline scale formation and there was none present. Also, no front end venting or corrosion problems occurred to indicate high first and second effect decarbonating rates.

During Run 16, this carefully monitored increase of pH operating level was continued. The deaerated-decarbonated brine pH control ranges specified were:

5.6 to 6.0 pH
5.8 to 6.2 pH
6.0 to 6.4 pH

At the middle pH range, caustic addition for excess acid neutralization was stopped. As in Run 15, since the total alkalinity to the first effect remained below our 25 ppm Ca CO₃ limit and no venting problems arose, the run was completed at the last set of conditions. No alkaline scale was observed in the preheat exchangers when they were inspected at the end of the run.

c. Carbon Dioxide and Total Alkalinity of Decarbonated Brine

Carbon dioxide analyses were performed on a routine basis on deaerator effluent and blended sea water samples. The results are shown on Table G-4-a. The carbon dioxide content of the deaerator effluent brine ranged from 0.3 ppm up to as high as 9.9 ppm.

It can be seen that there is considerable variability in the carbon dioxide content of the blended sea water. There are several reasons for this:

- 1) Blending - Concentrated brine from the last effect is blended with raw sea water in order to maintain the desalination plant feed at a concentration factor of about 1.0. Thus, concentrated brine, which contains essentially no carbon dioxide, is blended -- in significant but varying quantities -- with raw sea water to produce the blended desalination plant feed. Changes in blending rate will obviously cause changes in the blended brine carbon dioxide content.
- 2) Silt - Changes in the silt content of the incoming brine will affect the results of the carbon dioxide determination.
- 3) Raw Sea Water - The carbon dioxide content of raw sea water is not necessarily constant. It has been noted that the carbon dioxide content of raw sea water changes significantly when the sun goes down. It has been assumed that removing the sunlight causes a change in photosynthesis activity which in turn changes the carbon dioxide concentration of the raw sea water. Thus, it is probable that anything which changes the amount of sunlight (cloud cover, turbidity, etc.) will eventually result in a change in carbon dioxide content of the raw sea water.

Extreme care must be used when basing the design of a decarbonator on Freeport data. The raw sea water frequently has a concentration factor of less than 0.8; in the plant, it is mixed with a brine which has essentially a zero carbon dioxide content. Thus, our acid usage (and inlet carbon dioxide level) is considerably less than that which would be required in a desalination plant receiving a normal sea water feed.

TABLE G-4-a
CARBON DIOXIDE ANALYTICAL RESULTS

<u>Date</u>	<u>D/A Unadjusted Out, ppm CO₂</u>	<u>Blended Sea Water (P-3 Disch.) ppm CO₂</u>
<u>Development Run 15</u>		
5- 1-68	3.4	----
5- 1-68	5.5	----
5- 6-68	1.7	46.3
5- 6-68	4.1	55.4
5- 9-68	8.2	39.4
5-13-68	0.3	61.0
5-13-68	1.7	54.3
6- 3-68	4.4	70.0
6- 5-68	1.6	----
6-10-68	2.9	66.2
6-19-68	6.2	64.8
6-27-68	4.2	62.3
7- 3-68	1.4	54.7
7-15-68	2.3	72.5
7-24-68	2.8	72.2
7-31-68	3.1	65.8
8- 8-68	2.0	64.4
8-14-68	5.6	77.9
9- 5-68	9.9	73.3
<u>Development Run 16</u>		
1- 9-69	2.7	----
1- 9-69	3.0	----
1-16-69	3.4	----
1-16-69	4.7	----
2- 4-69	5.5	----

d. Dissolved Oxygen

1) Development Run 15:

The dissolved oxygen analyses taken during Development Run 15 are summarized on the attached Table G-4-b. Samples which were taken for testing purposes, such as analytical reagent blanks, etc., are not reported. The dissolved oxygen determinations were all made by the modified Winkler technique.

The dissolved oxygen content of the deaerator unadjusted outlet brine was above 50 ppb for the first part of Run 15. After several leaks in the deaerator were repaired, the dissolved oxygen of the deaerated brine dropped to between 30 and 50 ppb for the remainder of the run. The dissolved oxygen content of the deaerator inlet brine showed some variability. It should be noted that the dissolved oxygen content of the blended brine will run lower than the dissolved oxygen content of the raw sea water. This is because the brine that is used for blending has essentially no oxygen. Note also that because of the solubility of oxygen, the deaerator has less oxygen to remove per pound of sea water in the summer when sea water temperatures are high.

The dissolved oxygen of the deaerator effluent ran higher in Development Run 15 than it did in Development Run 14. There had been a change in Development Chemists between these two runs. As a check on the procedure on several different occasions, samples of deaerated sea water were obtained from the Materials Test Facility. These samples were known to contain close to zero dissolved oxygen; in both cases, no dissolved oxygen could be found. An outside person, quite experienced with the Winkler analysis, was asked to run the analysis on a sample of our deaerator effluent. The results obtained were in close agreement to those obtained by the Development Chemist.

TABLE G-4-b

DISSOLVED OXYGEN ANALYTICAL RESULTS (WINKLER ANALYSIS)

DEVELOPMENT RUN 15

	Deaerator Outlet Brine ppb	Blended Brine to Deaerator ppm
<u>Development Run 15</u>		
5/13	140	5.3
5/22	80	---
5/23	70	---
5/24	80	---
5/28	60	---
5/29	80	---
6/3	100	4.1
6/5	100	---
6/7	90	---
6/10	80	3.9
6/19	70	4.3
6/20	70	---
6/21 (1)	40	---
6/24	20	4.1
6/27	30	---
7/1	40	---
7/3	30	3.4
7/5	50	---
7/9	40	---
7/16	30	---
7/22	40	---
7/24	50	3.1
7/26	30	---
7/29	40	---
7/31	40	2.5
8/2	40	---
8/8	50	3.7
8/12	40	---
8/14	30	3.2
8/26	20	---
9/3	40	---
9/5	30	3.4
9/9	40	---
9/19	60	---

(1) A leak in the deaerator had been located and fixed.

d. Dissolved Oxygen (Contd)

2) Development Run 16:

The dissolved oxygen results for the development run are discussed in the next section, (G-4-e) Sulfite Addition. However, early in the run, a serious sampling problem was uncovered and solved.

During a series of sulfite residual determinations, it was noticed that the dissolved oxygen content of the deaerator effluent after the sulfite injection point (See Figure G-4-a), was zero at a time when no sulfite injection had taken place for over 12 hours. At this time, samples were being collected via the taps in the operations laboratory. Since the dissolved oxygen before the sulfite injection point (D/A Unadjusted Effluent) was found to be 80 ppb and since no sulfite was being injected, it did not make sense that the D/A Adjusted tap should show zero dissolved oxygen.

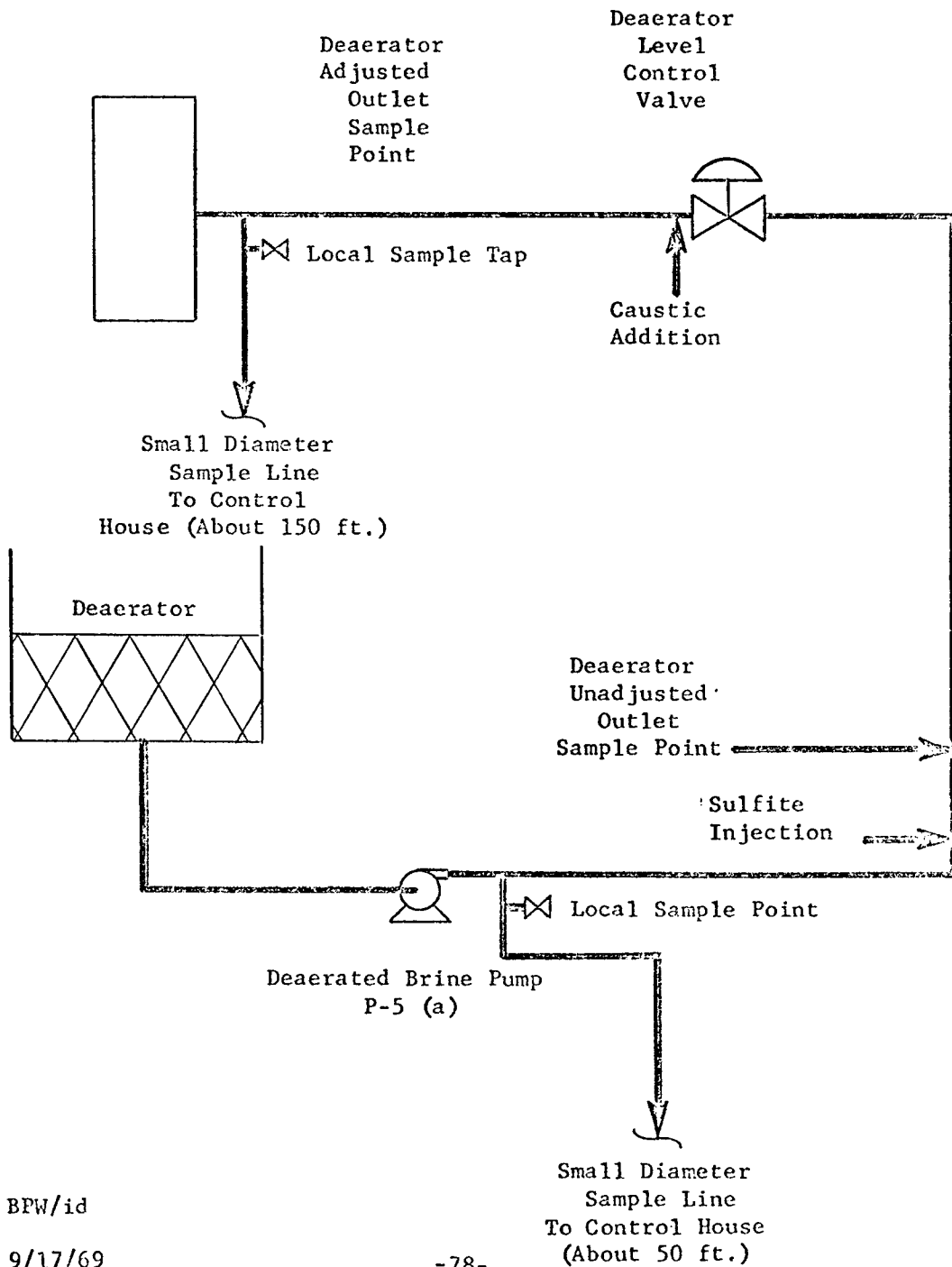
It was at first thought that possibly the sample line had been connected to a wrong point during the shutdown. Both of the sample lines were traced and found to be properly connected. It was then noticed that the packing gland on P-5a had been connected to the deaerator unadjusted sample line. It was thought that possibly oxygen was entering the line via the packing gland. The pump packing gland was valved off, and oxygen analyses were again run. Dissolved oxygen was still found at the deaerator unadjusted tap but not at the adjusted tap.

In the process of running many oxygen analyses at various sampling points it was established that the dissolved oxygen at the origin of the deaerator adjusted line was much higher than the dissolved oxygen at the tap in the laboratory; the dissolved oxygen at the origin being 120 ppb while at the tap it was zero. A similar situation was found for the deaerator unadjusted line. At the P-5a origin of the line, a dissolved oxygen content of 100 ppb was found, whereas the tap in the laboratory showed only 70 ppb.

The present proposed explanation for this difference in dissolved oxygen between the origin and termination of these two sample lines is that the oxygen is reacting with the sample line itself. Since the sample "sees" steel and copper tubing during its travel to the laboratory taps, and further since the surface

FIGURE G-4-a

Deaerator Sampling Points



BPW/id

9/17/69

area of the small diameter sample line is large relative to the velocity through the line, it is quite probable that this is happening. This also explains the greater dissolved oxygen content difference for the adjusted line versus the unadjusted line; the line for the adjusted effluent sample is approximately three times longer than the unadjusted line. This gives about three times more metallic surface and three times longer for the oxygen in the effluent to react within the line.

Since establishing the difference in dissolved oxygen content along the sample line, all dissolved oxygen samples have been taken at the point of origin in the plant. The sample for the unadjusted effluent is now taken at P-5a, and the adjusted effluent is taken on top of the module at a point well downstream of the sulfite injection point.

e. Sulfite Addition

1) Background

The economics of chemical scavenging of the residual dissolved oxygen was studied. Assuming the addition of 10 parts of sodium sulfite per part of oxygen at a cost of 20¢ per pound for the sodium sulfite, the cost of sulfite addition in LTV plants with high extraction ratios would not be much more than 0.2¢ per 1000 gallons of product water, based on deaerator effluent dissolved oxygen scavenging only. Scavenging of subsequent inleakage might raise the cost significantly.

Oxygen removal from distillation plant feed represents the unique case where the use of a combination two-stage process is more economical than a single-stage process. Laboratory studies indicate that oxygen reduction to the low part per billion range tends to reduce the corrositivity of brine in distillation plants. Reduction to this level by mechanical means or chemical means alone is economically prohibitive. However, the reduction to below 1 ppb can be achieved by a combination of mechanical deaeration and chemical scavenging at a very nominal cost.

In the past, attempts have been made to design a mechanical deaerator that would do the job in one step. However, the deaerator is only one part of the distillation process, and it is sometimes not practical to restrict the flexibility of the total process in order to maintain the invariant operating conditions at the deaerator which are necessary for optimum mechanical

deaeration. In an operating plant, the deaerator has to cope with such things as changes in the brine oxygen and carbon dioxide content, brine rate, stripping steam rate, brine flashdown, brine temperature, brine salinity, etc. It would probably be prohibitively expensive to design a mechanical deaerator that would handle all combinations of these changing conditions and also reduce the oxygen content of the brine significantly below the present 50 ppb range. However, a combined chemical scavenging and mechanical deaeration system can maintain the low oxygen level in the face of all combinations of the above mentioned variables. Additionally, if the scavenging system is installed, the plant can be immediately protected against malfunctions or leaks in the mechanical deaeration system. The effectiveness of scavenging agents in removing dissolved oxygen and reducing brine corrosiveness was tested in Run 16.

2) Analytical Problems

During Run 16, sodium sulfite was injected into deaerator effluent in an attempt to study the effects of sulfite as an oxygen scavenger in a seawater environment. A sulfite residual of 0.1 to 0.2 ppm was chosen. A suitable analytical method was required which would be usable to detect these low sulfite residuals. Since most of the published sulfite methods are for higher levels of sulfite, it was necessary to try to modify these methods to suit our purposes.

The first method tried was from Standard Methods⁽¹⁾ involving the reaction of sulfite with an excess of iodine (standard) and titration of the excess iodine with a standard sodium thiosulfate solution. Both the standard iodine and thiosulfate solutions were made much more dilute than that called for in an attempt to detect a low sulfite residual. This method failed in that the difference in volume of thiosulfate required to titrate the iodine before and after reaction with the sulfite was too small (zero in some cases) to be measured. Making the standard solutions more dilute would only "broaden" the titration end point. This method was thought to be insufficient for our purposes.

(1) Standard Methods For Examination Of Water And Waste Water (1960); American Public Health Association, et. al.; page 244.

The next method tried was ASTM method D 1339-54T involving reaction of the sulfite with hydrochloric acid and potassium iodide followed by titration to a starch endpoint with standard potassium iodate solution. Again the titrant was made more dilute than called for. The endpoint was very diffuse, making this procedure useless for low sulfite levels.

The next method tried was ASTM D 1339, Reference Method "A", specified as being usable down to 0.1 ppm. This is a very exacting analytical procedure involving a sample bomb such as the one presently used for the dissolved oxygen procedure. The sulfite is reacted with a known amount of iodine chloride, the excess iodine chloride being titrated to a starch endpoint with standard thiosulfate solution. The endpoint is fairly sharp, but a rather small thiosulfate volume difference is involved. It was found that this method was very difficult to run and that results were not reproducible. No further work on this method was pursued.

The first method (from Standard Methods for Water and Waste Water) was again tried and finally evolved into a usable method, good to ± 0.2 ppm as Na_2SO_3 . Note that since it takes at least an hour to take the sample and run the analysis, the observed sulfite residual may be low due to reaction during sampling and analytical time.

Toward the end of Development Run 16, it was decided to try to control sulfite additions to a threshold oxygen level rather than trying to measure an extremely small residual sulfite level. The sulfite addition rate was gradually backed off until the dissolved oxygen, after sulfite addition, showed a significant increase. Because of the time involved in running a Winkler analysis, this technique of scavenging agent addition control was tedious and unreliable. With a continuous dissolved oxygen meter, this method of control would be quite appropriate and could probably even be engineered to operate in an automatic control function.

3) Injection Point Plugging

During the first part of the run, the sodium sulfite solutions were made up at the strength recommended by the manufacturer. However, plugging of the injection point occurred necessitating cleaning at least once per shift. It was assumed that the plugging was due to the precipitation of calcium sulfite, which

is extremely insoluble, at the injection point. Experiments with different injection point locations and configurations did not solve the problem. The plugging problem was reduced but not totally solved by going to a much reduced sodium sulfite concentration.

At the end of Development Run 15, some small accumulation of scale was noted in the tubes of the first effect. There was some speculation that this scale might have been calcium sulfite due to the sulfite addition late in Development Run 15. The scale in the first effect had not been analyzed. However, sodium sulfite was added to the deaerator effluent brine during most of Development Run 16 and during this time the amount of scale in the first effect seemed to diminish.

4) Results

During Development Run 16, dissolved oxygen analyses were performed on a routine and/or as needed basis on deaerator effluent both before and after sodium sulfite injection.

Table No. G-4-c lists these analytical results for Run 16. Note that results involving very low dissolved oxygen are not reported as zero but as 5 ppb. This is due to inherent inaccuracies in the Winkler procedure at very low oxygen levels.

Early during Run 16, tests were conducted in an effort to find out why the dissolved oxygen in the deaerator effluent was higher than normal. On January 6, tests were conducted to determine if oxygen was being introduced into the deaerator along with the stripping steam. This was accomplished by running dissolved oxygen on a sample of effluent "before sulfite" (200 ppb D.O.) and then repeating the analysis after the stripping steam had been cut off (140 ppb D.O.). A drop in dissolved oxygen content of 60 ppb was found. However, after the stripping steam was turned back on a further drop in dissolved oxygen was found (110 ppb found in effluent). It is not felt that the test was conclusive. On January 14, tests were conducted to determine if oxygen was leaking in at the seals of P-5 (a). Samples of deaerator effluent were collected on the suction and discharge sides of the pump and analyzed for dissolved oxygen. Both samples showed 80 ppb D.O. It was thus concluded that the seals of P-5 (a) were

TABLE G-4-c

DISSOLVED OXYGEN ANALYSIS RESULTSDEVELOPMENT RUN 16

Date	D/A Effluent Before Sulfite, ppb	D/A Effluent After Sulfite, ppb
12-19-68	100	40
12-19-68	70	30
12-23-68	60	<5
12-26-68	110	<5
12-27-68	80	<5
12-30-68	120	---
12-30-68	70	---
12-30-68	100	---
*12-30-68	110	---
12-30-68	---	100 (1)
12-31-68	140	<5
12-31-68	---	<5
1-2-69	120	<5
1-3-69	---	170 (1)
1-3-69	180	<5
1-6-69	200	<5
1-6-69	150 (2)	---
1-6-69	110 (3)	---
1-9-69	110	<5
1-10-69	---	60 (4)
1-10-69	---	<5
1-14-69	80	<5
1-15-69	80	---
1-16-69	90	<5
1-22-69	---	<5
2-4-69	100	<5
2-5-69	140	150 (1)
2-5-69	---	60 (4)
2-6-69	---	50
2-11-69	---	120 (1)
2-12-69	130	140 (1)
2-12-69	---	100 (1)

Date	D/A Effluent Before Sulfite, ppb	D/A Effluent After Sulfite, ppb
2-13-69	---	100 (1)
2-13-69	---	< 5
2-25-69	---	< 5
2-26-69	---	100 (5)
3-4-69	---	< 5
3-18-69	---	100 (1)
3-18-69	---	40 (4)
3-19-69	---	60 (4)
3-19-69	---	50 (4)
3-27-69	---	20 (4)
4-7-69	---	< 5
4-9-69	---	40 (4)
4-14-69	40	< 5
4-15-69	---	< 5
4-17-69	---	< 5
4-28-69	---	50 (6)
5-1-69	---	< 5
5-6-69	< 5 (7)	---
5-7-69	35	---

- NOTES:
- (1) Sulfite pump not operating properly, no sulfite injection.
 - (2) D/A stripping steam off.
 - (3) D/A stripping steam back on.
 - (4) Sulfite injection rate too low.
 - (5) Sulfite injection off for other special tests.
 - (6) Sulfite solution strength reduced by dilution.
 - (7) Plant blended up - sulfite residual carryover.

not leaking oxygen into the plant. No further work was carried out as the dissolved oxygen dropped to and remained at a lower level immediately after these tests were conducted.

The results indicate that the addition of sodium sulfite apparently reduces the dissolved oxygen content of the deaerated brine below the threshold of measurement. The rate at which this occurs in the actual preheat train environment is, unfortunately, not known. Since it takes about fifteen minutes to collect and "fix" a sample for the Winkler determination, it is possible that residual sodium sulfite reacts with oxygen during analysis giving a low apparent dissolved oxygen. The actual instantaneous dissolved oxygen level in the preheat brine after sulfite addition can only be determined by use of an on-line continuous dissolved oxygen meter. It is recommended that one of these be obtained for future runs.

Although the cost of the sulfite itself is relatively low, the sulfite addition system itself represents an extra addition capital cost and an additional attention point for the people responsible for operating the plant. In many outlying areas, it is very important to keep the operations of the desalination plant as simple and straight forward as possible. The use of sodium sulfite for other scavenging agents to reduce the dissolved oxygen in the deaerated brine to nil cannot be recommended until it can be proved conclusively that this extra treatment step actually reduces the severity of corrosion in the desalination plant.

f. Deaeration Control

The proper operation of the deaerator - decarbonator for best deaeration during Runs 15 and 16 followed those conditions found during Run 14 that resulted in the lowest dissolved oxygen levels in the deaerator brine effluent. They are as follows:

Stripping Steam Rate	500-1,000 lb/hr.
Vent Steam Rate	Enough to allow for flashing feed
Degree of Feed Flash Down	0.5 - 1.0°F

These conditions were monitored by the plant operators and controlled manually. The degree of flash down was determined by reading a matched pair of calibrated thermometers. The stripping steam flow rate was indicated by a manometer and the flow was read on a set of curves, corrected for temperature.

g. Day/Night CO₂ and O₂ Study

During Development Run 14 it was noticed that around sunset each evening the carbon dioxide and dissolved oxygen content of raw sea water varies, the carbon dioxide content showing an increase and the dissolved oxygen a decrease. It is supposed that this is due to a cessation of photosynthesis at this time of day. Since the operation conditions of the deaerator depend on the carbon dioxide and dissolved oxygen content of the sea water it was felt necessary to further establish this observation of CO₂ and O₂ fluctuation.

On two evenings in August samples of raw sea water were taken at half-hour intervals for the purpose of evaluating this variation of carbon dioxide and dissolved oxygen content. The samples for CO₂ analysis were collected in a closed glass "bomb" and samples for O₂ analysis were collected in a BOD bottle.

The data shown in Table G-4-d indicate that a change in CO₂ and O₂ levels do occur at this time.

TABLE G-4-d

DAY/NIGHT CO₂ & O₂ ANALYSES

<u>Date</u>	<u>Time</u>	<u>Temp. °F</u>	<u>PPM CO₂</u>	<u>PPM O₂</u>
8-13-68	1930	84.9	74.3	3.0
	(1) 2000	84.3	78.9	2.6
	2030	84.0	86.1	2.5
	2100	84.1	80.2	2.6
8-14-68	1930	84.6	76.9	3.3
	(2) 2000	84.2	81.5	3.2
	2030	83.8	81.5	3.2
	2100	83.2	77.2	3.1

(1) Sunset at 2004

(2) Sunset at 2003

G. RESULTS AND COMPARISONS (CONTD)

5. MIST ELIMINATORS

During the shutdown between Runs 15 and 16, special piping was installed in the 14th and 15th effects to allow samples of condensate to be withdrawn from the steam chests of these effects for product purity analysis. At the same time, the mist eliminators were removed from the No. 14 effect steam product passage. Two sets of data were taken with this configuration. During a plant shutdown, the mist eliminators were reinstalled and an additional data set was taken. The condensate from the 14th effect was always sampled when sampling the 15th effect's condensate. The test data was always taken during high plant rates. Also, sampling was continuously composited and the conductivity was read hourly for a 24 hour period.

From the data shown on Table G-5-a it can be seen that the presence of a mist eliminator does improve the product water purity. However, considering the high plant rates, the increase in the impurity level due to the absence of the mist eliminator was very small and this effect did not need such a unit if it was only required to produce potable water.

TABLE G-5-a

PRODUCT PURITY WITH AND WITHOUT MIST ELIMINATORS

<u>Date</u>	<u>Sample Source Effect</u>	<u>Mist Eliminators</u>	<u>Plant Rate, %</u>	<u>Product Purity ppm NaCl</u>
2- 5-69	14	Yes	100	1.9
2- 5-69	15	No	100	26
3-12-69	14	Yes	105	2.3
3-13-69	15	No	111	26
3-24-69	14	Yes	100	1.4
3-25-69	15	Yes	101	8.9

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G. RESULTS AND COMPARISONS (CONTD)

6. VENTING STUDIES

A preliminary test of the effectiveness of the module venting system to remove non-condensable gas was performed during Run 16.

Air was allowed to enter the No. 13 steam chest thru a critical orifice which limited the flow to 30-40 lbs/hr, or about 0.2% of the steam to this effect. The deaerator vent rate was increased to the maximum when the air was admitted. Also, the 13th Effect external vent was open.

The data taken consisted of noticing the effect on the complete plant and determining if the test proper could be run. The observations made were as follows:

- a. Plant rate was reduced from 100 to 80%.
- b. Stripping steam to the deaerator increased from 700 to 1,000 lbs/hr.
- c. Final brine discharge rate increased 45% from 155,000 to 225,000 lbs/hr.
- d. Final concentration factor decreased from 2.0 to 1.4.
- e. Final condensing temperature increased 2°F.
- f. Steam rate to the first effect decreased 2% from 26,500 to 26,000 lbs/hr.

After approximately three hours, to allow the plant to approach steady conditions, the external vent of Effect 13 was closed. No noticeable change in plant operations resulted from this. After four and one-half hours all control room data indicated the plant was steady.

As a result of this preliminary test, the following conclusions can be made about the plant while running at 100% rate:

- a. Air inleakage of approximately 35 lbs/hr can be handled by the 5 Effect modular venting system, but with a severe effect on plant operations.
- b. The external bottom chest vent of the No. 13 Effect is not needed at 35 lbs/hr. or less non-condensable vent rates.
- c. Module venting tests can be run at a 30-40 lb/hr. air injection rate as a maximum.

G. RESULTS AND COMPARISONS (CONTD)

7. SODIUM SILICATE USE AS WATER STABILIZER

The work was done to gain information, on a continuing basis, about the corrosiveness of the high purity condensate that the plant produces and to determine the effect of the specific application of N-Sodium Silicate to reduce corrosiveness. To accomplish this, corrosion probes were installed at three locations shown on Figure G-7-a. These locations will allow three determinations to be made:

- a. The corrosiveness of plant product.
- b. The change in corrosiveness due to aeration in the split tank.
- c. The effect of adding product water stabilizing chemicals to the Freeport product.

Sodium silicate injection rate was varied to keep the product pH between 8.5 and 10.5. Treatment data are presented in Table G-7-a.

TABLE G-7-a

PRODUCT WATER TREATMENT DATA

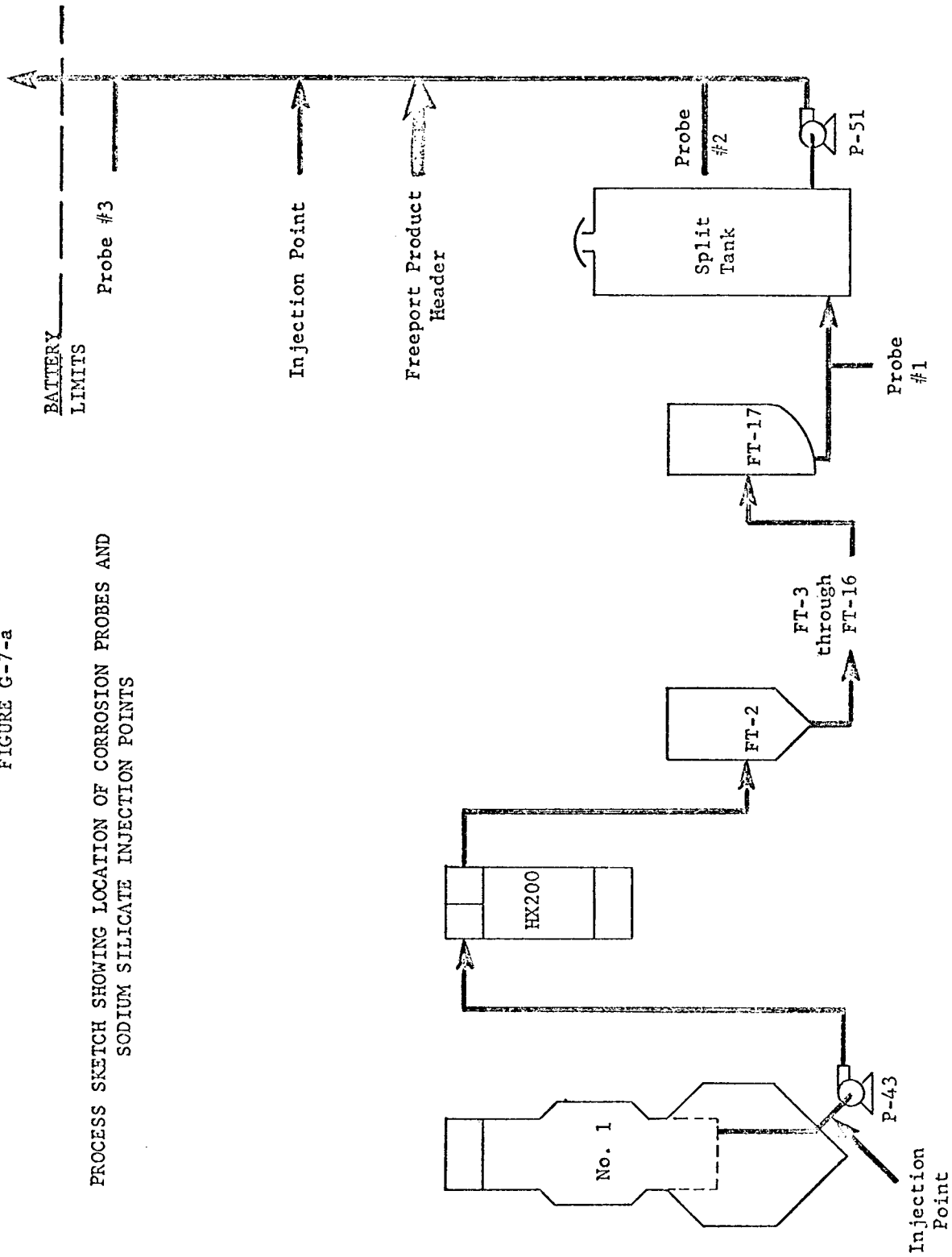
DEVELOPMENT RUN 15

<u>Month</u>	<u>Silicate Pumping Hours</u>	<u>Treated Water Gals.</u>	<u>Silicate Used Lbs.</u>	<u>Dosage ppm Silicate</u>
4-68	3	122,000	16	16
5-68	0	0	0	--
6-68	104	1,490,000	500	40
7-68	225	3,730,000	890	29
8-68	375	6,160,000	1,280	25
9-68	17	392,000	100	31

Corrosion probe readings were taken regularly at the three locations. The corrosion meter "dial readings" were converted into corrosion rate measured in MPY (mils per year). The data are shown in Table G-7-b.

FIGURE G-7-a

PROCESS SKETCH SHOWING LOCATION OF CORROSION PROBES AND
SODIUM SILICATE INJECTION POINTS



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TABLE G-7-b
CORROSION RATES MEASURED IN
PRODUCT WATER SYSTEM

Sodium N-Silicate Injection Point: Freeport Product Line

<u>Probe Location</u>	<u>Dial</u>	<u>Time Days</u>	<u>Corrosion Rate (MPY)</u>
1. Inlet to Split Tank	199	129	5.6 \pm 0.1
2. Outlet of Split Tank	301	129	8.5 \pm 0.1
3. Freeport Product Line	191	129	5.4 \pm 0.1

MAGNA's formula for corrosion rate calculations is:

$$\text{Corrosion Rate, MPY} = \frac{\text{Dial Reading}}{\text{Time (Days)}} \times 0.365 \times 10$$

From the results of the test, the following statements may be made:

- a. The corrosion rates found during this run are in all cases significantly lower (40-60%) than those measured in the last identical test.
- b. The corrosion rates at all locations were relatively uniform for the entire 4-month period following the initial 30 day new probe stabilizing period.
- c. Similar to previous tests, the corrosion rate indicated by the probe in the split tank outlet showed a higher corrosion rate than the probe in the inlet. This confirms aeration taking place in this vented vessel.
- d. The previous tests indicated pitting on the probe element at the split tank outlet. The same result was not shown during this test. At the completion of the run, both sides of the split tank were opened and an inspection revealed only light, uniform, surface rusting with no pitting of the vessel walls.

The original objective was expanded during Development Run 15 to include portable corrater testing because:

- a. A portable corrater and one new mild steel probe were in the plant and not in use.
- b. A multi-point recording corrater installation was planned for Run 16 and a preview of the data it would generate was possible at minimum expense.

- c. A correlation between corrater output and some process variable influencing corrosion rate might be made and make the planned corrater installation a valuable development and operating tool.
- d. An idea of probe life when continuously exposed to actual operating conditions could be made.

The equipment used consisted of a portable corrater, Model 1170, and two non-retractable process probes, Model 7002, Electrode Alloy 1010, Code No. 8001 - mild steel

Probe locations were No. 1 Evaporator sump and the No. 13 to 14 sump-to-sump transfer line, two points that the planned corrater system was to include. Reading frequency was twice a week following corrosometer readings.

Attached are:

- a. Figure G-7-b - A graph of corrosion rate and pitting index and dissolved oxygen vs. time for Development Run 15.
- b. Table G-7-c - A table of calculated corrosion rates for Development Run 15.

Dissolved oxygen (DO) was the process variable with which a correlation was sought. Three reasons for this decision were:

- a. A machine that would continuously monitor the deaerators performance would be a valuable operating aid.
- b. DO data was planned to be taken as part of another objective, and this would not increase the scope of and work planned for the development run.
- c. It was thought that an easily recognized correlation would be found between a major corrosion rate determining process variable and a corrosion rate measuring device.

In discussing this attempted correlation, which appears as Figure G-7-b, the following may be said:

- a. The corrosion rates at both points were very close for much of the run.
- b. The corrosion rates at both points increased and decreased together for much of the run.
- c. The pitting index at both points followed each other very closely during the majority of the run. A change in pitting index number has significance; the number itself does not.

FIGURE G-7-b

CORROSION RATE & PITTING INDEX vs. TIME

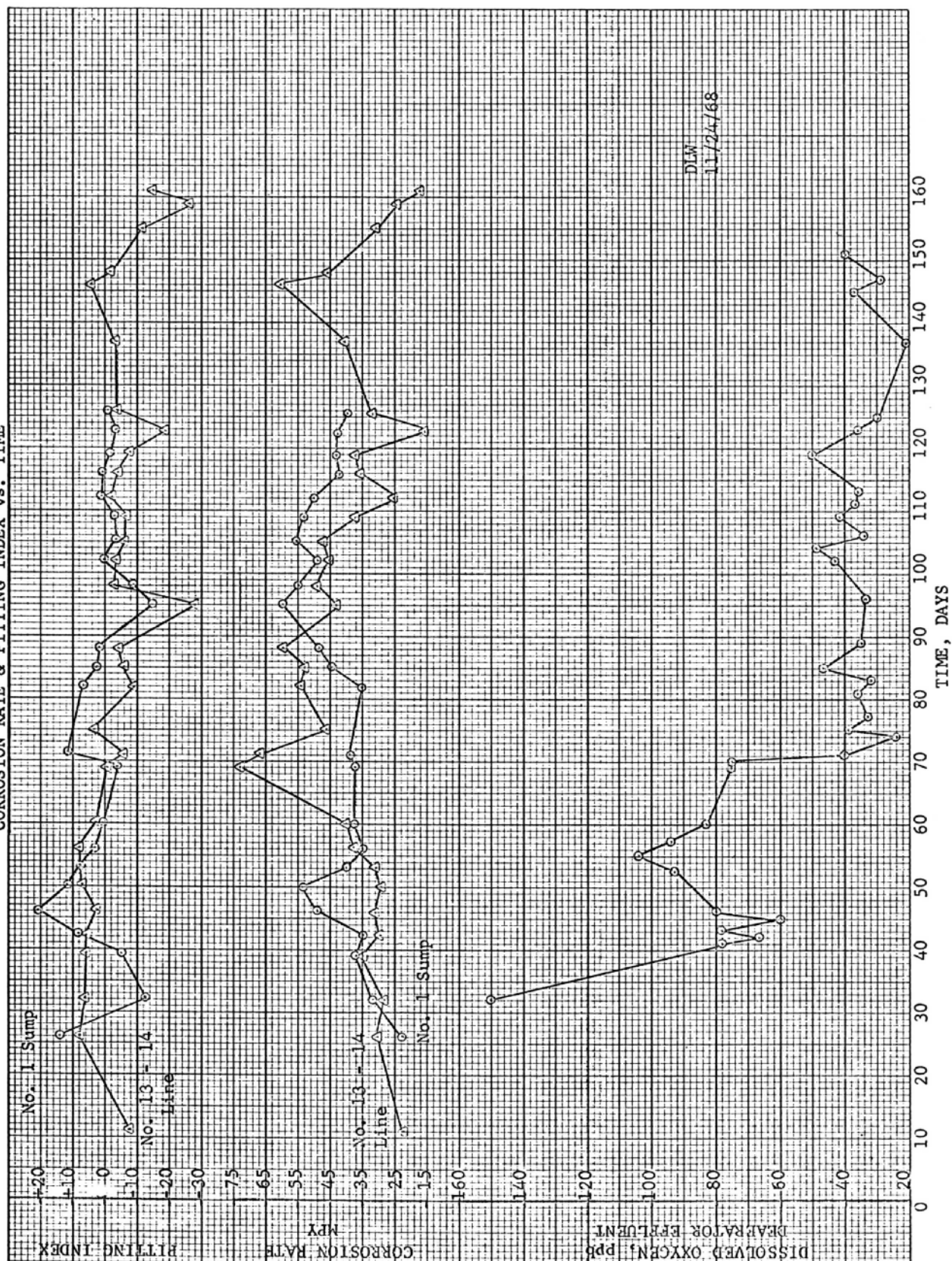


TABLE G-7-c

CORROSION RATES CALCULATED FROM PORTABLE CORRATER READINGS

Days ELAPSED	Date	Corrosion Rates, MPY		Pitting Index Change	
		No. 1 Sump	No. 13-14 Line	No. 1 Sump	No. 13-14 Line
11	4-22-68	----	22.6	----	----
26	5- 7-68	23.0	30.0	----	+ 15
32	5-13-68	31.8	28.2	- 26.6	- 2
39	5-20-68	37.3	36.5	+ 7	0
42	5-23-68	34.9	30.0	+ 13	0
46	5-27-68	50.0	32.5	+ 13	- 3
50	5-31-68	54.2	30.0	- 9	+ 5
53	6- 3-68	40.0	31.7	- 4	0
56	6- 6-68	34.9	37.0	- 4	+ 0.5
60	6-10-68	38.2	40.0	- 2	- 6.5
69	6-19-68	37.5	73.8	- 6	- 3
71	6-21-68	39.1	66.8	+ 15	- 5
75	6-25-68	----	46.2	----	+ 9
82	7- 2-68	35.8	54.2	----	- 11
85	7- 5-68	44.5	53.3	- 4	+ 3
88	7- 8-68	48.9	59.6	- 1	+ 1
95	7-15-68	60.3	43.6	- 17	- 24
98	7-18-68	55.2	50.0	+ 6	+ 25
102	7-22-68	48.9	46.2	+ 9	0
105	7-25-68	56.0	48.0	- 4	- 2
109	7-29-68	53.3	38.2	+ 1	- 1
112	8- 1-68	50.0	25.8	+ 4	+ 5
116	8- 5-68	42.8	35.8	0	- 2
119	8- 8-68	42.8	37.3	- 2	- 4
123	8-12-68	42.8	15.2	- 2	- 11
126	8-15-68	39.1	32.5	+ 2	+ 15
137	8-26-68	----	40.9	----	0
146	9- 4-68	----	60.3	----	+ 8
148	9- 6-68	----	46.2	----	- 6
155	9-13-68	----	30.8	----	- 10
159	9-17-68	----	25.0	----	- 15
161	9-19-68	----	17.3	----	+ 10.6

NOTE: Columns 3 and 4 (under Corrosion Rates, MPY) are from Figure 4,
"Operating Instructions, Model 1170, Corratel", Magna Corporation.

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- d. The large decrease in DO which occurred on Day 70, which was due to repair of a vacuum leak in the deaerator, did not show up as a corrosion rate decrease in No. 1 sump.
- e. The decline in corrosion rate at the end of the run might be due to the sulfite injection.

In conclusion, there are three reasons that may explain this apparent lack of data correlation:

- a. The corrater monitors many factors governing corrosion rate; dissolved oxygen is only one of these.
- b. It must be recognized that these data are spot, isolated readings and possibly should not be represented as a line graph. Continuous monitoring might allow for a different interpretation.
- c. The corrater's sensitivity to process variables might be greater than the plant's ability to perform at steady state conditions; i.e., 155.5 - 163.0°F at No. 13 - 14 sump line; 6.5 - 7.5 pH at No. 1 sump.

The operations of the continuous, on-line corrater during Development Run 16 is covered in section G-12 of this report.

G. RESULTS AND COMPARISONS (CONTD)

8. EFFECTIVENESS OF SHOCK CHLORINATION

This objective was to evaluate the chlorine residuals for specified injection rates and the effectiveness of shock chlorination in preventing the growth of marine organisms in processing equipment. The correct injection dosages were sought to give an outlet residual chlorine and to provide effective treatment as determined from an inspection of the water boxes of the protected exchangers. A Fischer & Porter Company Model 70 C 3660 Gas Dispenser was used to apply approximately 10 ppm Cl_2 to the P-3 circuit and 6 ppm Cl_2 to the P-1/P-2 circuit.

Following are: Figure G-8-a - Diagram of Chlorination Injection Circuits
Table G-8-a - Chlorination Data, P-1/P-2 Circuits
Table G-8-b - Chlorination Data, P-3 Circuit

Following each development run, the top water boxes of exchangers 213, 212 and 317 and the inlet, outlet and return water boxes of exchanger 318 were opened for inspection. At these times no attached-type marine organisms were found.

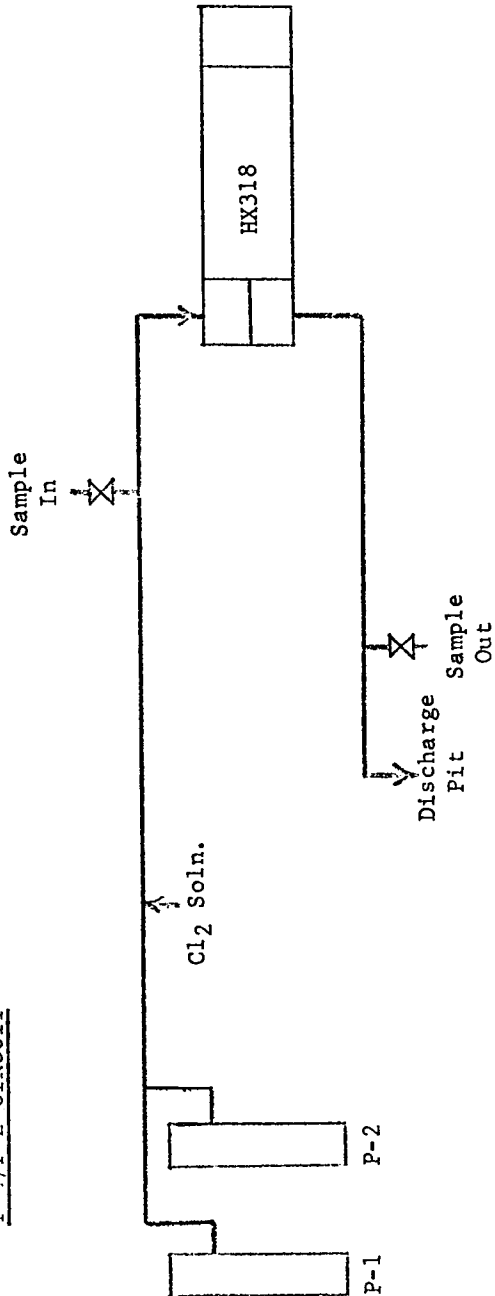
During Run 16, the procedure was made an operations task, including the timing to initiate the injection, operation of the equipment, record keeping and residual analyses. The Development Chemist set up an analytical procedure and trained the plant operators in its use. The major change for this run was a reduction of injection time to one hour for each circuit. The other change was the sequence of valve switching for flow reversals, the object being to wash out of the circuits any loosened material.

It is not felt that there is a significant economic incentive to investigate chlorination periods of less than an hour or chlorination frequencies of less than once per week.

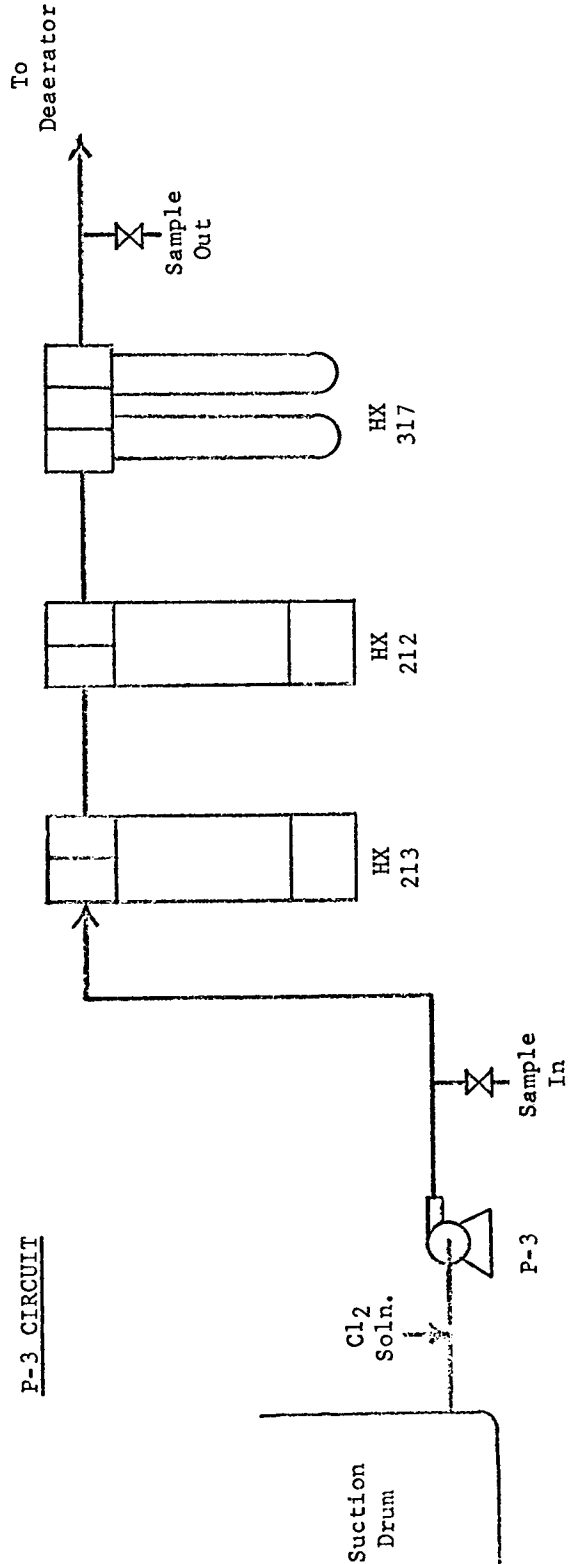
FIGURE G-8-a

CHLORINATION INJECTION CIRCUITS

P-2 CIRCUIT



P-3 CIRCUIT



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TABLE G-8-a

CHLORINATION DATA, P-1/P-2 CIRCUITS

TWO HOUR CHLORINATION PERIOD

<u>Date</u>	<u>Calculated Sea Water Flow, MPPH</u>	<u>Chlorine Used, Lbs.</u>	<u>Injection Strength</u>		<u>Residual Chlorine out of circuit ppm Cl₂</u>
			<u>Calc.</u>	<u>ppm Cl₂ Analysis</u>	
4-23-68	---	12	---	2.6	1.7
4-30-68	---	19	---	3.3	2.5
5- 7-68	---	16	---	1.8	1.4
5-13-68	2,100	14	3.3	3.4	2.5
5-20-68	2,300	20	4.3	2.1	0.5
5-27-68	1,600	16	5.0	3.1	2.1
6- 3-68	1,800	18	5.1	3.5	2.5
6-10-68	1,900	18	4.7	2.7	2.4
6-19-68	1,300	11	4.2	3.5	2.5
6-27-68	1,800	17	4.8	1.8	0.3
7- 2-68	1,600	15	4.7	2.3	1.3
7- 8-68	1,700	16	4.6	1.8	1.4
7-15-68	1,800	11	3.1	2.4	0.5
7-23-68	1,600	17	5.2	2.9	2.2
7-29-68	1,900	14	3.7	3.4	2.3
8- 5-68	1,600	14	4.4	3.8	2.1
8-12-68	1,800	16	4.4	4.0	2.8
8-26-68	1,800	17	4.7	2.9	2.1
9- 3-68	2,400	15	3.2	3.2	2.6
9-10-68	1,700	15	4.3	2.7	0.8
9-16-68	1,900	18	4.8	2.0	1.8

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TABLE G-8-b

CHLORINATION DATA, P-3 CIRCUIT

TWO HOUR CHLORINATION PERIOD

<u>Date</u>	<u>Metered Sea Water Flow, MPPH</u>	<u>Chlorine Used, Lbs.</u>	<u>Injection Strength</u>		<u>Residual Chlorine out of Circuit ppm Cl₂ Analysis</u>
			<u>Calc.</u>	<u>ppm Cl₂</u> <u>Analysis</u>	
4-23-68	465	11	11.8	7.2	5.1
4-30-68	432	8	9.3	5.5	4.1
5- 7-68	475	15	15.8	4.3	3.1
5-13-68	472	11	11.7	4.4	3.1
5-20-68	475	8	8.4	3.6	2.8
5-27-68	472	14	14.8	3.6	3.3
6- 3-68	485	10	10.3	4.6	3.0
6-10-68	482	8	8.3	4.5	3.8
6-19-68	480	8	8.3	4.7	4.5
6-27-68	477	10	10.5	---	---
7- 2-68	450	11	12.2	3.8	3.1
7- 8-68	457	6	6.6	3.7	2.7
7-15-68	467	8	8.6	3.5	3.5
7-23-68	478	8	8.4	6.7	6.2
7-29-68	482	7	7.3	7.2	5.4
8- 5-68	480	10	10.4	6.8	5.1
8-12-68	482	9	9.3	7.3	6.0
8-26-68	482	13	13.5	7.1	6.1
9- 3-68	489	10	10.2	5.5	4.9
9-10-68	498	10	10.0	6.4	4.2
9-16-68	400	8	10.0	4.5	3.8

DLW:alb
September 17, 1969

G. RESULTS AND COMPARISONS (CONTD)

9. ATU (AUXILIARY TEST UNIT) OPERATIONS

a. Introduction

The ATU (Auxiliary Test Unit) is a self-contained, totally instrumented, seven-tube evaporator that is used to test such things as scaling threshold, tube performance, venting effectiveness, etc. The ATU is set up to run in parallel with the first effect; it receives the same brine feed as the first effect.

During Development Run 15, there were three major ATU runs, all directed at determining first effect scaling threshold:

- Run 15 - 1 Base line operations at 268°F sump temperature with aluminum brass tubes.
- Run 15 - 2 Extended run at 273°F sump temperature with aluminum brass tubes.
- Sum 15 - 3 Extended run at 273°F sump temperature with stainless steel tubes.

During Development Run 16, the ATU was shortened to accommodate seven enhanced surface tubes. Test operations were directed at developing base line heat transfer coefficient data for the new tubes and determining the performance of these tubes under different operating conditions and equipment configurations, such as: flashing feed versus subcooled feed; SVL weirs versus spray nozzles; maximum venting versus no venting; etc.

b. Discussion of Runs - Development Run 15

Run 15-1 (268°F Sump): At the end of July, after being run for about 500 hours at operating conditions close to those experienced by the first effect, the ATU was shut down, blocked in, and drained. The tubes were scrubbed down with a plastic brush and washed down with fresh water. There was no sign of any scale in the tubes. The 1.84" diameter gauge ball easily passed down all of the 1.87" I.D. aluminum brass tubes. The top two or three feet of all the tubes was rather dull in appearance; the lower part of the inside surface of all the tubes was still quite shiny. There was no scale formed in the ATU tubes during the 268°F, 500-hour run.

There were several shutdown and unit upsets during this ATU run:

- 1) A bad leak developed in the discharge piping of P-2000. In order to band this leak, it was necessary for the operator to block off some steam to the ATU chest and cut off the brine feed to the ATU tubes for about half an hour.

The next day the above mentioned leak in the brine discharge piping got worse. As a consequence it was necessary to shut down the ATU long enough to have a new pipe spool put in. For approximately two hours, the ATU was fed directly from the main plant brine line to the by pass around Pump P-2000. The maximum obtainable flow in this configuration was 10 gpm. Looking through the portholes, brine flowed down all the tubes that could be seen during this mode of operation. At the same time, the steam chest was blocked in and vented to the atmosphere to avoid any possibility of overheating.

- 2) A malfunction developed in the ATU brine sump level controller. While this instrument was being worked on, the level control valve was closed; and the ATU filled up with brine. When this happened, the chest pressure went up to 40 psig; the vapors-out pressure went up to 35 psig; and brine flowed out the vapors-out line. This upset lasted no more than ten minutes.
- 3) On July 17 the ATU was shut down with the rest of the plant as follows: The steam, desuperheater water, condensate, and vent systems were all blocked in; the chest was vented to the atmosphere; the unit was water washed along with the main plant; and Pump P-2000 was shut off at the same time as P-12. The ATU was started up after the main plant was on the line and operating smoothly.
- 4) A fifteen-minute upset occurred during which the chest pressure went over 33 psig. There was no indication of any change in any of the ATU temperatures during this upset.

For the most part, the ATU ran quite smoothly during this 268°F, 500-hour run. The ATU was accomplishing approximately the same degree of concentration as the first effect during the run.

The ATU sump temperature ran from 0.2° to 0.5°F higher than the No. 1 sump temperature during most of the run. This is probably due to the pressure drop in the steam line from the ATU sump to the No. 2 Effect steam chest.

Run 15-2 (273°F Sump with Aluminum Brass Tubes): This run was started at 1800 on 7/26/68 and terminated at 1400 on 8/16/68 after 476 hours of unusually smooth operation. When the unit was shutdown and opened up for inspection, the tubes were found to be heavily fouled with calcium sulfate scale.

The plots on Figures G-9-a and b were made using data from the operator's log sheets. Only four points a day were plotted. These are raw, unreduced, uncorrected data and are plotted mainly to show trends. Table G-9-a lists the source of data for these ATU plots. Regarding the specific plots:

- 1) Sump Temperature: The plot is flat and quite steady, indicating that the operators were easily able to control the sump temperature well within the preset operating range.

FIGURE G-9-a

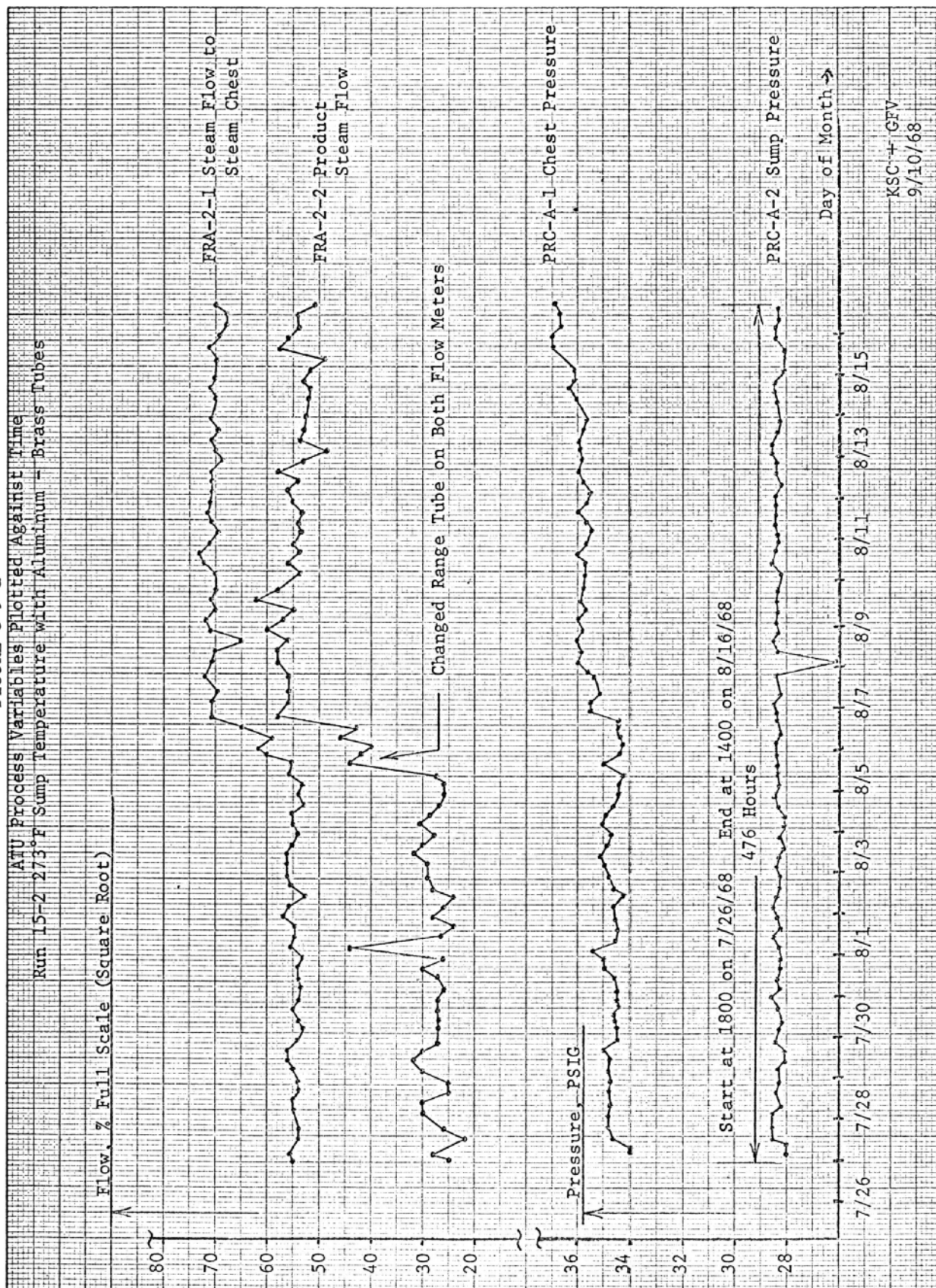


Figure G-9-b

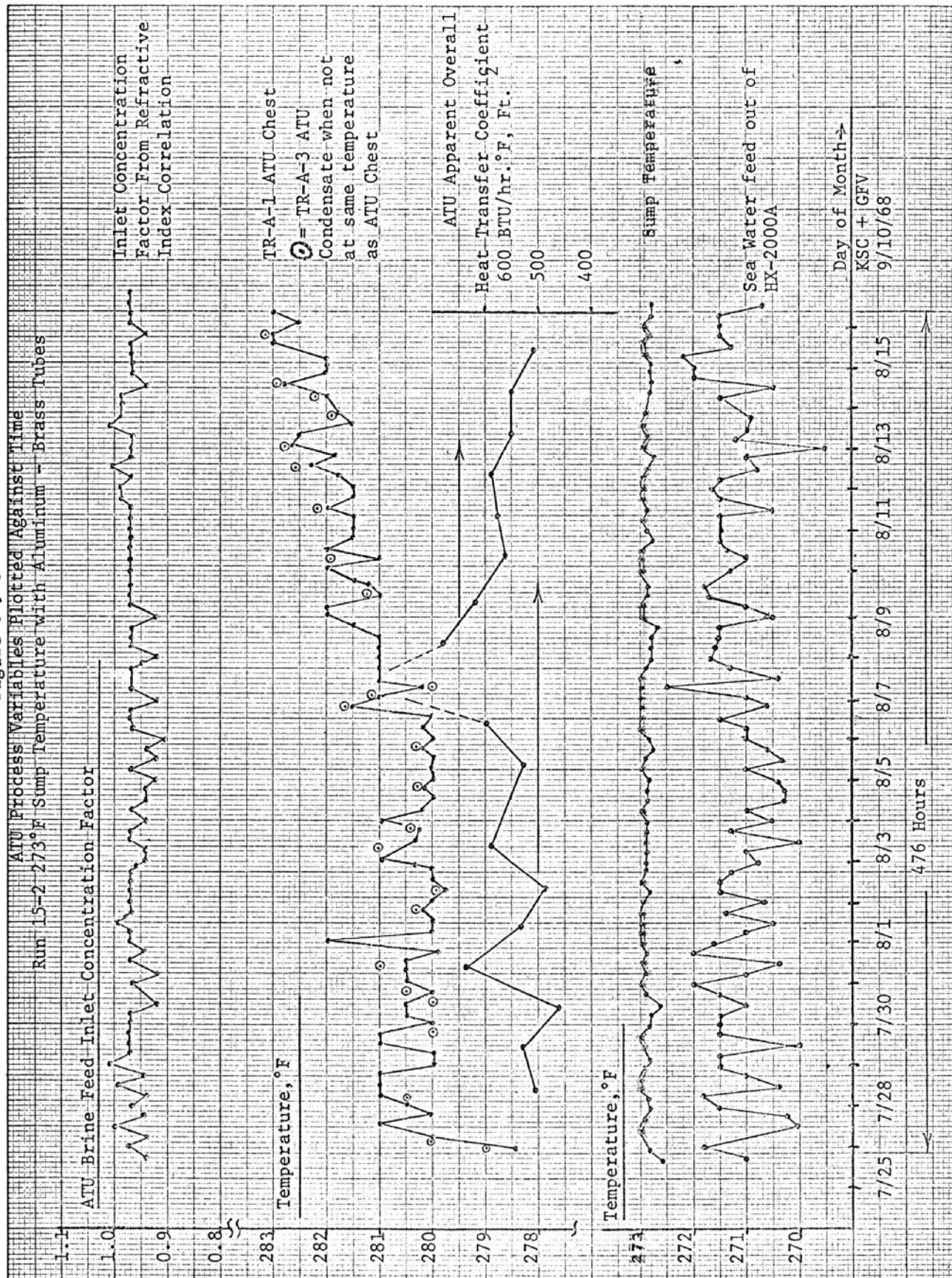


TABLE G-9-aSOURCE OF DATA FOR ATU
DATA PLOTS

<u>Points</u>	<u>Identification</u>	<u>Instrument</u>	<u>Accuracy</u>
ATU Steam Chest Temperature	TR-A-1	Thermocouple reading out on a board mounted strip chart recorder.	$\pm 2^{\circ}\text{F}$
ATU Condensate Temperature	TR-A-3	Thermocouple reading out on a board mounted strip chart recorder.	$\pm 2^{\circ}\text{F}$
Sump Temperature	-	Calibrated High accuracy mercury filled glass thermometer.	$\pm 0.2^{\circ}\text{F}$
Temperature of Seawater feed out of preheat exchanger	-	Certified high accuracy mercury filled glass thermometer	$\pm 0.2^{\circ}\text{F}$
Steam flow to ATU Steam Chest	FR-A-2-1	Board mounted strip chart recorder receiving output from DP cell across orifice plate.	$\pm 5\%$
Steam flow from ATU vapor body	FR-A-2-2	Board mounted strip chart recorder receiving output from DP cell across venturi tube	$\pm 5\%$
Chest Pressure	PRC-A-1	Board mounted strip chart recorder receiving signal from Bourdon tube system.	± 1 psig
Sump Pressure	PRC-A-2	Board mounted strip chart recorder receiving signal from Bourdon tube system.	± 1 psig
Inlet concentration factor from refractive index correlation		Spot sample, hand refractometer.	± 0.02 Concentration Factor units

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- 2) Seawater Feed Out of HX-2000a: The average preheated feed temperature for the run was about 271°F. There were frequent fluctuations of about $\pm 1^\circ\text{F}$ from this average temperature.

This plot points up the relative insensitivity of pneumatic strip chart recorders. This same temperature was sensed during the run by an expansion bulb thermometer, the output of which was changed to a 3-15 psi signal transmitted to the board-mounted Recorder-Controller TRC-A-1. This sensing device recorded a constant unvarying temperature during the course of the ATU run. This is in contrast to the fluctuations that were seen in the mercury-filled glass thermometer readings that are plotted on Figures 2 and 3.

The fluctuations in the seawater feed temperature are probably due to fluctuations in the temperature of the seawater feed to HX-2000a.

- 3) Steam Chest Temperature and ATU Condensate Temperature: Although the data are not conclusive, it looks as though the steam condensing temperature stayed between 280-281°F for the first half of the run and then gradually trended up to 282-283°F at the end of the run; thus the temperature driving force necessary to produce a constant degree of concentration increased at the end of the run. This is indirect evidence of scale formation.

Most of the time the ATU chest temperature and condensate temperature are quite close together. This indicates that the ATU desuperheater is operating satisfactorily and also that the ATU chest is operating isothermally.

- 4) Sump Pressure and Chest Pressure: The sump pressure stayed more or less constant at about 28.4 psig, which is roughly the pressure equivalent of 273°F sump temperature.

The chest pressure averaged about 34.6 psig for the first half of the run and then trended up to about 36.5 psig during the last half of the run. These pressures are roughly equivalent to the condensing temperatures that were discussed previously.

- 5) Steam Flow to Steam Chest and Product Steam Flow: These are raw, unreduced, flow chart readings. Several significant adjustments must be made in order to convert these readings to lbs/hr. The break in the middle of these curves is due to a change in the range tube at the DP cell that was made to both of these instruments. This was done to move the readings over to the high end of the scale where there is more distance (and thus more sensitivity) between scale divisions.

In general these curves are quite flat, indicating that the unit was being carefully controlled.

- 6) Brine Feed Inlet Concentration Factor: The operators were asked to reset blending rate in order to keep the concentration factor between 0.94 and 0.99. It can be seen that they were successful in keeping the feed concentration factor within the preset range. There is, however, considerable fluctuation in the inlet concentration factor within the given control range. This fluctuation is really beyond the control of the operating personnel and is a fact of life the way the ATU is set up.
- 7) Overall Heat Transfer Coefficient: An apparent overall heat transfer coefficient has been calculated from the operator log sheet data. Roughly speaking, these transfer coefficients are probably good to 15%.

The apparent overall heat transfer coefficient for the ATU averages about 550 Btu/hr, °F, ft² over the whole run. Starting at a value at about 500 Btu/hr, °F, ft², the coefficient trended up to in excess of 600 Btu/hr, °F, ft² during the middle of the run and then dropped back in a fairly even fashion to 500 Btu/hr, °F, ft². The initial rise in heat transfer coefficient toward the end of the run is consistent with the high degree of scaling that was found.

Run 15 - 3 (273°F Sump with Stainless Steel Tubes): For Run 15 - 3, the ATU was retubed with seven 2" x 20 BWG x 22' long, welded 316 L stainless steel tubes all from Red Heat No. 41923. These tubes have a nominal I.D. of 1.93". Operator data for Run 15-3 is plotted on Figure G-9-c, d, & e.

This run was started at 2200 on 8/31/68 and terminated at 1230 on 9/20/68 after a run of 458 hours, which was interrupted by one shutdown during which the tubes were inspected. Additionally, there were two serious upsets during the run. The first occurred on 9/11/68 when problems developed in the main plant acid pumping system. The net result was that unacidified seawater was fed to the ATU for a period of about an hour. Later, due to collateral power usage, one of the switch hours breakers was kicked out which shut down the ATU brine pump. For a short period of time, the ATU was run with no brine flow. Since the ATU steam flow was on manual, the chest was significantly overpressured. The upset condition was quickly rectified, and the ATU was brought back to the preset operating conditions for the balance of the run.

On September 13 the ATU was shut down for an interim inspection. The 1.84" gauge ball passed easily down all of the tubes. In an unwashed state, the tubes appeared shiny with a light speckle of adherent material on the surface. The scale was hard and not easily removed from the tube surface. There was not enough of it significantly to affect the heat transfer rate.

At the end of the run, the ATU tubes were again inspected. The 1.84" gauge ball passed easily down all of the tubes before they were washed. The 1.90" gauge ball would pass no further than two feet down any of the tubes even after they had been washed down with a nylon brush and fresh

FIGURE G-9 - c

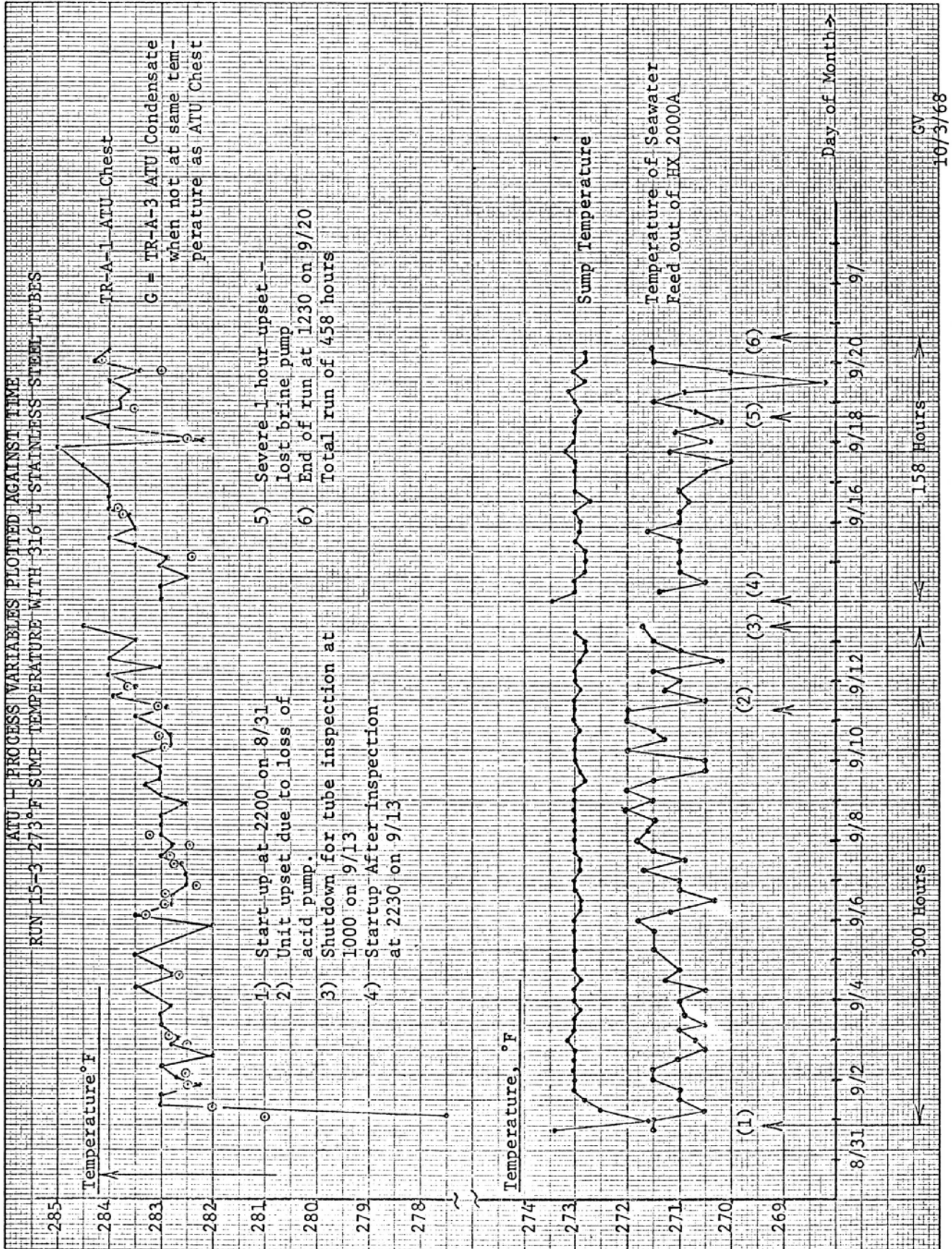


FIGURE G-9-d

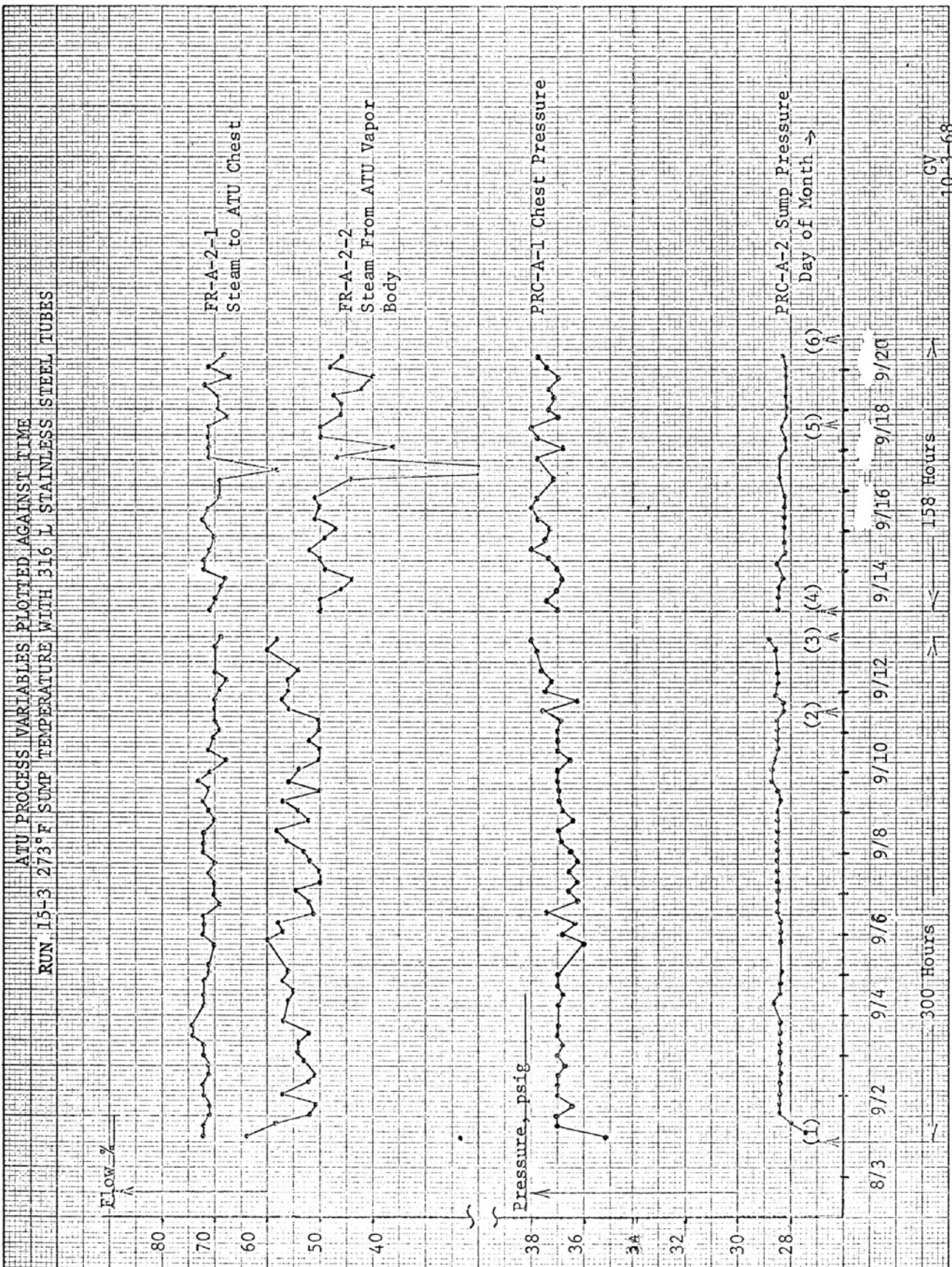
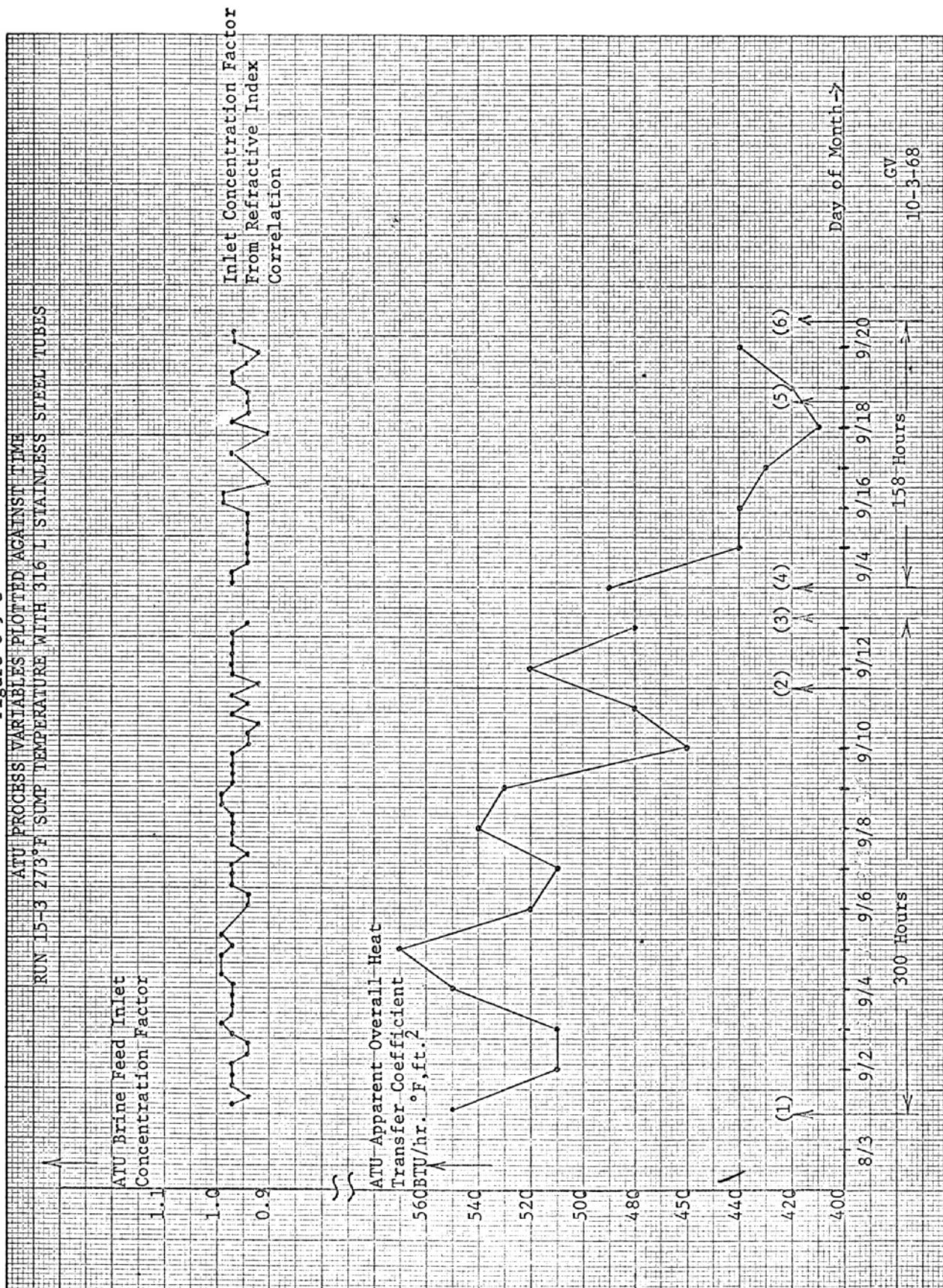


Figure G-9-e



water. The tubes had the same shiny, speckled appearance that they showed when inspected at the interim shutdown. There were some bigger patches of continuous scale; however, there was still not enough scale to affect heat transfer rates. It is estimated that about 5% of the surface area was spotted with scale. Referring to the data plots:

- 1) Sump Temperature: The plot is flat and quite steady, indicating that the operators were easily able to control the sump temperature well within the preset operating range.
- 2) Seawater Feed Out of HX-2000a: The average preheated feed temperature for the run was about 271°F with frequent fluctuations of about $\pm 1^\circ\text{F}$ from this average temperature.
- 3) Steam Chest Temperature and ATU Condensate Temperature: Although the data are not conclusive, it looks as though the steam condensing temperature stayed at about 283°F for the majority of the run and then gradually trended up to about 284°F at the end of the run; thus the temperature driving force necessary to produce a constant degree of concentration increased at the end of the run.

Most of the time the ATU chest temperature and condensate temperature are quite close together. This means that the ATU desuperheater was operating satisfactorily and also that the ATU chest was operating isothermally.

- 4) Sump Pressure and Chest Pressure: The sump pressure stayed more or less constant at about 28.4 psig, which is roughly the pressure equivalent to 273°F sump temperature.

The chest pressure averaged about 37 psig for the majority of the run and seemed to trend upward to about 37.5 psig during the last part of the run. These pressures are roughly equivalent to the condensing temperatures that were discussed previously.

- 5) Steam Flow to the Steam Chest and Product Steam Flow: These are raw, unreduced, flow chart readings. Several significant adjustments must be made in order to convert these readings to lbs/hr. These flows remained relatively constant over the course of the run.
- 6) Brine Feed Inlet Concentration Factor: The operators were asked to reset blended rate in order to keep the concentration factor between 0.94 and 0.99. It can be seen they were successful in keeping the feed concentration factor within the preset range. There were some excursions to below 0.94 concentration factor feed toward the end of the run when the operators found it very difficult to maintain control of blending at the lower main plant feed rates.
- 7) Overall Heat Transfer Coefficient: An apparent overall heat transfer coefficient has been calculated from the operator log sheet data. Roughly speaking, these transfer coefficients are probably good to 15%.

The apparent overall heat transfer coefficient for the ATU started at about 550 Btu/hr, °F, ft² at the beginning of the run and slowly trended down to about 420 Btu/hr, °F, ft² at the end of the run. This gradual downward trend is probably due to fouling of the new stainless steel tubes. It is doubtful whether the amount of actual non-washable scale that was observed inside the tubes could significantly affect the overall heat transfer coefficient.

c. Discussions of Runs - Development Run 16

At the beginning of Development Run 16, the ATU was modified and fitted with 14' long, spirally enhanced 2" O.D. Phelps Dodge tubes. These 90/10 copper/nickel tubes had a 0.049" wall thickness. The following tests were undertaken with the ATU during Development Run 16:

- Run 16 - 1 Shake-down to evaluate ATU performance with shorter tubes.
- Run 16 - 2 Constant operation tube performance test at No. 1 Effect conditions.
- Run 16 - 3 Venting studies.
- Run 16 - 4 Effect of flashing feed
- Run 16 - 5 SVL weir test.

The target operating conditions for ATU Run 16 - 2 at No. 1 Effect conditions are shown on Table G-9-b. These represent the limits within which the operators controlled the unit.

d. Heat and Material Balance Data

Development Run 15: The special ATU heat and material balance data that were taken during Development Run 15 are presented in Table G-9-c. These data were taken when the unit was well lined out and operating smoothly. The critical flow meters were zeroed and spanned just prior to the data-taking period. Seawater feed and brine sump samples were composited over the entire data-gathering period.

Certification corrections and stem corrections have been applied to the temperatures. The weight flow rate of the streams into and out of the ATU have been calculated from the observed pressure drop across the measuring device and the physical properties of the fluid flowing.

In Table G-9-c the metered steam production rates are compared with those calculated on a salt balance basis. There is considerable variance between the two ways of calculating production rate. In the early runs, the salt balance production is considerably higher than the metered production. This is probably due to the fact that for convenience the seawater inlet sample was taken at the discharge of the main plant blended brine pump

TABLE G-9-b

TARGET OPERATING CONDITIONSA.T.U. RUN 16-2 AT NO. 1 EFFECT CONDITIONS WITH"SPIRAL-ENHANCED TUBES"

	TARGET	LIMITATIONS	CONTROL MODE	REMARKS
<u>I. CONTROL POINTS</u>				
<u>A. Flow Rates</u>				
1. Sea Water Feed (FRC-A-3-1)	23.0 GPM (11,000 lbs/hr)	---	Automatic Flow Control	---
2. Steam Flow to Steam Chest (FR-A-2-1)	As needed to control steam out.	31.5 PSIG Maximum chest pressure. 267°F Maximum Sump temperature.	Automatic Pressure Control	Steam flow controlled by chest pressure controller (PRC-A-1). Sump temperature read on calibrated thermometer.
3. Steam Flow From ATU Sump (FR-A-2-2)	6.8 out of 10 Square Root (500 lbs./hr.)	6.7 - 6.9	---	Steam out flow rate controlled by chest pressure controller (PRC-A-1)
4. Vent Flow (FE-A-4)	45 lbs./hr.	40-50 lbs./hr.	Manual	Bottom vent only.
5. Condensate Flow Out (FI-A-2)	As Needed	---	Automatic Level Control	---
6. Brine Out (FR-A-3-2)	As Needed	---	Automatic Level Control	---
<u>B. Pressures</u>				
1. Chest Pressure (PRC-A-1)	As needed to control steam out.	31.5 PSIG maximum chest pressure. 267°F maximum sump temperature.	Automatic Pressure Control	Sump temperature read on calibrated thermometer.
2. Sump Pressure (PRC-A-2)	---	---	Manual	Control Valve Wide Open

TABLE G-9-b

(Contd)

	TARGET	LIMITATIONS	CONTROL MODE	REMARKS
3. Main Steam Reduced Pressure (PCV-A-3)	55 PSIG	50 - 60 PSIG	Automatic Pressure Control	---
C. <u>Temperatures</u>				
1. Desuperheated Steam Temperature (TCV-A-4)	305°F	295 - 315°F	Manual	Use plant desuperheating water without booster pump if possible.
2. Sea Water Temperature (TRC-A-1)	264.5°F	264 - 265°F	Automatic Temperature Control	Certified Thermometer.
II. <u>EQUIPMENT LINE UP</u>				
A. Vent	=	Bottom Vent Only		
B. Sea Water Source	=	P-2000 through HX-2000		
C. Steam Product To	=	No. 2 Effect Chest		
D. Condensate To	=	No. 3 Flash Tank		
E. Brine Out To	=	No. 2 Effect Sump		
F. Steam Inlet Drain	=	Wide Open		
G. HX-2000 Condensate	=	Through Steam Trap to No. 3 Flash Tank		

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TABLE G-9-c

Development Run 15

Summary of ATU Heat & Material Balance Data

I. BACKGROUND INFORMATION				15-2				15-3			
A. ATU Run Number	→	6/26/68 to 7/17/68	15-1	7/26/68 to 8/16/68	15-2	8/31/68 to 9/20/68	15-3	Repeat of Run 15-2	7 x 2" OD x 22' Long	x 20 BWG	316L Stainless Steel
B. Inclusive Dates	→	No. 1 Eff. Conditions		272°F Sump Temperature							
C. Intent of Run	→	7 x 2" OD x 22' Long		Same as Run 15-1							
D. Type of Tubes	→	Arsenical Aluminum									
		x 16 BWG									
		Brass									
II. BASIC DATA											
A. Temperatures, °F											
1. Steam Chest (1)		273.7	275.6	275.4	278.9	279.2	280.8	280.7	281.0	280.2	283.4 281.8 284.3
2. Sea Water Feed		264.3	265.4	265.0	272.0	270.8	271.3	270.2	270.2	270.9	270.6 270.0 270.5
3. Brine Sump (2)		267.2	268.2	268.0	272.9	272.9	273.0	272.9	272.8	272.8	272.9 273.0 272.8
B. Chlorinity, Wt. %											
1. Sea Water Feed		1.883	1.833	2.013	1.765	1.965	1.965	1.920	1.956	1.954	1.929 1.937 2.027
2. Sump Brine		1.945	1.932	2.136	1.844	2.054	2.054	2.019	2.101	2.031	2.009 2.003 2.101
C. Steam Production Rates, lbs/hr											
1. By Salt Balance		360	570	610	480	480	480	530	760	420	450 360 390
2. By Flow Meter		290	400	390	330	350	470	480	440	470	470 360 420
III. MATERIAL BALANCE, LBS./HR.											
IN											
Sea Water Feed (Metered)		11400	11000	10600	11100	11000	11000	10900	11000	10900	11100 10900 10900
Steam to Chest (Metered)		440	520	520	480	500	630	650	570	550	550 560 550

NOTES: (1) Calculated
(2) Calibrated Glass, Mercury Filled Thermometer

TABLE G-9-c
(Contd)

III. Cont'd. OUT	11110 10600 10210				10770 10650 10530 10420 10560 10430				10630 10540 10480			
	Brine Out Sump (By Difference)											
Condensate (Steam to Chest Minus Vent Out)	400	480	480		440	460	590	610	530	510	500	500
Vent Out (Metered)	40	40	40		40	40	40	40	40	40	50	50
Steam Produced	290	400	390		330	350	470	480	440	470	470	420
IV. HEAT TRANSFER												
A. Outside Heat Transfer Area (A _o), ft. ²	← ————— 80.6 ————— →											
B. Steam Chest Temperature - Brine Sump Temperature (L.M.T.D.), °F	6.5	7.4	7.4		6.0	6.3	7.8	7.8	8.2	7.4	10.5	11.5
C. Heat Transferred Through Tubes To Brine (Q _T), MM Btu/hr.	0.27	0.37	0.36		0.31	0.33	0.44	0.45	0.41	0.44	0.44	0.39
D. Overall Evaporative Heat Transfer Coefficient, Btu/hr, °F, ft. ²	520	630	610		640	640	700	710	620	730	520	470
											420	420

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(P-3). Arrangements have been made to take the seawater inlet sample at the discharge of ATU Pump P-2000 to avoid any future such difficulties. The steam production rates determined by the two different methods are in fairly good agreement for ATU Run 15-3.

A material balance for each set of data is also shown in the table. The metered flow for four of the six streams is shown. The condensate indicating flow meter is too erratic to be used; therefore, the condensate rate was calculated. The brine out of the ATU was calculated by difference to force the material balance. There is reasonable agreement between the calculated and the metered brine out-flow rate. An extra, non-significant digit is carried along in the brine out of sump numbers in order to close the balance.

Overall heat transfer coefficients have been calculated and are presented in the table for each set of heat and material balance data taken around the ATU during Development Run 15. These coefficients are good to $\pm 5\%$. The accuracy is limited by the temperature and steam production measurements.

Development Run 16: The ATU heat material balance data that were taken and reduced for Development Run 16 are shown on Table G-3-d. During this entire run, the ATU contained seven Phelps Dodge spirally enhanced tubes.

Considering the precision of the various data-gathering equipment, the estimated accuracy of these heat transfer coefficients is $\pm 5\%$. Before any other runs are undertaken, the ATU thermocouples should be recalibrated.

For most of Development Run 16, the ATU was run on unblended seawater. This was done to avoid any chance of scaling the experimental tubes. During most of the shutdowns, the ATU tubes were carefully inspected for any signs of scale. No scale was formed in the ATU tubes during Development Run 16.

e. Conclusions and Recommendations

In Run 15-1 no scale was formed after about 500 hours of operation at nominal No. 1 Effect conditions. In Run 15-2 the ATU aluminum brass tubes were heavily scaled with positively identified calcium sulfate anhydrite after having been run smoothly and continuously at a 273°F sump temperature for 476 hours. In Run 15-3 the 316 L stainless steel tubes were only slightly scaled after having been run irregularly at the sump temperature of 273°F for 458 hours. It can be said that at these conditions of temperature, pressure, and concentration factor the incipient scaling point is in the neighborhood of 273°F. Certainly no scale was observed during the earlier 268°F sump temperature run, and at least some scale was observed to form during both of the 273°F runs. It must be remembered that the calcium sulfate anhydrite scaling

TABLE G-9-d

Performance of Spiral Enhanced TubesATU - Development Run 16

(RUN DATA)

<u>1969 Operations</u>	<u>Date Start</u>	<u>Date Finish</u>	<u>Duration Hours</u>	<u>To Test</u>	<u>Type Weirs</u>	<u>Feed Condition</u>	<u>Tube Condition at End of Run</u>
<u>Run 16-1</u>	--	--	--	Shake-down	--	--	No Scale
<u>Run 16-2</u>	2/5	3/2	600	Tube Performance	Spray Nozzles	Non-Flashing	No Scale
2/7				Tube Performance	Spray Nozzles	Non-Flashing	
2/10				Tube Performance	Spray Nozzles	Non-Flashing	
2/18				Tube Performance	Spray Nozzles	Non-Flashing	
2/24				Tube Performance	Spray Nozzles	Non-Flashing	
2/26				Tube Performance	Spray Nozzles	Non-Flashing	
<u>Run 16-3</u>	3/6	4/2	400	Venting	Spray Nozzles	Non-Flashing	No Scale
3/7	3/6	3/8	50	Venting	Spray Nozzles	Non-Flashing	
3/13	3/12	3/18	140	Venting	Spray Nozzles	Non-Flashing	
3/13				Venting	Spray Nozzles	Non-Flashing	
3/26	3/24	4/2	210	Venting	Spray Nozzles	Non-Flashing	
<u>Run 16-4</u>	4/14	4/22	190	Flashing Feed	Spray Nozzles	Flashing	No Scale
4/16				Flashing Feed	Spray Nozzles	Flashing	
4/17				Flashing Feed	Spray Nozzles	Flashing	
<u>Run 16-5</u>	4/27	4/29	50	SVL Weirs	SVL Weirs	Non-Flashing	No Scale
4/28				SVL Weirs	SVL Weirs	Non-Flashing	
4/29				SVL Weirs	SVL Weirs	Non-Flashing	
<u>Run 16-6</u>	4/29	5/3	120	Scaling	SVL Weirs	Non-Flashing	No Scale
5/9				Scaling	SVL Weirs	Non-Flashing	

(1) $A_o = 51.3 \text{ Ft.}^2$

(2) Calibrated Thermocouples

TABLE G-9-d

Performance of Spiral Enhanced Tubes

ATU - Development Run 16

(OPERATING DATA)

<u>1969 Operations</u>	<u>Sump Temp.</u> <u>°F(2)</u>	<u>Feed Temp.</u> <u>°F(2)</u>	<u>Vent Rate</u>		<u>Feed Concentration</u>		<u>Feed Rate</u> <u>#/Hr/Tube</u>	<u>Overall (1) Heat</u> <u>Transfer</u> <u>Coefficient</u> <u>Btu/Hr., °F, Ft. 2</u>
			<u>#/Hr.</u>	<u>% Steam In</u>	<u>Wt. % Total</u>	<u>Dissolved Solids</u>		
<u>Run 16-2</u>								
2/7	266.7	263.1	20	5	3.02		1600	880
2/10	266.9	262.5	20	5	3.07		1600	820
2/18	265.7	264.3	40	8	2.56		1600	880
2/24	263.3	261.1	40	8	2.77		1600	780
2/26	265.9	263.3	40	8	2.65		1600	870
<u>Run 16-3</u>								
3/7	265.0	261.1	40	8	2.37		1600	770
3/13	265.7	262.1	0	0	2.40		1600	800
3/13	265.3	261.5	110	19	2.61		1600	800
3/26	266.0	263.2	290	39	3.11		1600	860
<u>Run 16-4</u>								
4/16	265.8	267.0	50	12	2.22		1600	830
4/17	264.9	265.7	50	11	1.95		1600	780
<u>Run 16-5</u>								
4/28	265.9	264.6	40	9	2.09		1600	820
4/29	266.6	264.8	50	11	2.23		1600	840
<u>Run 16-6</u>								
	267.0	263.6	50	7	3.64		1600	860

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point is not a uniquely determinable independently derived temperature, but rather it is a function of such things as concentration factor, brine temperature, residence time, nature of tube material, nature of any suspended phase in the brine, pH, temperature driving force, turbulence in the tube, etc.

In Development Run 15, coefficients all ran lower than the 700-800 Btu/hr, °F, ft² coefficients that are normally observed in the highest temperature (268°F) effect. The coefficients for Run 15-2 are significantly higher than the coefficients for ATU Run 15-1. This is partly due to the higher average operating temperatures during the later run. Although heavy scaling of the ATU tubes occurred during the later part of Run 15-2, there is no indication in these data of the decline in overall heat transfer coefficient that usually accompanies heavy scaling. The heat transfer coefficients for ATU Run 15-3 are significantly lower than those observed in the previous runs. This is probably due to the fact that stainless steel tubes were used in the ATU during this run.

In Development Run 16, the Phelps Dodge spirally enhanced tubes had heat transfer coefficients ranging from 770 to 880 Btu/hr, °F, ft², operating on raw, unblended seawater at a nominal sump temperature of 265°F. The heat transfer coefficients for these enhanced tubes were higher than the heat transfer coefficients for the smooth tubes in the ATU in Development Run 15. The values are, however, about what would be predicted from past data for a seawater evaporator operation at 265°F.

It is not known whether the higher heat transfer coefficients were due to the tube enhancement or due to the fact that the ATU was operating with 14' long tubes in a smaller steam chest. Test runs with 14' smooth 90/10 copper/nickel tubes were planned but not carried out because of the abrupt termination of Development Run 16. This test work will have to be repeated along with the projected base line run before any conclusions regarding relative performance of the spirally enhanced tubes can be made.

Runs were made to investigate the effect of variations in vent rate on the overall heat transfer coefficient in the ATU. The bottom chest vent on the ATU was regulated from totally closed to wide open with data being taken at the terminal and some intermediate points. According to theory, high steam velocity parallel to the tubes should increase the steam side film coefficient by keeping the tube surface clear of non-condensables. The results on Table G-9-d show almost no change in overall heat transfer coefficient when varying the bottom chest vent rate from zero up to 39% of the steam into the steam chest.

During the first part of the Development Run 16, the ATU tubes were equipped with porcelain spray nozzles; SVL weirs were substituted during the last part of the run. There is no significant increase in the heat transfer performance that could be attributed to the SVL weirs. According to these data, both distributing devices work equally as well with the spirally

enhanced tubes. The spray nozzles do use up some pressure drop. The spray nozzles were used with both a non-flashing and a flashing feed. Changing the condition of the feed did not significantly change the heat transfer performance of the spirally enhanced tubes.

It is recommended that future work with the ATU be discontinued and the unit be scrapped. The original intention was to be able to run a small "package" evaporator in parallel with various effects in the Freeport Plant. Experience has shown that a little evaporator such as this requires operating attention all out of proportion to its size. Additionally, any such unit needs an extremely high level of instrumentation in order to produce significant and relevant data. It is our recommendation that small units can be more appropriately and effectively operated at small unit test stations, such as Wrightsville Beach.

It is also recommended that the promising spirally enhanced tubes be further tested in small evaporators to determine their actual performance relative to smooth tubes of the same length, thickness, and composition.

G. RESULTS AND COMPARISONS (CONTD)

10. CHEMICAL ANALYSES

a. Calcium, Magnesium and Sulfate

Composite samples were collected on a routine basis throughout Development Runs 15 and 16 for calcium and sulfate analyses. During Development Run 15, the calcium analyses were run by a titrimetric analyses as described in Standard Methods for the Examination of Water and Waste Water, American Public Health Association, Page 67 (1960). This method involves titration of the calcium with EDTA at a pH and with an indicator which eliminates any magnesium interference. During Development Run 16, the calcium determinations were run on the Atomic Absorption Unit.

The sulfate analyses were run by a turbidimetric method like that described in the ASTM Standard Method D-516, non-referee procedure "A". This method involves formation of the precipitate of barium sulfate and uniform suspension of the precipitate. The result is found by reading the amount of turbidity on a colorimeter and interpolation of the reading on a calibration curve to find the sulfate concentration.

The results of the calcium and sulfate determinations on raw and blended sea water samples collected at various times are shown on Tables G-10-a and b. Also shown on these tables are the concentration factors of these samples and some calculated values.

Of particular interest are the calcium and sulfate values of raw and blended sea waters adjusted to a 1.0 concentration factor. On the basis of the adjusted raw sea water data, it appears that the local gulf water contains about as much sulfate and a little less calcium than normal sea water. The calcium and sulfate values both fluctuate; the ratio between sulfate concentration and calcium concentration that has been calculated for these raw sea water samples also fluctuates from a low of 5.9 to a high of 7.9. It appears that there is significant random, independent, variation in both the calcium and the sulfate concentration in the local Gulf of Mexico brine.

The blended sea water---adjusted to a 1.0 concentration factor---tends to run a little more highly concentrated in both calcium and sulfate than does normal sea water.

As with raw sea water, there is considerable variability in the blended sea water concentration of both calcium and sulfate. The ratio bet-

TABLE G-10-n
Results of Calcium & Sulfate Analyses - Development Run 15

Date 1968	Raw Sea Water		Blended Sea Water		ppm Sulfate		ppm Calcium		Raw Sea Water		Blended Sea Water		Concentration Adjusted to 1.00		Concentration Adjusted to 1.00		Concentration Product Blended Sea Water (Moles) ² X 10 ⁴	ATU Run No.
	Cal- cium ppm	Sul- fate ppm	Cal- cium ppm	Sul- fate ppm	Raw Water	Blended Sea Water	Cal- cium ppm	Sul- fate ppm	C. F.	Concentration Adjusted to 1.00	C. F.	Cal- cium ppm	Sul- fate ppm	C. F.	Cal- cium ppm	Sul- fate ppm		
5/7	---	---	415	3000	---	7.2	---	---	---	---	---	0.920	450	3300	3.2	---		
5/9	---	---	450	3300	---	7.3	---	---	---	---	---	0.966	465	3400	3.8	---		
5/20	---	---	435	3000	---	6.8	---	---	---	---	---	0.874	495	3400	3.4	---		
5/23	---	---	425	3000	---	7.0	---	---	---	---	---	0.920	460	3300	3.3	---		
5/27	---	---	415	2900	---	7.0	---	---	---	---	---	0.966	430	3000	3.1	---		
5/31	---	---	415	2900	---	7.0	---	---	---	---	---	0.920	450	3100	3.1	---		
6/3	---	---	430	3200	---	7.4	---	---	---	---	---	0.966	445	3300	3.6	---		
6/6	---	---	415	2800	---	6.7	---	---	---	---	---	0.920	450	3000	3.0	---		
6/10	---	---	435	2800	---	6.4	---	---	---	---	---	0.920	475	3000	3.2	---		
6/21	---	---	385	3000	---	7.7	---	---	---	---	---	0.990	390	3000	3.0	---		
6/25	---	---	350	2900	---	8.3	---	---	---	---	---	0.830	420	3500	2.6	---		
6/28	225	1500	425	2600	5.9	6.7	0.603	425	2500	0.964	440	2700	2.9	15-1	15-1	15-1		
7/1	320	2000	390	2600	6.3	6.7	0.799	400	2500	0.960	405	2700	2.7	15-1	15-1	15-1		
7/5	220	1300	390	2300	5.9	5.9	0.534	415	2400	0.913	425	2500	2.3	15-1	15-1	15-1		
7/8	240	1400	415	2500	5.8	6.0	0.584	410	2400	0.982	420	2500	2.7	15-1	15-1	15-1		
7/15	250	1900	390	3100	7.6	7.9	0.620	405	3000	0.992	395	3100	3.1	15-1	15-1	15-1		
7/17	290	2300	385	3000	7.9	7.7	0.787	370	2900	1.060	365	2800	3.0	---	---	---		
7/29	295	2300	400	3200	7.7	8.0	0.771	385	3000	1.033	385	3100	3.3	15-2	15-2	15-2		
8/1	345	2500	400	2800	7.2	7.0	0.896	385	2800	1.040	385	2700	2.9	15-2	15-2	15-2		
8/7	320	2300	430	2900	7.2	6.7	0.843	380	2700	1.049	410	2800	3.2	15-2	15-2	15-2		
8/9	305	2200	430	2800	7.2	6.5	0.804	380	2700	0.997	430	2800	3.1	15-2	15-2	15-2		
8/12	345	2400	400	3100	6.9	7.8	0.871	395	2800	1.070	375	2900	3.2	15-2	15-2	15-2		
8/27	310	2300	400	3000	7.4	7.5	0.766	405	3000	0.987	405	3000	3.1	---	---	---		
9/3	---	---	410	3100	---	7.7	---	---	---	---	---	0.97	420	3200	3.3	15-3	15-3	
9/6	---	---	400	3100	---	7.8	---	---	---	---	---	0.97	410	3200	3.2	15-3	15-3	
9/9	---	---	405	3100	---	7.7	---	---	---	---	---	0.97	415	3200	3.3	15-3	15-3	
9/12	---	---	410	3100	---	7.7	---	---	---	---	---	0.97	420	3200	3.3	15-3	15-3	
9/19	---	---	420	3000	---	7.1	---	---	---	---	---	0.97	435	3100	3.3	15-3	15-3	
9/20	---	---	415	3200	---	7.7	---	---	---	---	---	0.99	420	3200	3.5	15-3	15-3	
"Normal St."	---	---	---	---	6.62	6.6	1.00	410	2700	1.00	410	2700	2.9	---	---	---		

TABLE G-10-b

RESULTS OF CALCIUM & SULFATE ANALYSES - DEVELOPMENT RUN 16 (NO BLENDING)

Date	Raw Sea Water		Sulfate ppm Calcium ppm	Concentration Factor	Raw Sea Water		Concentration Adjusted to 1.0 Concentration Factor
	Calcium ppm	Sulfate ppm			Calcium ppm	Sulfate ppm	
1968							
12/30	320	2,500	7.8	0.81	400		3,100
1969							
1/3	310	2,400	7.8	0.74	420		3,200
1/8	325	2,500	7.7	0.78	420		3,200
1/14	320	2,300	7.2	0.74	430		3,100
1/17	325	2,400	7.4	0.78	420		3,100
1/20	330	2,300	7.0	0.74	450		3,100
2/4	310	2,200	7.1	0.74	420		3,000
2/18	330	2,500	7.6	0.78	420		3,200
3/4	290	2,200	7.6	0.71	410		3,100
3/14	260	1,800	6.9	0.64	410		2,800
3/19	320	2,600	8.1	0.85	380		3,100
3/28	320	2,700	8.4	0.92	350		3,000
4/14	250	1,900	7.6	0.64	390		3,000
"Normal"							
Sea Water	---	---	6.6	1.0	410		2,700

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ween sulfate and calcium also varies considerably. However, there does seem to be a day-by-day agreement in the sulfate to calcium ratio between the raw and blended sea water samples.

The molar concentration product⁽¹⁾ has been calculated for the blended sea water samples, unadjusted for concentration factor. With regard to the calcium sulfate scaling potential, it is not possible to calculate a rigorous solubility product without knowing the activity of the brine. However, it is known that in brines approaching the concentration of raw sea water, the activity coefficient does not vary a great deal with small changes in ionic strength. Thus, all other things being equal, the activity coefficient for 0.96 concentration factor blended sea water would be about the same as the activity coefficient for 1.0 concentration factor blended sea water. Under these circumstances, the molar concentration product can be used qualitatively to predict relative scaling tendency.

Looking at the data on Table G-10-a, it can be seen that the molar concentration product for the blended sea water was generally about 10% higher than the molar concentration product for normal sea water. This implies that---all other things being equal---the scaling potential of the Freeport blended sea water is greater than the scaling potential of normal sea water. Thus, from the standpoint of calcium and sulfate concentration alone, the Freeport 1st Effect high temperature scaling data can be considered conservative. Note that for ATU Test Run 15-1, during which no scale was formed, the molar concentration product of the blended sea water (and ATU feed) was lower than the molar concentration product of normal sea water. The reverse was true for ATU Runs 15-2 and 15-3; that is, the molar concentration product of the feed was high---relative to normal sea water---and scale was formed.

Toward the end of Development Run 16, a technique for determining magnesium content on the atomic absorption unit was developed. The weekly composite samples were analyzed for magnesium content; the results are shown on Table G-10-c. Also shown are the magnesium contents of the samples adjusted to a concentration factor of 1.0; these are almost 20 per cent higher than the normal sea water value.

These results show that there can be a considerable deviation in composition between actual and normal sea water. In designing a desalination plant, it is therefore, important to obtain a good representative analysis of the sea water that is to be processed in the plant.

Magnesium concentration can be of critical importance. Normal acid treating is generally sufficient to suppress magnesium hydroxide formation in the preheat train. However, the pH of the brine in the high temperature effects is somewhat above the pH of the brine in

(1) Molar Concentration Product - Molar Calcium Concentration
(Moles/Liter) X Molar Sulfate Concentration (Moles/Liter)

TABLE G-10-c

RESULTS OF MAGNESIUM ANALYSES

<u>Date</u>	<u>C. F.</u>	<u>Plant Feed Magnesium ppm</u>	<u>Adjusted to 1.0 Concentration Factor Magnesium ppm</u>
1-20-69	0.74	1,130	1,530
2-4-69	0.74	1,100	1,490
2-18-69	0.78	1,170	1,500
3-4-69	0.71	1,070	1,510
3-14-69	0.64	960	1,500
3-19-69	0.85	1,260	1,480
3-28-69	0.92	1,350	1,470
4-14-69	0.64	1,000	1,560
"Normal" Sea Water	1.0	---	1,270

the sea water feed out of the last preheater. If the magnesium content of the sea water is unusually high, magnesium hydroxide can form in the tubes of the higher temperature effects, if the system pH is not carefully controlled.

b. Chlorinity

Late in 1968, several attempts were made to evaluate evaporator performance from chlorosity data. It was shown that the standard chlorosity analytical technique lacked the precision required to evaluate evaporator performance on a salt balance basis. An analytical technique was developed that yielded weight percent halide values good to four significant figures; this is the degree of accuracy that is required to material balance the evaporators on a salt balance basis.

The main modification to the standard chlorosity technique is the use of a weighing buret to measure the amount of titrant added. Thus the sample and the titrant are both weighed. Additionally, by using a small plastic tip, the drop size has been reduced to about 1/80 of a millimeter, allowing a finer resolution of the end point.

Initial experiments showed that the delivery of titrant was so slow that the silver chloride precipitate was absorbing chloride ions from the titration media. It has been found necessary to titrate the chlorine to an apparent end point and then crush the precipitate with a glass rod, titrate further, and repeat the crushing.

At the present, it is felt that the precision of the chlorinity (wt % total halide) determination is limited by the selection of an end point to about 10-20 ppm.

c. Copper and Iron

- 1) Background - Work was continued during Development Runs 15 and 16 on monitoring the iron and copper content of the liquid streams leaving the plant. When corrosion occurs, the products of corrosion will either accumulate in the plant, or leave the plant dissolved or suspended in the reject brine and product streams.

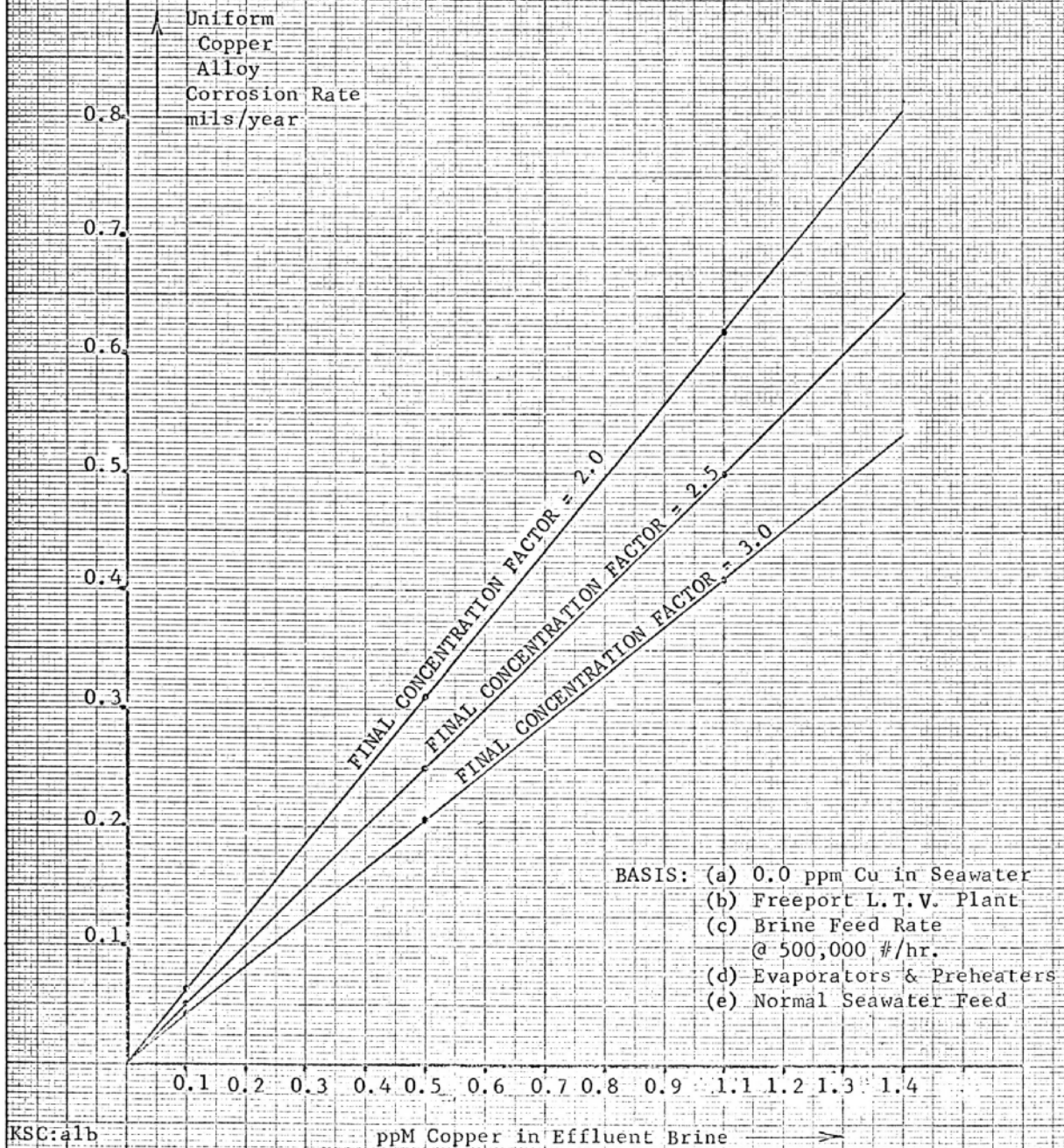
Knowing the physical dimensions and composition of the evaporator and preheater tubes, it is possible to calculate the equivalent copper surface in the plant that is exposed to brine. Assuming uniform corrosion, a copper alloy corrosion rate in mils per year can be calculated as a function of the copper content of the effluent brine; this has been done and plotted in Figure G-10-a. Note that this graph is good only for uniform copper dissolution and does not take into account corrosion products remaining in solid form in the evaporators; this approach will not pick up pitting attack.

FIGURE G-10-a

UNIFORM COPPER ALLOY CORROSION RATE

versus

COPPER CONTENT OF EFFLUENT BRINE



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This graph is specific for the Freeport Test Facility in its present configuration. It can be used to get a rough idea of corrosion rate from data on the copper concentration of effluent brine.

- 2) Development Run 15 - Composite samples of brines and condensates were collected on a weekly basis throughout Development Run 15 for copper and iron analysis. All of the iron analyses were run by a colorimetric procedure adapted from ASTM Standard Method D 1068. The copper analyses were run by a colorimetric procedure adapted from ASTM Standard Method D 1688. The iron and copper results from these analyses are shown on Tables G-10-d and e.

Inspection of the data shows that there are unexplained random changes in the iron and copper content of the brines. The concentration of both these cations in the condensate is very low.

It will be noticed that on two different days (5/7 and 7/8) more iron went into the plant than was seen in the blowdown of Effect 17 (when adjusted for evaporation of water in the plant). The iron content of the sea water is quite high and shows considerable fluctuation.

The ratio shown on Table G-10-e between iron content of the last effect blowdown and iron content of the blended brine shows that, in general, the iron content of the discard brine went up and down with the iron content of the feed brine. This ratio gives a rough idea of the rate of iron dissolution in the plant. If this ratio equals the plant extraction ratio, the apparent iron dissolution rate is quite low.

Ratios higher than the extraction ratio indicate increasing iron dissolution rates. If the ratio is lower than the extraction ratio, it means that iron is accumulating in the plant.

- 3) Development Run 16 - Composite samples were collected on a routine basis throughout Development Run 16 for copper and iron analysis. All of these analyses were run by atomic absorption; the results are shown in Tables G-10-f and g.

It should be noticed that the copper and iron concentrations in Effects 12 and 17 brine blowdown vary quite drastically from week to week. This same variation was noticed during Runs 14 and 15. To determine the cause, two sampling programs were carried out.

TABLE G-10-d

TOTAL INORGANIC COPPER ANALYSES RESULTS

DEVELOPMENT RUN 15 - ppm as Cu

<u>Date</u>	<u>Condensate ppm</u>	<u>Blended ppm</u>	<u>No. 12 B.D. ppm</u>	<u>No. 17 B.D. ppm</u>
5/7/68	0.01	0.09	0.42	1.12
5/20/68	0.05	0.07	0.33	0.60
5/27/68	0.00	0.14	0.45	0.58
6/3/68	0.04	0.14	0.80	0.96
6/10/68	0.00	0.06	0.68	0.82
6/25/68	0.04	0.44	1.40	4.56
7/1/68	0.03	0.07	0.54	3.00
7/8/68	0.04	0.05	0.50	1.92
7/22/68	0.02	0.02	0.33	0.44
7/29/68	0.03	0.04	0.64	0.80
8/12/68	0.02	0.02	0.22	1.62
8/27/68	0.04	0.12	1.00	3.44
9/3/68	0.08	0.04	1.00	1.76
9/9/68	0.01	0.02	0.12	0.50
9/19/68	0.03	0.02	0.12	0.09
9/20/68	0.74	0.76	6.20	10.70

TABLE G-10-e

TOTAL INORGANIC IRON ANALYSES RESULTS

DEVELOPMENT RUN 15 - ppm as Fe

<u>Date</u>	<u>Condensate ppm</u>	<u>Blended ppm</u>	<u>No. 12 Blowdown ppm</u>	<u>No. 17 Blowdown ppm</u>	<u>RATIO</u>
					<u>#17 Blowdown, ppm</u> <u>Blended Brine, ppm</u>
5/7/68	0.1	2.6	4.6	7.3	2.8
5/20/68	0.1	3.7	7.8	10.2	2.8
5/27/68	0.1	1.8	4.2	6.6	3.7
6/3/68	0.2	3.2	8.0	12.1	3.8
6/10/68	0.1	1.6	4.4	7.1	4.4
6/25/68	0.1	3.0	8.6	11.3	3.8
7/1/68	<0.1	5.5	12.3	24.4	4.4
7/8/68	<0.1	6.5	6.8	10.3	1.6
7/22/68	<0.1	1.4	3.4	4.9	3.5
7/29/68	<0.1	1.9	4.0	5.3	2.8
8/12/68	<0.1	0.8	2.1	2.9	3.6
8/27/68	<0.1	1.8	6.3	9.1	5.0
9/3/68	0.1	1.6	3.9	5.2	3.2
9/9/68	< 0.1	3.0	8.3	12.1	4.0
9/19/68	< 0.1	3.7	8.7	10.7	2.9
9/20/68	0.1	2.4	13.6	7.7	3.2

TABLE G-10-f

ROUTINE TOTAL INORGANIC COPPER ANALYSES

DEVELOPMENT RUN 16 - ppm as Cu

<u>Date</u>	<u>Cond.</u>	<u>Feed</u>	<u>12 BD</u>	<u>17 BD</u>
12-30-68	<0.1	<0.1	1.6	4.0
1-3-69	<0.1	0.1	3.6	3.0
1-8-69	<0.1	0.4	1.8	1.0
1-14-69	<0.1	<0.1	0.4	0.5
1-17-69	<0.1	<0.1	0.5	0.4
1-20-69	<0.1	<0.1	1.5	1.1
2-4-69	<0.1	<0.1	0.2	0.2
2-18-69	<0.1	<0.1	0.8	0.2
3-4-69	<0.1	<0.1	3.5	1.9
3-14-69	<0.1	<0.1	4.9	0.4
3-19-69	<0.1	<0.1	1.7	0.4
3-28-69	<0.1	<0.1	0.2	0.1
4-14-69	<0.1	<0.1	1.4	0.6
5-1-69	<0.1	<0.1	3.7	4.7

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TABLE G-10-g

ROUTINE TOTAL INORGANIC IRON ANALYSES

DEVELOPMENT RUN 16 - ppm as Fe

<u>Date</u>	<u>Cond.</u>	<u>Feed</u>	<u>12 BD</u>	<u>17 BD</u>
12-30-68	<0.1	0.5	2.5	3.4
1-3-69	0.1	0.8	5.2	6.8
1-8-69	0.1	0.6	2.4	3.5
1-14-69	<0.1	0.1	0.6	1.1
1-17-69	<0.1	0.1	0.6	1.1
1-20-69	0.1	0.7	3.0	4.1
2-4-69	<0.1	0.8	4.0	4.8
2-18-69	0.1	1.1	3.6	4.2
3-4-69	0.1	1.0	6.3	7.2
3-14-69	0.1	1.6	8.2	4.9
3-19-69	0.1	1.7	5.3	6.6
3-28-69	<0.1	0.9	5.8	4.7
4-14-69	0.1	1.9	5.0	6.9
5-1-69	0.1	2.8	9.3	14.1

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The first program was begun by collection of spot samples of plant feed and Effect 17 blowdown every 10 minutes for a period of 2 hours and every hour for a period of 6 hours. Special care was used during the sampling to insure that no copper or iron was picked up from sampling lines, etc. Plastic sample lines were used, along with stainless steel fittings, valves, etc., for both sampling points.

The results from the atomic absorption analysis of these samples are shown in Tables G-10-h and i. All of the iron results are for total iron. The samples taken at 10 minute intervals showed very little change in copper or iron throughout the two hour sampling period. The samples taken at one hour intervals, showed a slight, random fluctuation in copper and iron content. Notice that the copper and iron concentrations both increased quite sharply on the sample taken at 1300 hours. The reason for the sharp increase at this time is not known. The wide fluctuation seen in the weekly samples (Tables G-10-f and g) is not apparent.

On February 13, samples of plant feed and Effect 17 blowdown were composited for 30 minute intervals every half hour for a total of 4 hours. At the time these samples were taken, sodium sulfite was being injected into the deaerator effluent in order to reduce the dissolved oxygen to zero. Copper and iron analyses were run on these samples by atomic absorption. The results from these analyses are shown in Table G-10-j. Notice that on half-hour composite samples there is very little variation in copper content, but the fluctuation in iron content is still present.

In order to further investigate any "cycling", and in order to determine the effect of sulfite addition (zero dissolved oxygen) on copper and iron corrosion, another sampling program was initiated on February 25. Hour long composite samples were taken every hour for a 23-hour period. When the sampling was started, sodium sulfite was being injected into the deaerator effluent. Dissolved oxygen analyses showed no oxygen at this time. After collection of the samples at 2400, the sulfite injection was again started. The dissolved oxygen again fell to zero. Atomic absorption analyses of these samples gave the results shown in Table G-10-k.

The iron and the copper content of the Number 17 effect brine blowdown was quite constant for this 24-hour period. The extreme fluctuation in iron and copper that is seen in the weekly composite samples does not appear in these results. The copper content of the Number 17 effect brine blowdown is quite low; using Figure G-10-a, the apparent uniform copper alloy corrosion rate

TABLE G-10-h

IRON AND COPPER ANALYSES

10 MINUTE SPOT SAMPLES

1-23-69

<u>Time</u>	<u>Plant Feed</u>		<u>No. 17 Blowdown</u>	
	<u>ppm Fe</u>	<u>ppm Cu</u>	<u>ppm Fe</u>	<u>ppm Cu</u>
1420	0.6	<0.1	3.1	0.4
1430	0.6	<0.1	3.0	0.4
1440	0.6	<0.1	3.3	0.4
1450	0.6	<0.1	3.2	0.4
1500	0.7	<0.1	3.2	0.4
1510	0.8	<0.1	3.1	0.4
1520	0.8	<0.1	3.2	0.4
1530	0.8	<0.1	3.3	0.4
1540	0.9	<0.1	3.0	0.4
1550	0.9	<0.1	3.3	0.4
1600	0.9	<0.1	3.4	0.4
1610	0.7	<0.1	3.3	0.4
1620	0.7	<0.1	3.3	0.4

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TABLE G-10-i

IRON AND COPPER ANALYSES

1 HOUR SPOT SAMPLES

1-27-69

<u>Time</u>	<u>Plant Feed</u>		<u>No. 17 Blowdown</u>	
	<u>ppm Fe</u>	<u>ppm Cu</u>	<u>ppm Fe</u>	<u>ppm Cu</u>
0800	2.0	0.2	4.7	0.8
0900	1.4	0.1	5.0	0.6
1000	1.5	0.1	5.7	0.6
1100	1.2	0.1	5.5	0.5
1200	1.2	0.1	5.0	0.5
1300	1.1	0.1	9.9	1.6

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TABLE G-10-j

COPPER AND IRON ANALYSES

1/2 HOUR COMPOSITE SAMPLES

2-13-69

<u>Time</u>	<u>ppm Cu Feed</u>	<u>ppm Cu No. 17</u>	<u>ppm Fe Feed</u>	<u>ppm Fe No. 17</u>
1230	0.1	0.5	0.7	4.7
1300	0.1	0.4	0.7	4.7
1330	0.1	0.4	0.8	4.9
1400	0.1	0.4	0.8	4.7
1430	0.1	0.4	1.0	4.5
1500	0.1	0.4	1.2	4.8
1530	0.1	0.4	1.4	5.0
1600	0.1	0.4	1.4	5.2

PW:alb
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TABLE G-10-k

IRON AND COPPER ANALYSES

1 HOUR COMPOSITE SAMPLES

2-25/26-69

<u>Time</u>	<u>ppm Fe</u>		<u>ppm Cu</u>	
	<u>Feed</u>	<u>No. 17</u>	<u>Feed</u>	<u>No. 17</u>
1600	1.5	5.0	0.08	0.41
1700	1.2	5.3	0.04	0.32
1800	1.1	5.3	0.04	0.55
1900	1.1	4.7	0.05	0.45
2000	1.1	5.0	0.04	0.43
2100	1.0	4.4	0.04	0.29
2200	0.9	4.4	0.04	0.36
2300	0.9	4.5	0.04	0.58
2400	0.9	4.5	0.04	0.39
Sulfite Addition Off				
0100	1.1	4.2	0.04	0.36
0200	1.0	4.3	0.05	0.41
0300	0.9	4.1	0.05	0.37
0400	0.9	4.2	0.04	0.39
0500	1.0	4.5	0.04	0.44
0600	1.1	4.6	0.04	0.42
0700	1.2	5.0	0.04	0.36
0800	1.3	5.4	0.05	0.34
0900	1.0	5.5	0.04	0.37
1000	1.1	5.5	0.04	0.36
Sulfite Addition Back On				
1100	1.0	6.0	0.04	0.38
1200	1.1	5.7	0.04	0.36
1300	1.0	5.4	0.04	0.35
1400	0.9	5.3	0.05	0.33

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is about 0.1 mils per year. There was no significant change in either iron or copper concentration of Number 17 effect blowdown caused by the 10-hour termination of sulfite addition.

On March 4th, through March 13th, the final special sampling program was carried out which consisted of collecting 8-hour composite samples of plant feed and Effect 17 blowdown. The sampling was started right after a plant start-up. At this time sodium sulfite was being injected into the deaerator effluent to keep the dissolved oxygen content at a very low level. On March 7, 1969, at 1600, the sulfite injection was stopped--sampling was continued.

Copper and iron analyses were run on these samples by atomic absorption. The results from these analyses are shown in Table G-10-1. The results in this table show a random variation in copper and iron from one sampling period to the next. There is, however, a definite trend shown--the copper and iron content of the plant blowdown starts out relatively high, drops rapidly after start-up, and shows a slow rise when the sulfite injection is shut off.

On the basis of the Development Run 16 work, the following conclusions can be drawn.

- a) The concentration of iron and copper in the effluent brine starts high at start up and then drops sharply to a much lower equilibrium value.
- b) The average copper concentration in the effluent brine is equivalent to a copper alloy uniform corrosion rate of 0.1 to 0.2 mils per year.
- c) Sulfite addition does not seem to greatly change the concentration of copper or iron in the effluent brine.
- d) The copper and iron pickup on the condensate side is extremely low.

d. Calibration of Hand Refractometer (Brix Meter)

The calibration of the hand refractometer was checked. Brix readings were made on samples from two different heat and material balances from Run 15. Accurate chlorinity determinations had already been made on these samples. The results were used to calculate concentration factors and the resulting factors plotted against Brix readings. Figure G-10-b shows the results.

TABLE G-10-1

IRON AND COPPER

8 HOUR COMPOSITES

<u>Date</u>	<u>Time</u>	<u>ppm Cu Feed</u>	<u>ppm Cu No. 17</u>	<u>ppm Fe Feed</u>	<u>ppm Fe No. 17</u>	
3-4-69	0800	0.14	1.5	1.2	7.2	
	1600	0.07	1.4	1.3	7.4	
	2400	0.07	0.5	1.0	5.2	
3-5-69	0800	0.09	0.4	1.0	4.6	
	1600	0.07	0.4	1.3	6.7	
	2400	0.09	0.4	1.1	5.0	
3-6-69	0800	0.07	0.4	0.9	4.6	
	1600	0.09	0.6	1.3	5.9	
	2400	0.06	0.4	1.0	5.3	
3-7-69	0800	0.06	0.4	0.9	4.9	
						Stop
						Sulfite
						Addition
	1600	0.05	0.4	0.9	4.8	
	2400	0.06	0.5	0.9	4.7	
3-8-69	0800	0.15	0.5	1.0	4.9	
	1600	0.06	0.5	1.0	5.7	
	2400	0.13	0.5	1.4	5.3	
3-9-69	0800	0.05	0.5	1.4	5.3	
	1600	0.17	0.6	1.3	5.6	
	2400	0.08	0.5	0.9	4.6	
3-10-69	0800	0.05	0.6	1.2	4.8	
	1600	0.07	0.8	1.6	6.3	
	2400	0.07	0.6	1.2	6.1	
3-11-69	0800	0.05	0.6	1.2	5.3	
	1600	0.05	1.5	1.4	7.1	
	2400	0.04	0.7	1.4	6.7	
3-12-69	0800	0.04	0.6	1.0	5.0	
	1600	0.04	0.6	1.1	5.3	
	2400	0.04	0.6	1.1	5.9	
3-13-69	0800	0.04	0.6	0.9	4.7	
	1600	0.04	0.6	1.4	6.6	
	2400	0.04	0.5	0.8	4.9	

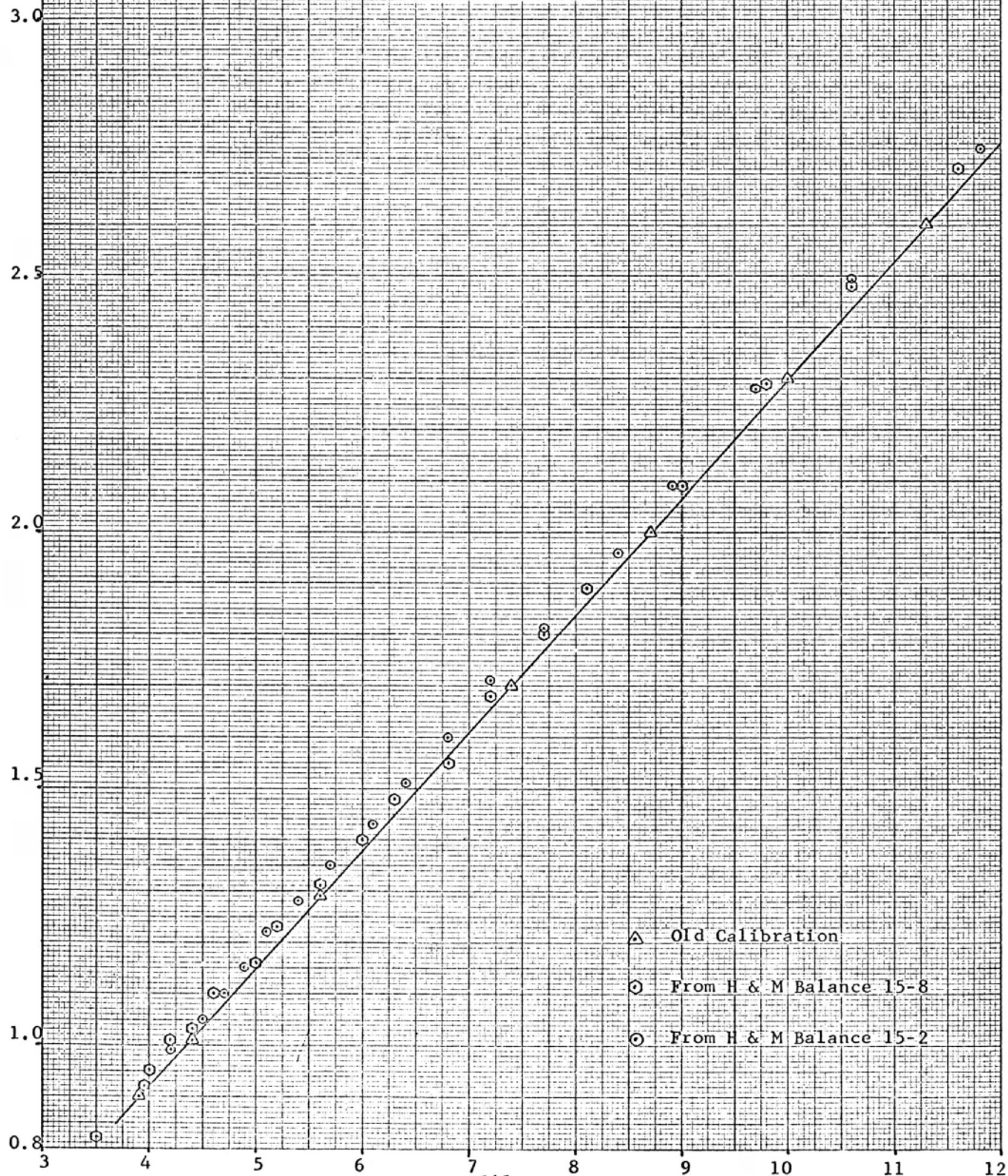
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FIGURE G-10-b

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CONCENTRATION FACTOR vs. BRIX°

DEVELOPMENT RUN 15



△ Old Calibration

⊙ From H & M Balance 15-8

○ From H & M Balance 15-2

Inspection of the data shows that the results from both sets of samples are quite close together with only a small amount of scatter of the points; both of the curves are quite close to the standard calibration curve. The slight scatter is probably caused by inability to read the hard refractometer scale to closer than one unit.

e. Deposit Analyses

The approximate analysis of several samples that were taken during the shutdown after Run 15 are shown on Table G-10-m. It should be noted that these are only approximate analyses which serve to characterize the sample. In the procedure that is used in this type of analysis, a whole group of related compounds is analyzed as one. For instance, Fe_2O_3 , Al_2O_3 , MnO_2 , etc. are all precipitated as the hydroxides and ignited back to the oxide form. The resulting possible mixture of oxides is referred to as R_2O_3 and in all cases is called Fe_2O_3 since this is the most likely of these compounds to be found in the samples here at Freeport. X-ray diffraction or differential thermal analysis would be required to get any more exact data regarding these samples.

There are few surprises in the results. Most of these samples were scraped from alloy surfaces which are exposed to hot brine. The samples are predominately copper oxide; the deposits most likely result from the reactions that occur when the alloy becomes the cathode in a galvanic couple formed with near-by carbon steel. The presence of significant metallic copper in these samples is further evidence of electrochemical reaction.

The fact that these samples gain weight on ignition is probably further substantiation of the presence of significant amounts of metallic copper. Oxidation of metallic copper during ignition would cause the sample to gain weight, as would a change in oxidation state from cuprous to cupric.

The origin and significance of the silica content of these samples is not known. Visual observation did not reveal any significant silt entrapment by these deposits.

A sample scraped off of the column pipe of P-5(a) was also analyzed. It was found to be primarily organic material. It is surmised that this is the residual from an organic protective coating that was applied to the outside of this pump column some years ago.

f. Atomic Absorption Standardization

Analytical procedures for use on the atomic absorption unit have been set up and checked out for copper, iron and calcium:

TABLE G-10-mDEPOSIT ANALYSIS RESULTSDEVELOPMENT RUN 15

<u>Sample Point(1)</u> <u>and Description</u>	<u>Approximate Analysis, %</u>				
	<u>Gain/Loss</u> <u>on Ignition</u>	<u>Silica</u> <u>as SiO₂</u>	<u>Iron</u> <u>as Fe₂O₃</u>	<u>Copper(2)</u> <u>as CuO</u>	<u>Calcium</u> <u>as CaSO₄</u>
<u>Effect No. 5:</u> Hard deposit from top water box adhering to alloy liner close to carbon steel tube sheet.	-1	4	20	75	4
<u>Evaporator No. 9:</u> Deposit adhering above "water line" to upper water box outlet alloy riser.	5	6	None	87	None
<u>HX-302a:</u> White/blue/gray deposit w/brown underneath; adhering to alloy inlet pass baffle.	12	3	None	99	None
<u>HX-303b:</u> White/blue/gray deposit adhering to alloy inlet pass baffle.	13	4	None	100	None
<u>HX-308:</u> Brown deposit from north side of alloy water box.	10	9	None	96	None
<u>HX-308:</u> Surface of alloy inlet baffle & south side of water box.	11	6	None	100	None

- (1) All samples taken on 10/1/68 by B. P. Webb and K. S. Campbell.
 (2) Metallic copper found to be present in all cases.

1) Copper Analysis

The atomic absorption instrument was checked in the 0 to 5 ppm range and found to be perfectly linear with respect to copper.

Actual plant samples were analyzed by atomic absorption and by the colorimetric method. The results are shown in Table G-10-n. Inspection of these results shows that quite close agreement was obtained by the two different methods, the difference being about 0.1 to 0.2 ppm. This is within the basic accuracy of both techniques.

The atomic absorption method cannot be matched as far as speed is concerned. Twenty samples by the colorimetric method would take one to two days for analysis; by atomic absorption they can easily be run in 15 minutes.

2) Iron Analysis

The atomic absorption iron analysis was set up and checked in the range 0 to 10 ppm. The instrument was quite linear in this range.

Actual plant samples were run by the colorimetric method and by the atomic absorption method. A new colorimeter curve was prepared using the same standards that were used to calibrate the atomic absorption unit. The results from the analyses by these two methods are shown in Table G-10-o. It is to be noted that these results are for ionic iron.

Inspection of these results shows about a 0.1 to 0.4 ppm difference between the results from the two methods. It is felt that the results by atomic absorption are more accurate than by the colorimetric method since, in the colorimetric method for iron, the colored complex formed is not a strong absorber of light and thus makes the analysis tenuous.

3) Calcium Analysis

A different problem was encountered for the atomic absorption analysis of calcium. The concentration of calcium in the samples was so high that it was impossible to run the samples without prior dilution. The atomic absorption method is so sensitive for calcium (this is always the case for this element) that any concentration over 25 ppm drives the instrument "off scale". It is reported in the literature that a less sensitive wave length can be used for calcium up to 1000 ppm. The instrument turned out to be quite nonlinear in this range.

TABLE G-10-n

COPPER ANALYSIS COMPARISON

ATOMIC ABSORPTION vs. COLORIMETRY

<u>Date</u>	<u>Sample</u>	<u>ppm Cu Colorimeter</u>	<u>ppm Cu Atomic Absorption</u>
12-30-68	Cond.	0.1	<0.1
12-30-68	Feed	<0.1	<0.1
12-30-68	No. 12	1.4	1.6
12-30-68	No. 17	4.4	4.0
1-3-69	Cond.	<0.1	<0.1
1-3-69	Feed	0.1	0.1
1-3-69	No. 12	3.5	3.6
1-3-69	No. 17	3.1	3.0
1-8-69	Cond.	<0.1	<0.1
1-8-69	Feed	0.3	0.4
1-8-69	No. 12	1.6	1.8
1-8-69	No. 17	0.8	1.0
1-14-69	Cond.	<0.1	<0.1
1-14-69	Feed	<0.1	<0.1
1-14-69	No. 12	0.4	0.4
1-14-69	No. 17	0.3	0.5
1-17-69	Cond.	<0.1	<0.1
1-17-69	Feed	<0.1	<0.1
1-17-69	No. 12	0.3	0.5
1-17-69	No. 17	0.2	0.5

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TABLE G-10-o

IRON ANALYSIS COMPARISON

ATOMIC ABSORPTION vs. COLORIMETRY

<u>Date</u>	<u>Sample</u>	<u>ppm Fe Colorimetry</u>	<u>ppm Fe Atomic Absorption</u>
12-30-68	Cond.	0.1	<0.1
12-30-68	Feed	0.4	0.5
12-30-68	No. 12	2.3	2.5
12-30-68	No. 17	3.3	3.4
1-3-69	Cond.	0.1	0.1
1-3-69	Feed	1.0	0.8
1-3-69	No. 12	5.6	5.2
1-3-69	No. 17	7.1	6.8
1-8-69	Cond.	<0.1	0.1
1-8-69	Feed	0.4	0.6
1-8-69	No. 12	2.0	2.4
1-8-69	No. 17	3.3	3.5
1-14-69	Cond.	<0.1	<0.1
1-14-69	Feed	<0.1	0.1
1-14-69	No. 12	0.1	0.6
1-14-69	No. 17	0.3	1.1
1-17-69	Cond.	<0.1	<0.1
1-17-69	Feed	<0.1	0.1
1-17-69	No. 12	0.2	0.6
1-17-69	No. 17	0.3	1.1

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In order to run blended and raw sea water samples for calcium content it is necessary to dilute the sample to the point where the calcium concentration falls in the 0 to 25 ppm range. A simple 1:20 dilution with deionized water accomplishes this, and also decreases the sodium by a factor of 20.

The linearity of the instrument was checked in this 0 to 25 ppm range; it was found to be quite linear. The analysis on diluted samples was very "quiet" and easy to run.

Samples of plant feed were run by the atomic absorption method and by the titrimetric method. The results are shown in Table G-10-p. Quite good agreement was found for the two methods, the average difference being 8 ppm. Probably the titrimetric method is totally at fault here. The endpoint in this method can be reproduced to ± 0.1 ml, which corresponds to $80.1 \times 0.1 = 8$ ppm Ca^{++} .

TABLE G-10-p

CALCIUM ANALYSIS COMPARISON

TITRATION vs. ATOMIC ABSORPTION

PLANT FEED SAMPLES

<u>Date</u>	<u>ppm Ca⁺⁺ Titration</u>	<u>ppm Ca⁺⁺ Atomic Absorption</u>
12-30-68	320	340
1-3-69	310	325
1-8-69	325	330
1-14-69	320	320
1-17-69	325	330
1-20-69	325	320

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G. RESULTS AND COMPARISONS (CONTD)

11. MATERIALS PERFORMANCE REPORT SYSTEM

In 1967 a Materials Evaluation and Reporting System (MERS) was instituted by the Office of Saline Water to be used by its various facilities to report to a central location the degree of success each contractor had with the application of the wide variety of materials being tested and/or used.

After the important details of report form, system dictionary and information retrieval were worked out, MERFS (Material Evaluation Report Forms) were filled out at the various facilities and transmitted to the central files. As of June, 1968, the MERF catalog contained 759 reports; 422 of these, or 55%, were from the Freeport Plant.

Late in 1968 the Materials Performance Report System (MPRS) replaced the MERS and absorbed all the information already submitted. The idea of the new system is similar with the old, the major differences being the method of information storage and style of report form.

During Development Run 15 and 16 and the modification periods following each, equipment material failures were recorded on a special materials evaluation short form. This form was created to allow various people to report failures simply but to have all reports to flow through the Development Engineer, who in turn would combine the significant operating parameters on a MPRS Form.

At the completion of the 17 Effect Plant Operation, 208 MPRS Forms were completed and transmitted to the Materials Information Center (MIC) at Oak Ridge National Laboratory. This will make the Freeport Plant inspection completion date September 30, 1969.

In this latest inspection report transmittal, the following equipment categories are covered:

<u>Catagory</u>	<u>Number of Reports</u>
1. Piping	118
2. Pumps	25
3. Valves	23
4. Heat Exchangers	13
5. Evaporators	13
6. Flash Tanks	9
7. Deaerator	2

<u>Catagory</u>	<u>Number of Reports</u>
8. Acid Tanks	2
9. Caustic Tanks	2
10. Equipment Housing	1

Failures in the following types of materials were noted:

<u>Material</u>	<u>Number of Reports</u>
1. Carbon Steel	127
2. Glass Reinforced Epoxy	14
3. 316 Stainless Steel	13
4. Phenolic	10
5. Carbon	8
6. Buna-N	8
7. Epoxy	6
8. Cast Iron	6
9. Glass Reinforced Polyester	4
10. Bronze	4
11. Monel	4
12. Nordel Rubber	4
13. Glass	3
14. Admiralty, (As.)	2
15. Asbestos	2
16. Ceramic	2
17. Cement	2
18. Brass	2
19. Ni-Resist	1
20. Neoprene Rubber	1

Some of the more interesting material failures and inspections are compiled below:

- a. Failure of the carbon steel pipe spools at reducers and elbows in the high-temperature brine suction and discharge lines continued to be an operating and maintenance problem.
- b. During the Development Run 14-15 shutdown, glass reinforced epoxy pipe was put in hot brine transfer service. Early in Development Run 15, there were some failures of the glass reinforced epoxy components primarily due to expansion problems or inadequate fabrication.
- c. In June, 1968, two of the 2" welded 90/10 Cupro/Nickel tubes were pulled out of the spare stock evaporator 13 bundle and shipped out for evaluation by the vendor. There was no measurable corrosion or preferential attack in the weld area. The tubes had only been in operation about 130 days.

- d. A 45° fibercast plastic pipe elbow at the discharge of P-13 showed signs of serious "checkered" attack. In many places the outside clear layer had been almost completely eroded away down to the fiber level.
- e. The lining on the P-16 discharge control valve had been torn partly away. With the lining gone, the valve seat also showed considerable erosion and corrosion. The brine piping downstream of this control valve was coated carbon steel (baked on phenolic). This coating had failed in several places. There were several holes in the first flange and one half-inch deep hole about the size of a half dollar just above the weld in the reducer. The coating in the straight run piping appeared to be in good condition.
- f. The 4" recycle line to No. 15 Effect was temporarily replaced with carbon steel. The recycle line that was taken out, was inspected, and found to be in bad shape. This line had been internally coated with phenolic plastic, but the coating was completely washed out downstream of the welds that were made to field fit the pipe.
- g. The condensate drain line from Heat Exchangers 302a and 302b that was replaced was hammer tested and found to be in fairly good shape, except where the leak had developed at one of the elbows. Similar leaks were also found in the 45° elbows on the condensate drains of HX-305, 306, 307, 309, 310, and 311.
- h. A number of holes were found in the steam chest of No. 2 Effect just above where the monel expansion joint is welded to the carbon steel evaporator steam chest.
- i. P-5(a) was inspected. The shaft was badly worn (approximately 30% reduction in dia.) at the packing gland. There was some pitting of the pump bowls and impellers. As was noticed at the last inspection of this pump, the outside of the column pipe was heavily pitted. All parts are of 316 stainless steel.
- j. The plastic SVL distribution weirs in the higher temperature effects (Effect 9) show signs of shrinking and general deterioration. In some of the other effects, the unrestrained weirs had a tendency to pop out of the tubes.
- k. The polyester corite distribution plates in the first five effects (high temperature) all showed signs of disintegration. They were replaced with 316 stainless steel plates of similar design.
- l. The Brutem 110 Epoxy deaerator coating was in pretty good shape with some scattered rust spots. The failed areas were repaired with the same epoxy.

- m. After more than 8,500 operating hours, the cement-mortar lined brine sumps of Effects 8 and 9 appear to be in good condition.
- n. A fibercast 6" x 8" polyester reducer at the discharge of P-20 was removed when very serious wear due to erosion was discovered.
- o. The bronze impeller on the deaerator feed pump (P-3) showed signs of impingement wear on both the inlet and outlet vanes. The wear took the form of many small depressions in the bronze similar to cavitation attack but not as deep.
- p. The tube ends in the top tube sheet of HX-213 showed signs of considerable wear. This is the first heat exchanger that the blended, aerated sea water contacts. The admiralty tubes are rolled into a tube sheet lined with 70-30 Cu-Ni.
- q. Superficial scraping with a pick of the 70-30 Cu-Ni liners in the top water box of HX-304 and 305 revealed a soft layer of coppery material on the surface. Hard scraping revealed the sub-surface of whitish 70-30 Cu-Ni. It was thought that the outer layer of copper was due to "parting" corrosion which leads to denickelification. Theoretically, as the result of galvanic action between the metal crystals, nickel leaves the alloy and goes into solution as an ion. It is also possible that this external layer of copper could be due to plating out of copper ions from the brine solution. It was felt that if this were the case, the external layer of copper would have been a great deal softer.
- r. The carbon steel next to the monel expansion joint on the suction of Hot Brine Pump P-11 failed, presumably as a result of galvanic corrosion.
- s. A new shop-fabricated Bondstrand Series 400 fiberglass reinforced plastic pipe elbow at the discharge of Hot Brine Pump P-15 failed at the weld. The failure might have been caused when the P-15 discharge spool was subjected to severe brine surging due to erratic action of the level control valve.

The areas where non-metallic material is used in the process plant at Freeport Test Facility are listed in Table 11-a.

There are four vessels in the plant where the testing of non-metallic coating protection of carbon steel waterboxes are in progress. Four materials are under test in real desalting applications. All materials, except one, lumnite cement, have failed in some way. Table 11-b summarizes the performance of these materials.

TABLE G-11-a

NON-METALLIC MATERIAL AT THE FREEPORT TEST FACILITY

<u>Item</u>	<u>Material Type</u>	<u>Vendor</u>	<u>Application</u>	<u>Service</u>	<u>Remarks</u>
1.	Glass Reinforced Polyester	Chemical Proof Corp.	16" Pipe	Cold, Raw Seawater	Failed once due to water hammer; OK since then.
2.	Glass Reinforced Polyester	Chemical Proof Corp.	Acid Mixing Chamber	90-120°F Acidified Seawater	OK
3.	Glass Reinforced Polyester	Dow-Smith Corp.	8" Pipe	Deaerated Seawater	Flange to pipe weld failure after approx. 2 years service.
4.	Glass Reinforced Epoxy	Amercoat Bondstrand Series 4000	8" Pipe	267°F Deaerated Brine	6 months service; still OK
5.	Glass Reinforced Epoxy	Amercoat Bondstrand Series 2000	8" Pipe	258°F Deaerated Brine	Flange to pipe weld failure after 4 months service; replaced in kind.
6.	Glass Reinforced Epoxy	Amercoat Bondstrand Series 2000	8" Pipe	258°F Deaerated Brine	6" x 8" reducer flange weld failure after 30 days replaced in kind.
7.	Glass Reinforced Epoxy	Amercoat Bondstrand Series 2000	8" Pipe	258°F Deaerated Brine	45° Ell flange weld failure after 30 days. Replaced in kind.
8.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	249°F Deaerated Brine	45° Ell flange weld failure after 30 days. Replaced in kind.
9.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	249°F Deaerated Brine	45° Ell flange weld failure after 30 days. Replaced in kind.
10.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	249°F Deaerated Brine	45° Ell flange weld failure after 60 days. Replaced in kind.
11.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	249°F Deaerated Brine	45° Ell flange weld failure after 60 days. Replaced in kind.

TABLE G-11-a (CONTD)

NON-METALLIC MATERIAL AT THE FREEPORT TEST FACILITY

<u>Item</u>	<u>Material Type</u>	<u>Vendor</u>	<u>Application</u>	<u>Service</u>	<u>Remarks</u>
12.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	249°F Deaerated Brine	6" x 8" reducer flange weld failure after 30 days. Replaced in kind.
13.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	249°F Deaerated Brine	6" x 8" reducer flange weld failure after 90 days. Replaced in kind.
14.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	249°F Deaerated Brine	45° Ell flange weld failure w/internal erosion after 5 months. Replaced in kind.
15.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	249°F Deaerated Brine	6" x 8" reducer flange weld failure after 30 days.
16.	Glass Reinforced Epoxy	Amercoat Bondstrand Series 2000	8" Pipe	231°F Deaerated Brine	5 months service small cracks showing up around flange weld.
17.	Glass Reinforced Epoxy	Amercoat Bondstrand Series 2000	8" Pipe	223°F Deaerated Brine	5 months service. Still OK.
18.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	223°F Deaerated Brine	Pipe saddle weld failure, repaired and reinstalled. 5 months total service. Still OK.
19.	Glass Reinforced Epoxy	Fibercast RB 2530	8" Pipe	223°F Deaerated Brine	Pipe flange weld failure. Repaired and reinstalled. 5 months total service. Still OK.
20.	Glass Reinforced Epoxy	Fibercast OG 2025	8" Pipe	213°F Deaerated Brine	10 months service. Showing cracks at flange welds. Still OK.
21.	Glass Reinforced Epoxy	Fibercast OG 2025	8" Pipe	204°F Deaerated Brine	10 months service. Still OK.

TABLE G-11-a (CONTD)

NON-METALLIC MATERIAL AT THE FREEPORT TEST FACILITY

<u>Item</u>	<u>Material Type</u>	<u>Vendor</u>	<u>Application</u>	<u>Service</u>	<u>Remarks</u>
22.	Glass Reinforced Epoxy	Fibercast OG 2025	8" Pipe	204°F Deaerated Brine	5 months service. Still OK.
23.	Glass Reinforced Polyester	Chemical Proof Corp.	Evaporator Dome	193°F Flashing Deaerated Brine	10 months service Still OK.
24.	Glass Reinforced Epoxy	Fibercast OG 2025	8" Pipe	193°F Deaerated Brine	10 months service. Still OK.
25.	Glass Reinforced Polyester	Chemical Proof Corp.	Flash Tank	193°F Flashing Condensate	5 months service. Still OK.
26.	Glass Reinforced Polyester	Fibercast CL 2025	8" Pipe	182°F Deaerated Brine	6" x 8" reducer erosion failure after 6 months. Rotated 90° and left in service.
27.	Glass Reinforced Polyester	Chemical Proof Corp.	Evaporator Dome	155°F Deaerated Brine	14 1/2 months service. Still OK.
28.	Glass Reinforced Polyester	Chemical Proof Corp.	Evaporator Dome	143°F Deaerated Brine	14 1/2 months service. Still OK.
29.	Glass Reinforced Polyester	Chemical Proof Corp.	Evaporator Dome	131°F Deaerated Flashing Brine	14 1/2 months service. Still OK.
30.	Glass Reinforced Polyester	Chemical Proof Corp.	Evaporator Dome	116°F Flashing Deaerated Brine	14 1/2 months service. Still OK.
31.	Glass Reinforced Polyester	Chemical Proof Corp.	Evaporator Dome	100°F Flashing Deaerated Brine	14 1/2 months service. Still OK.

TABLE G-11-a (CONTD)

NON-METALLIC MATERIAL AT THE FREEPORT TEST FACILITY

<u>Item</u>	<u>Material Type</u>	<u>Vendor</u>	<u>Application</u>	<u>Service</u>	<u>Remarks</u>
32.	Glass Reinforced Epoxy	Fiberblast Corp.	8" and 10" Pipe	155°F Deaerated Brine	14 1/2 months service. Still OK.
33.	Glass Reinforced Epoxy	Fiberblast Corp.	4", 8" and 10" Pipe	100°F Deaerated Brine	14 1/2 months service. Still OK.
34.	Glass Reinforced Epoxy	Fiberblast OG 2025	4" Pipe	143°F Deaerated Brine	5 months service. Still OK.
35.	Glass Reinforced Polyester	Corite-Reynolds Corp.	Evaporator Distributor Sheets	Flashing Deaerated Brine at Various tempera- tures	3 failed by cracking after 17 months service; 2 failed by cracking after 13 months service; 3 and 2 are still in service after 18 and 14 1/2 months, respectively.
36.	Polyvinyl Dichloride Plastic	Cabot Corp.	3" Pipe	Sulfuric Acid Tank Vent	6 months service. Still OK.
37.	Polyvinyl Dichloride Plastic	Cabot Corp.	2" Pipe	Acid, Caus- tic, and sodium sulfite tank drains	6 months service Still OK.

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TABLE G-11-b

NON-METALLIC COATING SUMMARY

Coating	Operating Hours	Environment			Vessel Location and Application	Remarks
		Temp.	Press.	Fluid		
RTV-112 (G.E.) Silicone Rubber	3,100	266°F	50 psig	Deaerated Brine	ORNL-VTE Test noz- zle tube sheet.	Failed. Became soft and tore.
RTV-112 (G.E.) Silicone Rubber	6,200	210°F	35 psig	Deaerated Brine	HX 307 Inlet tube sheet.	Pulled loose and tore.
RTV-112 (G.E.) Silicone Rubber	3,100	195°F	Vacuum	Deaerated Brine	No. 9 Tubesheet and flange.	Tore
Epoxy-Epon No. 828 (Shell)	6,200	266°F	50 psig	Deaerated Brine	ORNL-VTE Test noz- ube tube sheet.	Cracked at edges.
Lumnite Cement	6,200	266°F	50 psig	Deaerated Brine	ORNL-VTE Test noz- zle tube sheet.	Still good.
Gaco EHS-55 (Gates Eng)	9,100	85°F	5 psig	Raw Sea Water	HX 318 inlet, outlet and return water boxes and baffles	Chipped at edges of drain holes and walltube sheet and wall- baffle joints.

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G. RESULTS AND COMPARISONS (CONTD)

12. CORRATER

During Development Run 16, a Magna Corporation Model 1180, 6-Point corrater, together with a Model 1190 programmer, was added to the plant control room instrumentation. The purpose of this installation was to determine the reliability of the instrumentation package as an operating tool. Trial evaluation of the unit was still underway at the end of the run. The test objectives were to be accomplished, following the initial de-bugging period, by demonstrating the corrater's usefulness in pointing to trouble spots in the operating system. Data would have been taken when a known process upset was caused and the corrater output would be noticed. It was planned to accumulate, in book form, a series of upset vs. corrater traces, to be used as a trouble-shooter's guide to the process plant.

The six locations of the process probes follow:

<u>Location</u>	<u>Environment</u>	<u>Temperature °F (Approx.)</u>
1. Heat Rejection Condenser Inlet Cooling Water	Raw Sea Water	Ambient
2. Deaerator Brine Effluent	Deaerated Brine	100
3. No. 1 Effect Blowdown	Deaerated Brine	266
4. No. 8 Effect Blowdown	Deaerated Brine	200
5. No. 13 Dome	Deaerated Brine	160
6. No. 17 Sump	Deaerated Brine	100

G. RESULTS AND COMPARISONS (CONTD)

13. MISCELLANEOUS STUDIES

a. Steam Jet Ejectors

The steam jet ejectors that were installed in the Freeport Test Facility in the late spring of 1966 have been a continuing operating problem.

Since the steam jet ejectors were installed, considerable time and effort have been expended trying to find the exact cause of the problem. No easy analytical solution has been found. Regarding the jets, the following can be said:

- 1) The existing steam jet ejectors were designed to operate with saturated steam. Most of the time there is superheated steam at the Freeport Test Facility. The jet nozzles have been bored out to compensate for the difference in specific volume between the saturated and superheated steams. The manufacturer indicated that occasionally jets designed to operate on saturated steam did not work too well with superheated steam even after the nozzles had been bored out.

Other manufacturers report difficulties in using superheated steam in saturated steam ejectors because the increase in specific volume tends to choke the diffuser.

It is not possible to desuperheat the jet steam with existing equipment. The only system that is at all compatible is the ATU desuperheater which is run at pressures significantly below that required by the steam jet ejectors. Any new system to desuperheat only the ejector steam would be an additional potential operating obstacle to be avoided at almost all costs.

- 2) The steam jet failures are totally random and unexplained. Enough data has been taken to assure us that 99% of the time the ejectors work on their nominal design curve. No correlation between the failures and any other process or environmental variables has been found.

The source of the trouble has been traced to the second stage. The trouble starts when there is, for no apparent reason, a decay in the inter-stage pressure. This may happen an hour, a day, or even a week after the plant has been started up and is lined out.

During the first part of Development Run 15, the steam jets performed satisfactorily. In the middle of the Run, there were repeated random failures. It was found that the failures could be minimized if the outside of the second stage throat chamber and diffuser were kept cool; a cooling water jacket for this purpose was installed. Spraying water on the throat chamber and second stage diffuser and allowing the excess to drain off through the cooling water jacket improved the reliability of the steam jet ejectors for the balance of Development Run 15.

During the shutdown between Run 15 and 16, two additional changes to the steam jet system were made. First, the second stage nozzle throat was drilled out to one larger drill size. This was done to determine if the additional quantity of steam passing through the nozzle and diffuser would help to steady the second stage down.

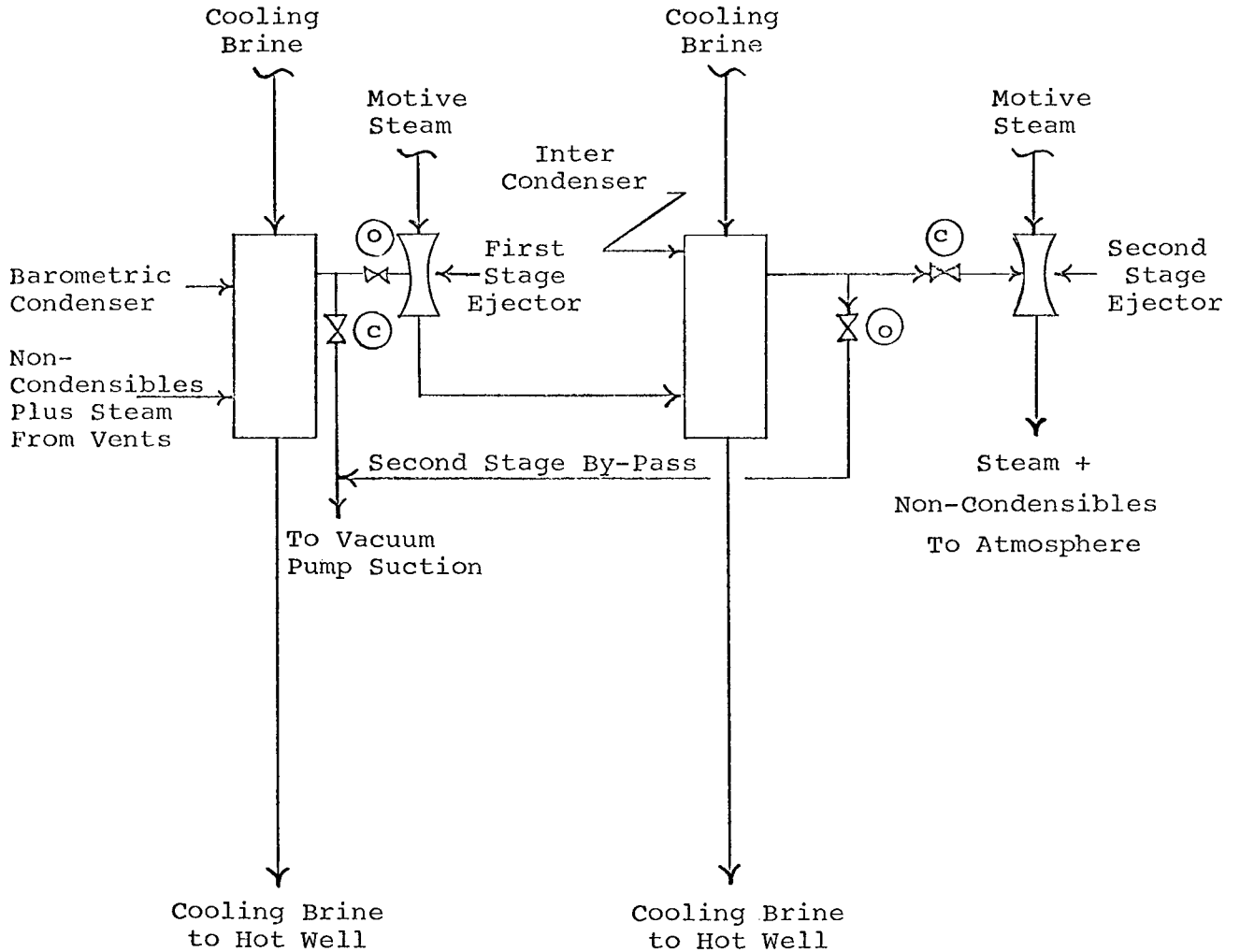
Additionally, the by-pass shown in Figure G-13-A was put in so that if the second stage failed, the first stage jet and the vacuum pump could be operated in series. This provided two alternate, separable systems for pumping non-condensibles up from the back-pressure achieved by the temperature of the water in the hot well.

At the start of Run 16, the second stage jet still would not pump. At this point, the new first stage jet-vacuum pump combination was tried out. Operations in this mode were extremely smooth. Absolute pressures as low as 0.7 inches of mercury were observed at the barometric precondenser. According to the manufacturer's performance chart, this is equivalent to about 120 lbs./hr. of non-condensibles and water vapor.

To determine the source of the non-condensibles, several vents around the plant were shut off one at a time. There was no noticeable change when the vents from the 12-effect plant were shut off. There was a decrease in barometric

FIGURE G-13-a

Process Flow - Steam Jet By-Pass



NOTE: Letters refer to valve position when 2nd stage ejector is off.
 (C) = Closed
 (O) = Open

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suction pressure equivalent to a reduction of about 30-40 lbs./hr. of non-condensibles when the vent from the heat rejection condenser was shut off.

The combination vacuum pump and first stage jet operation has afforded the plant stable operations and excellent non-condensable pumping capacity. This system remained in operation for over 3,100 operating hours, all Run 16, and experienced no failures.

The cause of all the trouble is probably the second stage diffuser. It is obviously badly corroded and eroded. It has been thermally shocked and wire brushed several times. It has been recommended that this carbon steel diffuser be replaced with a new stainless steel diffuser of the same size. This would give the plant two competent vacuum producing systems.

b. Plant Response Test

Toward the end of Run 15, a plant response test was performed. For testing purposes the operations of the plant were changed as follows:

- 1) Forward Feed Brine Pump P-18 was turned off for 10 minutes and data and samples were taken during the upset and subsequent readjustment period.
- 2) After the plant was back at steady state, the steam rate to the first effect was decreased by one-half. It was noted that the plant responded very sluggishly to this change. The front end temperature profile, the water production, the steam flow to the module, all trended gradually downward over a period of about 1 hour. It was also observed that there was considerable level buildup in the sump to sump effects. This was probably due both to the increase in brine flow caused by the decreased water production and by the decrease in the pressure drop driving force between stages.
- 3) About an hour and a half after the steam rate change but before steady state was reached, the brine rate was cut down to about 300,000 lb/hr. Because of the malfunction of several of the level control devices in the old plant,

the total plant performance was somewhat erratic during this phase of the test work. After about 15 minutes, it was noted that P-4a was cavitating badly and the feed rate was moved up to about 400,000 lb/hr in order to avoid doing any damage to the pump. After another half hour at this condition the plant was returned to normal operating conditions.

c. Tube Gaging - High Temperature Effects

Table G-13-a shows the results of gaging randomly selected tubes in the first four effects at the Freeport Test Facility after Development Run 15.

Table G-13-a

Evaporator Tube Gaging - December 1968

Evap. No.	Nominal Tube I.D. Inches	No. of Tubes Tested	Number of Tubes that Gauge Ball Passed Completely Thru		
			Gauge Ball Size		
			1.84"	1.80"	1.75"
1	1.87	55	11	26	15
2	1.87	40	36	4	--
3	1.87	41	38	3	--
4	1.87	40	37	3	--
5	1.87	0	--	--	--

Effects 2 through 4 are considered to be clean and without scale. The 1st Effect is showing signs of some deposit buildup. During Development Run 15 this effect was run at a 268°F brine sump temperature receiving a feed with an average concentration factor of 1.0. No deterioration of the heat transfer coefficients that could be caused by scaling was observed during Development Run 15. However, the 268°F brine sump temperature is quite close to the incipient scaling point that was observed during two of the ATU runs (about 273°F).

In picking tubes to be gaged, the probologed tubes were chosen. These tubes had originally been selected on a random number basis. Unfortunately, the random number selection technique for a circular tube bundle is partially invalidated by the fact that almost half of the tubes in these bundles were pulled at the time of the module installation. In statistics, when the size and the orientation of the base population is changed, the basis for random number selection has to be changed accordingly.

At the end of Development Run 16, every other tube in every other row in the first effect was gaged with a 1.800 inch diameter gage ball. The gage ball would not pass down about five of the 90 odd tubes gaged. This is a very significant improvement over the gaging results observed for the first effect at the end of Development Run 15. It appears that running the first effect at a temperature of 265° and with unblended Gulf of Mexico sea water served to dissolve some of the scale that had accumulated on the tubes. This is in direct contradiction to normal experience. Once it is formed, calcium sulfate scale cannot be washed off; usually, it tends to get worse even if operating conditions are eased.

d. Effect 14

At the post Development Run 15 shutdown, two tubes in Effect No. 14 were found to be almost plugged with gypsum scale. Forty or fifty other tubes in this bundle were inspected and found to be clean and totally free of any evidence of gypsum scale. No reason could be found for two tubes in this bundle being plugged. The tubes were cleaned by scraping and water-washing.

Forty to sixty randomly selected tubes in Effects 15 through 17 were inspected. There was no evidence of any surface deposit or gypsum scale in these tubes (a foreign object had become lodged halfway up on tube in Effect 15; gypsum scale was starting to form around it).

At the end of Development Run 16, 22 tubes in effect 14 were found to contain gypsum scale. After the effect was water washed for 3-4 days, two tubes still contained sypsum scale. It is assumed that these are completely blocked.

H. APPENDIX

SECTION 1

Symbols & Bibliography

SECTION 2

Schematic Flow for double fluted
tube bundle (13th Effect) with
basis for heat transfer calculations

H. APPENDIX

SECTION 1

A. LIST OF SYMBOLS AND ABBREVIATIONS

ΔT	Temperature difference, °F
A_o	The active outside heat transfer surface, square feet
BPR	The boiling point elevation resulting from the salt content of the brine.
BOD	Biological Oxygen Demand
CF_i	Concentration factor in
CF_o	Concentration factor out
D/A	Deaerator
DO	Dissolved Oxygen
GPM	Gallons per minute
GTD	Greater Temperature Difference, °F
h	Enthalpy, Btu/#
Hg	Mercury
HX	Heat Exchanger
LB/HR	Pounds per hour
LMTD	Log mean temperature difference
LTV	Long-Tube Vertical
ME	Multiple-Effect
MLB/HR	Thousands of pounds per hour
MPPH	Thousands of pounds per hour

A. LIST OF SYMBOLS AND ABBREVIATIONS (CONTD)

MMGPD	Millions of gallons per day
MV	Millivolt
OD	Outside Diameter
P	Pump
pH	Negative logarithm of hydrogen ion concentration
PPB	Parts per Billion
PPM	Parts per million
PSIA	Pounds per square inch absolute
PSIG	Pounds per square inch gage
Q_L	Heat Loss, Btu/hr
Q_T	Heat transferred, Btu/hr
SS	Stainless Steel
SVL	Swirl Vane, Liquid distribution weirs
T	Temperature, °F
TDS	Total dissolved solids, Wt.%
TEMP	Temperature, °F
T_{sb}	Temperature of sump brine
U_o	The calculated actual overall coefficient of heat transfer in Btu/hr-Ft ² -°F
V	Vent
W	Weight Flow Rate, °F
VTE	Vertical Tube Evaporator

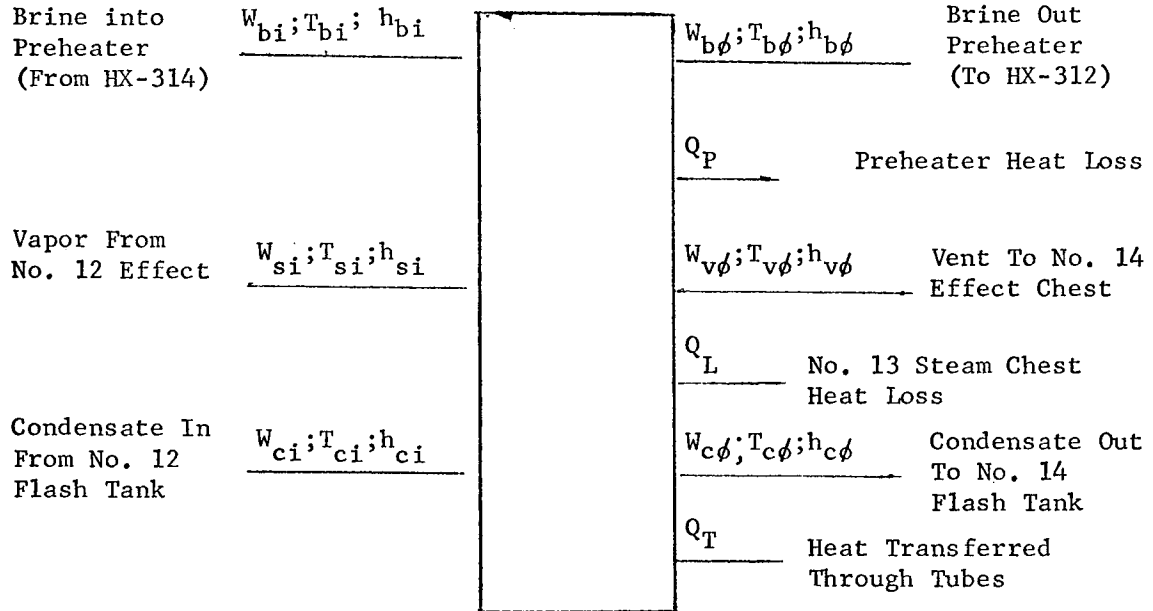
B. BIBLIOGRAPHY

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- (2) Standard Methods for Examination of Water and Waste Water (1960); American Public Health Association, et.al.

APPENDIX

SECTION 2

SCHEMATIC FLOW FOR DOUBLE FLUTED TUBE BUNDLE (13th EFFECT) WITH BASIS FOR HEAT TRANSFER CALCULATIONS



I. LEGEND

A. Symbols

W = Weight flow rate, lb/hr

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

h = Enthalpy, Btu/#

Q_T = Heat transferred through tubes, Btu/hr

Q_L = No. 13 steam chest heat loss, Btu/hr

Q_P = Preheater heat loss Btu/hr

B. Subscripts

bi = Brine in

bo = Brine out

si = Steam in

vo = Vent steam out

ci = Condensate in

co = Condensate out

II. CALCULATIONS

A. Heat Transferred Through Tubes (Q_T):

$$Q_T = (W_{si} - W_{vo}) (h_{si} - h_{co}) \quad \text{Heat in with steam minus vent rate}$$

$$\begin{aligned}
 &+ W_{ci} (h_{ci} - h_{c\phi}) && \text{Heat in with condensate flashdown} \\
 &- W_{bi} (h_{b\phi} - h_{bi}) && \text{Heat out to preheat brine} \\
 &- Q_L \\
 &- Q_P && \text{Heat Losses}
 \end{aligned}$$

B. Overall Evaporative Heat Transfer Coefficient (U_0):

$$U_0 = \frac{Q_T}{A_0(T_{sc} - T_{bs})}$$

Where: U_0 = Overall evaporative heat transfer coefficient, Btu/hr., $^{\circ}\text{F}$, ft.²

A_0 = Outside Heat Transfer area, Ft.²

T_{sc} = No. 13 Effect steam chest temperature, $^{\circ}\text{F}$

T_{bs} = No. 13 Brine sump temperature, $^{\circ}\text{F}$

Note that this assumes isothermal condensation in steam chest and no temperature change in tubes.

III. DATA USED

	<u>Term</u>	<u>Value</u>	<u>Source</u>	<u>Estimated Precision</u>
A.	Heat Transferred Through tubes (Q_T)			
1.	Heat In With Steam Minus Vent			
	W_{si}	-	Measured with	± 400 lb/hr
	$W_{v\phi}$	200 lb/hr	Design for	± 100 lb/hr
		300 lb/hr	Calculated for	± 100 lb/hr
	$(h_{si} - h_{c\phi})$	-	From steam tables assuming isothermal condensation.	-
2.	Heat In With Condensate Flashdown			
	W_{ci}	-	Measured with flow	$\pm 15,000$ lb/hr tube
	$(h_{ci} - h_{c\phi})$	=	$C_p(T_{ci} - T_{c\phi})$	
	C_p	1.0 Btu/lb, $^{\circ}\text{F}$	Heat Capacity of Water	-

	<u>Term</u>	<u>Value</u>	<u>Source</u>	<u>Estimated Precision</u>
	T_{ci}	-	Measured with mercury filled glass thermometer.	$\pm 0.5^{\circ}\text{F}$
	$T_{c\phi}$	-	Assumed Complete flashdown of condensate. $T_{c\phi}$ is same as chest temperature	-
3.	Heat Out to Preheat Brine W_{bi}	-	Measured with orifice meter	$\pm 20,000$ lb/hr

$$(h_{b\phi} - h_{bi}) = C_P(b) (T_{b\phi} - T_{bi})$$

	$C_P(b)$	0.96 Btu/hr, $^{\circ}\text{F}$	Heat Capacity of normal sea water at 160°F	-
	$T_{b\phi}, T_{bi}$	-	Measured with mercury filled glass thermometer.	$\pm 0.5^{\circ}\text{F}$
4.	Heat Losses $Q_L + Q_P$	300,000 Btu/hr	Assumed	$\pm 200,000$ Btu/hr

B. Overall Evaporative Heat Transfer Coefficient (U_O)

1.	Outside Area A_O	1722 ft. ²		± 10 ft. ²
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Double-Fluted Tube Outside Heat Transfer Area (A_O),

Fluted Surface = $(0.8181 \text{ Ft}^2/\text{Ft}) (150 \text{ fluted tubes}) (0.850 \text{ Ft fluted length}) (1.3 \text{ enhancement factor}) = 1,571 \text{ Ft}^2$

Smooth Surface = $(0.8181 \text{ Ft}^2/\text{Ft}) (150 \text{ fluted tubes}) (1.233 \text{ Ft unfluted length}) = 151 \text{ Ft}^2$

$$1,571 + 151 = 1,722 \text{ Ft}^2$$

2.	Temperatures T_{sc}	-	Measured with calibrated mercury filled glass thermometer	$\pm 0.1^{\circ}\text{F}$
	T_{bs}	-	Measured with certified mercury filled glass thermometer	$\pm 0.1^{\circ}\text{F}$

- C. Overall heat transfer coefficient is estimated to be good to ± 50 Btu/Hr, Ft², °F. Most significant limiting measurement is temperature difference between steam chest and brine in the tubes; this is good to about $\pm 0.2^\circ\text{F}$ or $\pm 3\%$ based on the average temperature difference.

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