Analysis and Summary of Reports and Data from the Freeport, Texas, Test Bed Plant (1961-1969)

United States Department of the Interior

March 3, 1849

Office of Saline Water • Research and Development Progress Report No. 759

Int-OSW-RDPR-71-759

Research and Development Progress Report No. 759 • December 1971
Int-OSW-RDPR-71-759

Analysis and Summary of Reports and Data from the Freeport, Texas, Test Bed Plant (1961-1969)

By Dr. G. H. Shroff, I. C. Watson, and R. D. Cross, of Control Systems Research, Inc., Arlington, Virginia, for Office of Saline Water; J. W. O'Meara, Acting Director; J. W. Heintz, Assistant Director, Project Management & Plant Engineering; W. L. Barnes, Jr., Chief, Plant Engineering Division; and D. I. Dykstra, Project Engineer.

Contract No. 14-30-2697

United States Department of the Interior • Rogers C. B. Morton, Secretary
James R. Smith, Assistant Secretary for Water and Power Resources

For sale by the Superintendent of Documents, U.S. Government Printing Office
Washington, D.C. 20402 - Price $3.25
Stock Number 2400-0722

As the Nation's principal conservation agency, the Department of the Interior has basic responsibilities for water, fish, wildlife, mineral, land, park, and recreational resources. Indian Territorial affairs are other major concerns of America's "Department of Natural Resources".

The Department works to assure the wisest choice in managing all our resources so each will make its full contribution to a better United States—now and in the future.

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.

TABLE OF CONTENTS

TABLE OF CONTENTS (cont'd)

TABLE OF CONTENTS (cont'd)

TABLE OF CONTENTS (cont'd)

TABLE OF CONTENTS (cont'd)

LIST OF FIGURES

LIST OF FIGURES (Cont'd)

LIST OF FIGURES (cont'd)

LIST OF TABLES

LIST OF TABLES (cont'd)

LIST OF TABLES (cont'd)

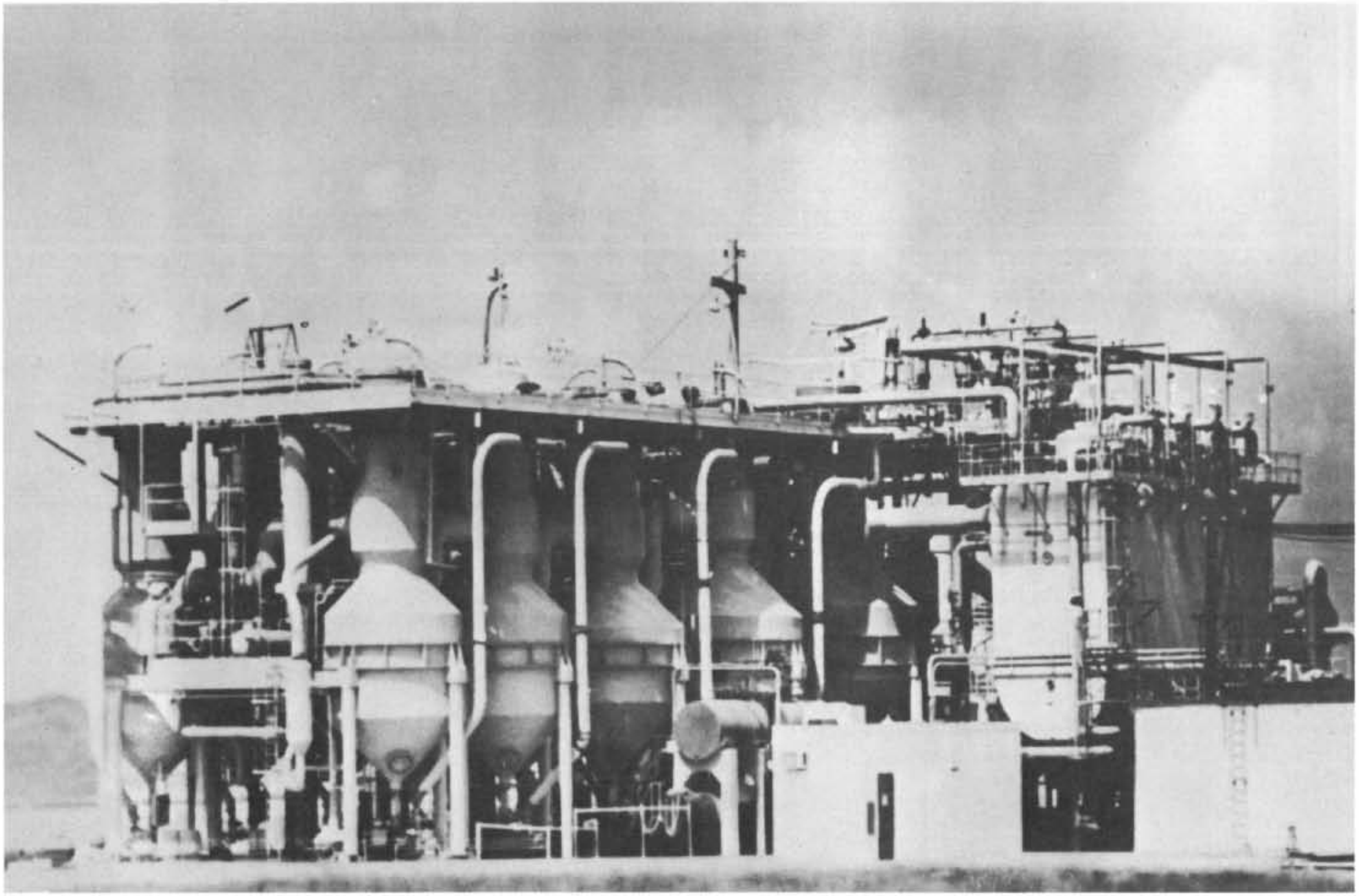
PREFACE

This report presents a summary and analysis of all data pertaining to the Freeport, Texas, VTE seawater desalting demonstration plant. The study, initiated in September 1970, was performed by Control Systems Research, Inc., under contract with the Office of Saline Water, U. S. Department of the Interior.

Ian C. Watson, Manager of CSR's Resource Studies Group, directed the work, while Dr. Gajendra H. Shroff, Associate, performed the bulk of the technical analysis. Mr. Robert D. Cross, Associate, developed the bibliography, and assisted in the development of maintenance and operation material and in the economics analysis.

Mr. Ferris Standiford of W. L. Badger and Associates contributed some of his time to discussion and response to questions, as did Mr. David Kays of Envirogenics. Their assistance is gratefully acknowledged.

Mr. Dewey Dykstra provided invaluable guidance on behalf of the Office of Saline Water throughout the study.



Overall View of the Freeport Vertical Tube Evaporator Test Bed Plant Facility in the 17-Effect Configuration

Chapter 1

INTRODUCTION

A. PLANT HISTORY

In a bill introduced by Senator Clinton P. Anderson of New Mexico, and enacted by Congress in 1958 (P. L. 85-883) the Office of Saline Water was empowered to construct and operate no less than five saline water conversion plants to demonstrate the technical and economic feasibility of desalting brackish and sea water.

The Freeport site and the demonstration plant type of process were announced on March 2, 1959. The design contract was awarded to W. L. Badger Associates, Ann Arbor, Michigan, and the construction contract to Chicago Bridge and Iron Company. Stearns-Roger Corporation of Denver operated the plant until final shutdown in 1969.

The plant was designed to produce 1 MGD of water from seawater feed taken from the Gulf of Mexico. Process steam was supplied from the Dow Chemical Company installation nearby, and brine was discharged via canal back to the Gulf. The product water was divided between the City of Freeport and the Dow facility.

Startup was accomplished in April 1961. Several weeks were spent in shakedown operations, prior to the first official production run in July of 1961. On September 9th, Hurricane "Carla" hit the Texas coast, dealing extensive damage to the installation, and flooding to a depth of six feet. When the hurricane had dissipated, a crash maintenance program was initiated, and the plant was restarted, using river water feed, under emergency conditions. Hurricane damage continued under repair into October 1961, and conditions were finally normalized in November 1961, although long term effects of the hurricane flooding appeared for some time, mainly in the instrument systems.

A major plant modification was instituted in 1967, increasing the number of effects from twelve to seventeen. The additional five effects were added in modular form, a design concept developed by Stearns-Roger Corporation. Also in 1967, a seven-tube Auxiliary Test Unit (ATU) was erected for parallel or series operation with the first effect evaporator, for the purpose of investigating high temperature operation in relation to heat transfer performance and scale control.

The plant was shut down in May 1969 for inspection and non-destructive materials testing. In September of 1969, the plant was decommissioned prior to a major reconstruction of the facility.

B. PURPOSE OF THE REPORT

Control Systems Research, Incorporated, was placed under contract number 14-30-2697 on September 1, 1970, for the purpose of reviewing and summarizing all reports and data generated as a result of Freeport operations for the period 1961-1969. The goal of this work was the reduction of pertinent operations, maintenance, cost and research data to a level which would provide the Office of Saline Water with a single document from which VTE data could be readily extracted. In addition, all reports, letters, memos and miscellaneous data would be cataloged, and a bibliography developed to provide ready access to the original filed material. This report represents the culmination of the one year contractual effort.

C. SUMMARY OF FINDINGS

The Freeport Saline Water Conversion Demonstration Plant No. 1 at Freeport, Texas operated from April 1961 to May 1969. In its initial stages the plant was run primarily as a production-size facility to demonstrate the economic feasibility and long-term operability of the multiple-effect falling-film evaporation process to produce potable water from seawater. After demonstrating these two aspects, the plant became a test-bed facility in 1964 to evaluate on a large scale, concepts that showed promise of improving the plant operations technically and to some extent economically. The following is a brief summary of findings of the Freeport Test Facility:

- The plant capacity could be increased from 1 MGD to 1.21 MGD with 30 percent less than original surface and at a plant ΔT of 126°F (original $\Delta T = 130°F$).

- The high temperature limit of the first effect was raised from 250°F to 268°F while the low temperature limit of 120°F was decreased to 103°F. The brine flashing range thus increased from 130°F to 165°F.

- This ΔT increase allowed conversion of the 12-effect plant to a 17-effect plant, both of which represented an engineering "first" throughout the world. The plant was also the first to use successfully the falling-film process for seawater desalting and the first to use temperatures of 250°F and higher.

- The problem of even distribution of feed water to the heating element tubes (about 500) was solved first by using triangular or V-notch weirs in each tube-top and later by improved swirl-vane weirs and porcelain spray nozzle devices.

- Series 200 product water-to-seawater feed preheating exchangers were replaced, without loss in the steam economy ratio, by the flash tanks associated with each effect except the last. This change resulted in a savings in equipment, pumps, and piping and power consumption.

- Double-fluted and spirally indented surfaces showed more than double the heat transfer rates of the normal smooth surfaces.

- Carbon steel can be used as a material of construction except when high velocity or turbulent conditions exist — for example, it cannot be used for waterboxes, top tubesheets, or for heat element tubes.

- Copper-nickel, monel, and stainless steel were corrosion-resistant even in non-deaerated seawater service. Aluminum brass tubes withstood high velocity (to 7 ft/sec) and turbulence of seawater as heating element tubes in both the heat exchangers and the evaporator effects.

- If the brine evaporation path lies below the gypsum and hemihydrate solubility curves then acidification of feed to pH = 4.0 - 4.5 and the use of polyphosphate compounds at 3 to 4 ppm will ensure scale-free plant operation between 268°F-300°F at a maximum brine concentration factor of 3.2.

- Actual annual average water cost ranged between \$0.76/1000 gal. to \$4.04/1000 gal. Typical month water cost ranged from \$1.12 to \$1.73 per thousand gallons over the period 1963-1969, with October 1964 being recognized as the most economic: 34,845,900 gallons of water being produced at an average cost of 95¢ per kgal.

- The normalized capital cost per gallon of daily capacity was estimated to vary between \$1.424 to \$1.827 over the 1962-1969 period.

Chapter 2

PLANT DESIGN

The basic purpose of a desalination plant is to separate potable water from non-potable waters such as seawater, brackish water, etc., at a reasonably low cost. Speaking broadly, there are only two methods to obtain pure water — either by separation of pure water from the saline water, or by rejection of salts from the saline water. Solar distillation, multiple-effect distillation, multistage flash distillation, freezing out of water, and extraction using hydrate formation fall into the first category while membrane processes (reverse osmosis, electrodialysis) come under the second category. Out of these various desalination techniques, multiple-effect evaporation was selected for the first demonstration plant of OSW at Freeport, Texas, because this was the most developed technique at that time (1960). Also, preliminary studies by W. L. Badger and Associates[1] indicated that low cost water could be produced using this technique and conventional equipment.

A. BASIC PRINCIPLES

Multiple effect evaporators have been used in industry for more than 100 years and have been in use for seawater distillation for about 70 years. The development of seawater evaporators, however, has followed an entirely different course than those built for industrial use. The principal reason for this is the fact that seawater evaporators were originally built for shipboard use where the main requirements were for compactness, simplicity of operation and utmost reliability. Corrosiveness of seawater and its scale depositing

---

(1) "Preliminary Description of W. L. Badger LTV Demonstration Plant," Report 450 by W. L. Badger and Associates, Inc. (June 1959), submitted to OSW.

problems were minimized by using expensive alloys and evaporator arrangements that would allow easy descaling of surfaces.

In large, land based industrial evaporator plants the requirement is much less stringent, a fact which permits the use of cheaper materials of construction, high boiling temperatures, elaborate descaling methods, and most important, the use of the cheapest type of evaporator in flowsheets that are much more complex than usually considered practical. It is important to bear in mind here that unit water costs are relatively unimportant in naval ships, while for civilian use, cost assumes critical importance. This cost reduction was proposed to be achieved by employing industrial evaporation methods that made use of the factors mentioned above.

1. Principles Underlying Multiple Effect Evaporation

When salt is dissolved in fresh water a certain amount of energy is released from the system. Thermodynamically, this is the minimum amount of energy that would be required when separating fresh water from the saline water. This minimum energy requirement has been calculated as 2.7 to 4 KWh (kilowatt-hours) per 1000 gallons product water in the range of temperatures normally used for desalting[1]. Actual energy consumption is many times this value. If saline water is at its boiling temperature, separation of a pound of water vapor requires about 1000 Btu of thermal energy as the latent heat of vaporization which is the equivalent of 2442 KWh per 1000 gallons of water. It is obvious that the water cost would be exorbitant unless this substantial amount of heat were reused. This is the basis of the principle of multiple effect evaporation. A typical single effect used at Freeport is shown in Figure 2.1. The principle of multiple effect evaporation can be understood by referring to Figure 2.2. The first effect, which operates at the highest temperature, is supplied with prime steam from a boiler. For each pound of steam condensed in the steam chest of this effect, approximately one pound of water is evaporated from the seawater flowing down the tubes of the effect. Concentrated brine collects in the bottom of the effect while the water vapor generated is used as the heating medium in the next (second) effect. Brine blowdown from the first effect is used as the feed to the tubes in this second effect. Vapor produced in the first effect now generates approximately one more pound of vapor in the second effect, which is operated at a slightly

---

(1) M. d'Orival in "Water Desalting and Nuclear Engineering", published by Verlag Karl Thiemig, KG München 90, Germany (1967).

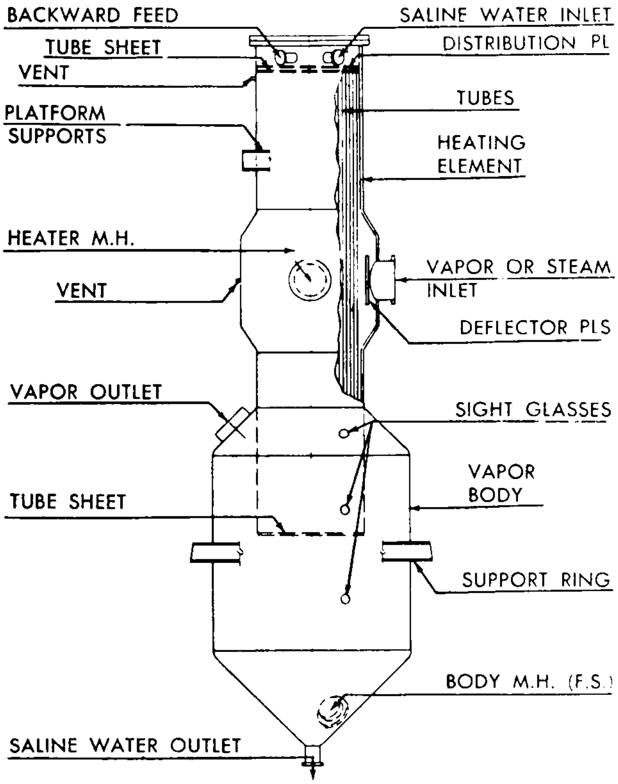


Figure 2.1. Typical Freeport Falling Film Evaporator Effect (Single Pass)

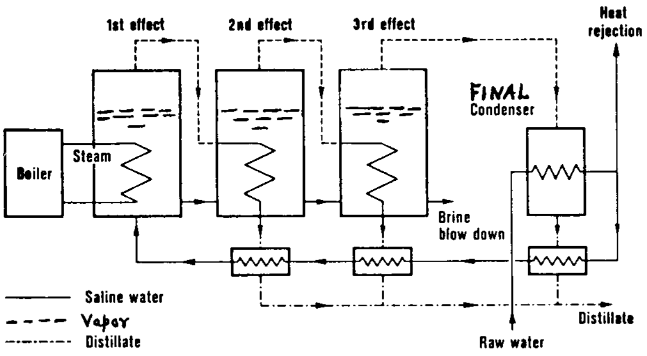


Figure 2.2. Principle of Multiple Effect Evaporation

lower pressure than the first effect. The condensate from the steam chest becomes the product water. Vapor generated in the second effect is again used in the third effect as the heating medium. This process is repeated until the last effect is reached, where cold seawater is used to condense the vapor coming out of this effect. Theoretically then, such an arrangement would produce, for each pound of steam input to the first effect, as many pounds of water vapor as there are effects in the unit. However, due to heat losses, design imperfections, and the temperature driving forces, the actual values are slightly less.

There are two terms commonly used to evaluate the performance of a multiple effect evaporator system. These are the evaporator capacity and the steam economy. The former is defined as the pounds of water vapor generated per unit time. The steam economy is defined as the pounds of vapor produced per pound of prime steam condensed in the first effect. In chemical engineering texts (e.g., "Unit Operations of Chemical Engineering" by McCabe and Smith, McGraw-Hill Book Co.) it is shown that the capacity of a multiple effect evaporator is the same or even less than that of a single effect operating over the same total temperature range as the multiple unit. The increased surface area of the multiple unit, therefore, increases only the steam economy. The optimum number of effects must be determined for each individual case on the basis of fuel costs and the fixed charge rates on capital costs. The cheaper the fuel the less are the number of optimum effects indicated for a given plant.

For better steam economy — which is also referred to as the gain ratio or the performance ratio — the following factors must be taken into consideration while designing the desalting plant:

- Heat losses from all equipment except the heat rejection condenser (condensing vapors from the last effect) must be kept down to economic minimum.
- Cold seawater should be brought to its saturation temperature by regenerative preheating in gradual stages using only low-grade heat from condensate and vapors produced in effects.
- Thermal energy in brine and condensate should be utilized in flashing vapor out for further heating either in the next effect or in the preheat train. Brine flashing in the sump of lower temperature effects is preferable over its flashing in the top water box while separate flash-tanks are used for condensate flashing. Condensate could also be used for

seawater feed preheating but liquid-to-liquid heat transfer coefficients are low relative to condensing vapor-to-liquid transfer coefficients.

- Higher feed liquor temperatures and lower product liquor temperatures improve the steam economy.
- Introduction of feed liquor into the system where the minimum amount of preheating is necessary prior to ebullition. This is the advantage of the backward feed technique but without adequate pretreatment it can lead to scaling problems.

Most of these energy-saving and economy-improving techniques were adopted in the Freeport Plant.

Since unit water costs are significantly reduced by increased plant capacities it is important to list and incorporate in the design the factors included in the basic equation governing the evaporator capacity, which is:

$$W_T = \sum_{i=1}^{i=N} U_i A_i \, \Delta T_i / \lambda_i$$

where

W_T = total water production rate excluding first effect condensate, lb/hr

U_i = overall coefficient of heat transfer in i^{th} effect, Btu/hr ft^2 ^oF

A_i = heat transfer area in i^{th} effect, ft^2

ΔT_i = temperature driving force in the i^{th} effect, ^oF

λ_i = latent heat of vaporization in i^{th} effect, Btu/lb

and N = total number of effects

In this equation $U_i A_i \Delta T_i$ represents the heat transferred, q_i, across the i^{th} effect heating element tubes to seawater flowing down through them. In addition to the water evaporated due to this heat, there is some vapor produced by flashing of brine as it enters lower pressure effects. This is not included in the above equation.

It is obvious from this equation that increased capacities can result from higher values of the overall coefficient U, area A, and temperature driving forces, ΔT, between the condensing steam and the evaporating brine. Latent heat of vaporization increases with decreasing boiling temperatures.

Chapter 4 on "Heat Transfer History" in this report deals with factors that affect U and ΔT. Higher overall ΔT's, though desirable, are limited by the highest temperature that can be attained without scaling problems. Chapter 5 in this report deals with scale control techniques.

Increased heat transfer area A leads to higher capacities but it also increases capital costs. This is the reason for selecting the cheapest materials of construction that can withstand the corrosive tendencies of seawater.

A factor that decreases capacity due to decrease in ΔT is the boiling point elevation (BPR) of seawater. The loss in ΔT due to BPR increases with the increasing concentration of seawater as more water is extracted from the seawater in each succeeding effect.

Since several effects and exchangers are used in a multiple effect evaporator plant their contribution to unit water cost is significant and this dictates careful selection of these pieces of equipment. This is discussed later in subsection E entitled "Selection of Equipment and Materials".

B. ORIGINAL DESIGN

W. L. Badger and Associates of Ann Arbor, Michigan, provided the original design of the Freeport Demonstration Plant. The design requirements of the desalination process for producing 1 million gallons per day (MGD) of fresh water were submitted in the form of process flowsheet, heat and material balance diagram, instrumentation diagram, and drawings of equipment, etc. A document entitled "Specifications No. 195" was prepared by OSW and used for soliciting construction bids. Out of the four bids submitted, the Chicago Bridge and Iron Company's lowest bid for $1,246,250 was accepted.

The original design submitted by Badger and Associates was based on three main considerations:

- use of the cheapest types of evaporator in a flowsheet that was complex but thermally efficient so as to minimize water costs;

- use of magnesium hydroxide seeds in a sludge recycle method of scale control that would permit mild steel as the (cheap) material of construction; and
- corrosion control by use of a deaerator to reduce oxygen level in the seawater feed to the plant.

1. Process Flowsheet

Before entering into the description of the process flowsheet, it is considered worthwhile to recollect the main objectives and design basis of the Freeport Demonstration Plant. "Specifications No. 195" of OSW on page IV-3 states:

> "This plant will be used to demonstrate the technical feasibility of processes developed by W. L. Badger Associates, Inc., for scale prevention in seawater evaporators, to prove that such evaporators can be maintained free of scale at temperatures to at least 250°F, and to determine the economic and operating potentials of flowsheets and equipment for seawater evaporators developed by W. L. Badger Associates, Inc., for the Office of Saline Water of the U. S. Department of the Interior."

It is further stated that:

> "The plant design set forth in these specifications will permit flexibility of operation for determination of the optimum conditions of steam economy, scale prevention, maximum capacity, product water quality, operating and maintenance requirements. Where changes in operating characteristics are not possible in the plant as constructed, provisions have been made in the design to permit changes in piping and auxiliary equipment at minimum future expense. The plant also is designed to determine materials of construction most suitable for a plant operating under these conditions."

The evaporator consisted of twelve effects operated in forward feed manner with seawater feed heated by both the distilled water product and by vapor bled from each effect. The evaporator was the long tube vertical type (LTV) arranged for falling film operation. Boiling pressures ranged from 30 psia in the first effect to 3 in. Hg absolute in the last effect.

The design objective was stated in "Specifications No. 195" as:

> "The plant has been designed to produce on million gallons of potable water per stream day containing no more than 50 ppm total dissolved solids.
>
> "The design has been based on steam at 175 psia and 530°F and seawater feed with 3.5% total dissolved solids. Design capacity must be achieved at all seawater temperatures up to 88°F; product water temperature must be below 100°F when the seawater temperature is below 85°F. Product water to be delivered to the City of Freeport shall be cooled to a temperature no greater than 85°F."

A simplified form of the Badger flowsheet is shown in Figure 2.3. There are three main process streams:

- prime steam and distilled vapor;
- liquor — raw seawater or blowdown from each effect; and
- condensate or product water.

Steam at a pressure of 175 psia is reduced to 35 psia prior to entering the first effect. Vapors generated from the seawater flowing down the heating element tubes of the first effect are used in the second effect steam chest as the heating medium. This is repeated in the rest of the effects until the last effect (XII) is reached. Vapors from this effect are finally condensed by raw seawater in a surface condenser. Vapors are bled from all effects except the XIIth effect to supply the 300 series of liquor preheaters with sufficient thermal energy (supplementing that extracted from the condensate by 200 series of preheaters) to enable the seawater feed (SWF) to be raised to its saturation temperature at the first effect conditions. A small amount (about 600-1000 lb/hr) of the vapors from the XIth effect is used to strip dissolved gases (O_2, CO_2, N_2, etc.) from SWF in the deaerator. Noncondensible gases are removed from the deaerator and the final condenser by means of a vacuum pump to maintain absolute low pressure conditions.

The second stream, the raw seawater feed, enters the plant at the inlet to the product cooling exchanger HX-214 then to HX-213, HX-311,

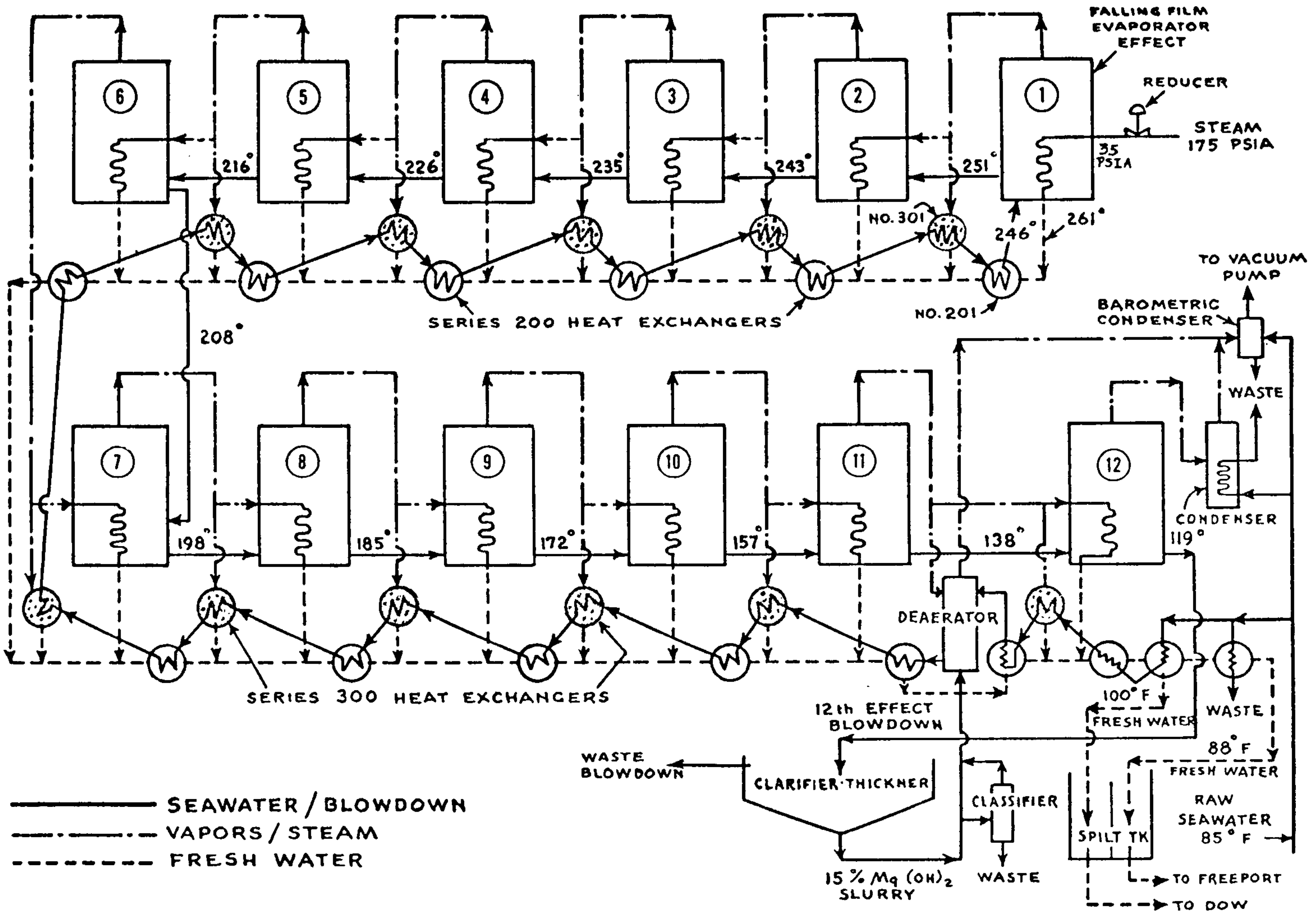


Figure 2.3. Original Freeport M.E.-F.F. Process Flowsheet

HX-212 and finally is sprayed by a nozzle at the top of the deaerator. Vapors from XIth effect pass up countercurrent to the SWF flowing down over the raschig ring packing in the deaerator. In the lower portion of the deaerator a 15-20% slurry of $Mg(OH)_2$ is mixed with the deaerated SWF forming a 1% by weight mixture. This liquor is pumped from the deaerator through 21 more preheaters before entering the top liquor box of the first effect. As mentioned earlier, for maximum efficiency thermal energy is taken in (from vapors and condensate) at the lowest level possible to perform the preheating.

Final heating of the SWF to its boiling point is accomplished in the heating element of the first effect. The concentration of dissolved solids in the liquor increases in proportion to the amount of water removed. The SWF leaving the first effect becomes the "blowdown". This is pumped to the top waterbox of the second effect where it is distributed evenly to all the tubes by means of a distributor plate having 3/16 inch diameter holes (not coinciding with tube openings but over the web section of the tubesheet). The blowdown flows in parallel with the vapors from effect to effect until finally it is rejected from the twelfth effect sump with a concentration factor of 4.0.; i.e., the dissolved solids concentration is four times that in the raw seawater feed. The suspended $Mg(OH)_2$ slurry concentration also increases to 4% from the initial 1%.

The blowdown from the twelfth effect is pumped to the clarifier-thickener where $Mg(OH)_2$ crystals are recovered. The underflow of the thickener (15-20% slurry) returns to the mixing section of the deaerator.

The third stream is the product water collected from the steam chest of each effect and from those of the 300 series preheaters. The condensate formed in the first effect from prime steam is at a temperature of 261°F. This passes through HX-201 and is cooled. Condensate from HX-301 joins this stream as does the condensate from the second effect, all of which pass through HX-202 for further cooling (while heating SWF) and this pattern is repeated in the subsequent effects and exchangers. The last stage in cooling is performed by HX-215 which brings down the temperature of half of the product distillate to 88°F (or lower) for supply to the city of Freeport. The other half of product water is supplied to the nearby Ethyl-Dow Company.

It is worth mentioning here that the Freeport Plant was the first in the world to use twelve (later 17) effects in series, the first to use the falling film type evaporator for seawater, and the first designed to operate at 250°F or above.

The important factors and equations governing the plant operation and performance are shown in Tables 2.1 and 2.2. Table 2.3 lists the variables under the designer's control and the desired results[1].

Table 2.1

Factors Governing Evaporator Performance

PERFORMANCE ITEM	PRIMARY FACTORS	SECONDARY FACTORS
CAPACITY	Heat transfer area Effective temperature drop Overall heat transfer coefficient	Preheating required from sea water intake to Effect 1 Extraction ratio
NET ECONOMY	Number of effects	Preheating required from sea water intake to Effect 1 Approach of sea water feed temperature to preheating vapor temperature at each effect Variation in latent heat of steam with temperature Heat losses to environment Bypassing and hot product withdrawal

---

(1) "Development of a Program to Design ME-LTV Evaporator for Sea Water Conversion" by D. D. Kays of Stearns-Rogers Corp., a paper presented at Western Region Summer Meeting of COMMON, (July 6-8, 1966) Denver, Colorado.

Table 2.2

Primary Equations Relating To
Plant Operation and Performance

NUMBER	ITEM	EQUATION	DEFINITION OF TERMS
1	CAPACITY	$w = \frac{Q}{h_{fg}} = \frac{UA\Delta t}{h_{fg}}$	w = rate of pure water vapor produced Q = rate of heat transfer U = overall coefficient of heat transfer A = surface area of heat transfer Δt = effective temperature difference h_{fg} = latent heat of vaporization
2	EXTRACTION RATIO	$XR = (1 - \frac{CF_i}{CF_o})$	XR = pounds water extracted per pound sea water feed to Effect 1 CF_i = Effect 1 feed concentration factor CF_o = Effect 12 blowdown concentration factor
3	NET ECONOMY RATIO	$ERN = (SFR)(XR)$ or $ERN = \frac{(SWF)(XR)}{HSR}$	ERN = pounds water produced per pound steam fed SFR = pounds sea water feed to Effect 1 per pound steam to Effect 1 XR = extraction ratio SWF = rate of sea water feed to Effect 1 HSR = heating steam rate
4	NET PRODUCTION RATE	$P = (SWF)(XR)$	P = rate of pure water extraction SWF = rate of sea water feed to Effect 1 XR = extraction ratio

Table 2.3

Important Variables and Desired Results

VARIABLES UNDER THE DESIGNER'S CONTROL	DESIRED RESULTS
	Principal Result:
1. Vapor Temperature Profile	Minimum Cost of Water
2. Sea Water Preheating Profile	Subordinate Results:
3. Extraction Ratio	1. Minimum Heat Transfer Surface 2. Minimum Pumping Energy Requirement
4. Heating Steam Temperature	3. Optimum Thermal Economy
5. Heat Rejection Temperature	4. Optimum Geometry 5. Minimum Energy Input
6. Equipment Geometry	6. Minimum Capital Cost 7. Optimum Plant Life

C. CHANGES FOR ACTUAL CONSTRUCTION

The constructors of the Freeport Plant, Chicago Bridge and Iron Company, accepted the basic design supplied by W. L. Badger and Associates, Inc., with only minor changes. This is seen from the Heat and Material Balance Flowsheet drawing No. 506 of W. L. Badger Associates and drawing No. 4 of Chicago Bridge and Iron Co. The latter drawing contains one additional heat exchanger HX-215 for cooling the Freeport product water to 88°F.

The only important process design change was the abandonment of the $Mg(OH)_2$ sludge recycle method of scale prevention within two days of operation of the plant. This was replaced with the acid or pH method of scale control. Concentrated sulfuric acid (66°Be) was added in stoichiometric amounts (corresponding to the bicarbonate ion concentration in raw seawater) upstream of the deaerator which now also acted as a decarbonator.

D. MODIFICATIONS DURING PLANT LIFE

Because of the developmental nature of work at the Freeport Plant a number of process and equipment modifications were introduced and tested during the operational period of the plant. These are discussed in this subsection and in subsection E, "Selection of Equipment and Materials".

1. Process Modifications

The first process modification came with the realization that scaling would occur at the initial design concentration factor of 4.0. This resulted from the use of 1849 gypsum solubility data of Usiglio[1]. Subsequent data[2] indicated that saturation with respect to gypsum is reached at significantly lower concentration factors as shown in Figure 5.4, Section 5, "Scale Control History". Final concentration factor was then lowered to 3.0.

A second important process modification was the use of flash tanks to recover heat from the condensate stream. This was achieved in stages. Initially (during 1961-62) only seven liquid-to-liquid heat exchangers HX-202 through 207 were bypassed and the flash tanks associated with effects I through VII put into service. Condensate pumps P-32 through P-37 were also made redundant with this process modification which produced reductions in power consumption and maintenance costs without appreciably affecting the steam economy (OSW R&D Report No. 71, p. IV-2). The rest of the 200 series exchangers except the final product coolers were eliminated with the major plant modification to 17 effects. Condensate heat recovery was then in the all-flash mode. The additional heating load resulting from removal of 200 series exchangers actually improved the performance of the vapor condensing preheaters (Series 300). This is discussed in Chapter 4.

Another important process modification was the use of polyphosphates to control (and even remove) gypsum scale. This allowed the use of high concentration factors in the low temperature (less than 190°F) effects. Use of brine recirculation from the sump of a given effect to the top water box of the same effect was investigated to see if it improved heat transfer rates and prevented scaling arising from maldistribution of brine feed to tubes in the evaporator effects. Annual Report No. 2 by Stearns-Roger (OSW R&D Report No. 100) contains the following relevant comment: "In conclusion, the brine recirculation (approximately 100 gpm brine flow to Effect XII) is somewhat disappointing in that the Plant still cannot be run continuously at a

---

(1) Standiford, F.C. and H.F. Bjork, Chem. Eng. Prog., 63, No. 1, 70 (January, 1967).

(2) Tanaka, Y., K. Nakamura, and R. Hara, J. Chem. Soc. Japan, Ind. Chem. Sect., 34, 779 (1931).

final brine concentration factor of 4.0, but tests have established that it is of value in maintaining scale-free conditions at a concentration factor of 3.0." Brine recirculation is discussed in more detail in Chapter 4, Heat Transfer History.

2. Equipment Modifications

Several equipment modifications were carried out at the Freeport Plant with the ultimate aim of increasing the efficiency of the process and/or reducing costs. These changes are described in detail in the eight annual reports by Stearns-Roger to OSW. The following were some of the important modifications:

a. Brine Distribution Weirs — Distribution of feed brine to about 500 tubes of the heating elements in the various effects was recognized to be a problem from the design and construction stage. The original distributor plate with some 600 holes of 3/16 inch diameter installed by the constructors (Chicago Bridge & Iron Co.) did not provide even distribution of liquor to the tubes, which led to starvation of some tubes to the point where scale deposition became a serious problem. The 3/16 inch diameter holes would easily choke up with silt, scale or corrosion products, worsening the liquor distribution problem. A triangular notch weir was therefore installed on each tube opeing in the upper water box. This technique was quite successful in ensuring an even liquor distribution. For further improvement in operation, Swirl Vane Liquor distributing weirs (SVL) made of both an alloy and a plastic, and Porcelain Spray Nozzles were studied. Further discussion is given in the Heat Transfer History Chapter.

b. Change of Carbon Steel Tubing — Evaporator Effects I, V, VI, VIII, and XI were equipped with carbon steel tubes to test their performance in seawater service. Annual Report No. 2 (OSW R&D No. 100) by Stearns-Roger has the following comment:

> "In January 1963, during a scheduled major shutdown, a pressure test on the Effect I evaporator revealed approximately twenty-five leaking tubes. A similar test on Effect V revealed seven leaking tubes. The remainder of the carbon steel tubed effects (VI, VII, and XI) did not leak, but definite pitting was evident. It was necessary to completely remove the

> carbon steel tubing from all effects so equipped, and replace with aluminum brass.
>
> "This alloy was selected because of its relatively good corrosion resistance and competitive price. At this date there have been no tube failures of copper-base alloy tubes which can be attributed to corrosion. Test specimens indicate that of all the copper-base alloys investigated, aluminum brass and 90-10 copper-nickel have the lowest corrosion rates in seawater and hot brine service."

Carbon Steel tubes from the preheaters had to be removed for similar reasons but much earlier because they were exposed to low pH (acidified and containing dissolved CO_2), whereas evaporator tubes (other than the first effect) were exposed to CO_2-free liquor.

c. High-accuracy Instrumentation — Numerous instruments such as a seawater pH recorder, deaerator level recorder, Effect XII temperature and flow rate recorders, level controllers on all effects, long thermowells, etc., were installed during the fiscal year 1963. In addition, it was necessary to equip Effects I, VI, X, XI, and XII with instrumentation to obtain highly accurate data on two-phase pressure drops in the tubes and the associated temperature drops to verify Dukler's theory (OSW R&D Report No. 74). This theory indicated that the calculated values of the overall heat transfer coefficient in the low temperature effects (especially Effects X, XI, and XII) were low in comparison to their true values. This discrepancy resulted from the use of a value for the temperature driving force, ΔT, that was too large. The effective, or actual, ΔT was less than the temperature difference between the condensing steam and evaporating brine at the tube exit (apparent ΔT) because of the pressure drop inside the tubes. Specific volumes of water vapor increase rapidly at low temperature/low pressure conditions so that their velocities are high inside the 2 inch diameter tubes of the heating elements of these effects. Pressure drop in a 2-inch tube was found to be five times as high as that in a 3-inch tube of same length.

The high accuracy instrumentation was also useful in obtaining accurate heat and material balance data which resulted ultimately in better evaluation of plant characteristics. Figure 2.4 shows the typical high accuracy instrumentation of an effect.

d. Vacuum Maintenance System — The original mechanical vacuum system could maintain a vacuum corresponding to about 115°F final

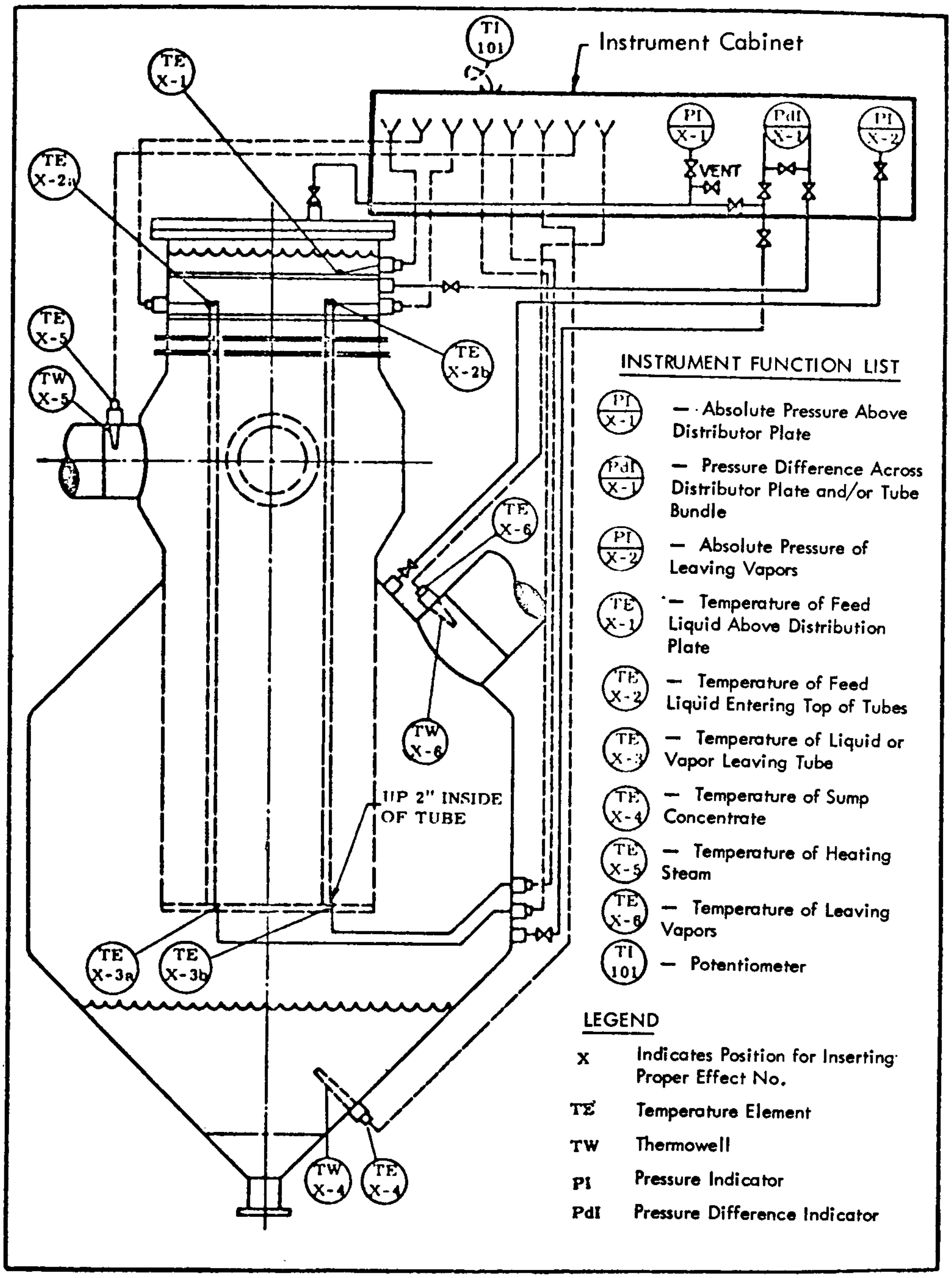


Figure 2.4. Typical Effect Instrumentation Diagram

condenser temperature. To increase the process capability of operating the Plant at a final condensing temperature of 100°F, a new steam jet ejector system including barometric condenser, steam inter-condenser, and two-stage ejectors was installed in June of 1966[1]. This system was operated in parallel with the mechanical system. Required dry and saturated steam supply pressure was 125 psig at the inlet nozzle to the jets and the steam flow rate was 405 lb/hr per stage for removing 180 lb/hr vapor (equivalent air) from the evaporator system. Performance was stated to be satisfactory under normal load conditions.

e. Double-fluted Tubes in Effect XIII — The installation and testing of these enhanced surface tubes in Effect XIII (of the five-effect module) was carried out in March of 1967 (OSW R&D No. 479, p. 171). The tubes were Olin Alloy 194 material formed from 3.125 inch O.D. x 0.065 inch tubing. The finished tubes were 11 feet 1 inch overall length, with 9 foot 10 inch section fluted both sides to a resultant 0.064 inch wall thickness at the flute and having a 4 1/2 and 10 1/2 inch smooth ends. The tube bundle had 151 tubes. Tremendous improvement in the overall coefficient U was noted with these tubes. Chaper 4 on Heat Transfer History contains further details.

f. Horizontal Vapor Shear Preheater, HX-306 (X) — The Oak Ridge National Laboratory designed this horizontal enhanced (spiral-indented) surface preheater. Seawater passed through the horizontal tubes in a single pass while steam (vapor from Effect VI) condensed on the outside of the tubes in a shell having three passes to increase steam velocity. This preheater replaced an original HX-306 (b) unit with smooth vertical tubes. The new condenser was installed to demonstrate, under field conditions, that the ORNL spiral-indented tube did, in fact, enhance the overall heat transfer coefficient, and that the high steam mass velocity was essential in a condenser subject to noncondensibles accumulation. As discussed later in the Heat Transfer History chapter, these objectives were basically achieved since an overall U of about 970 Btu/hr ft^2 °F was obtained for 3100 hours of operation (Annual Report No. 8 by Stearns-Roger, p. 23) with 300 lb/hr ft^2 steam mass velocity (5% of inlet steam vented) and the increase in mole fraction of noncondensibles from 0.01 to 0.23 did not appreciably reduce the overall U[2]. It should be noted that the previous smooth tube unit had U values around 370 Btu/hr ft^2 °F.

---

(1) OSW R&D Report No. 253, pp. 177-179. This is the Fifth Annual Report by Stearns-Roger Corporation.

(2) Communicated by Dr. D. M. Eisenberg of ORNL, Jan. 4, 1971.

There were a number of other modifications to the original plant design amongst which the following are worth mentioning:

- Change of original 3-inch ceramic raschig ring packing in the deaerator to 3 3/4 inch Maspac FN-90 polystyrene packing which reduced the damage to the inside coating of the deaerator steel tower.
- Installation of wire-mesh entrainment separators in the low temperature effects to reduce the brine carryover.
- Installation of alloy water box liners in effects and preheaters to reduce corrosion.
- Use of clarifier-thickener as a settling tank for silt in raw seawater feed.
- Use of fiberglass reinforced epoxy plastic material for piping, domes, etc.
- Product water stabilization to reduce its corrosive tendencies.
- Shock chlorination of seawater feed once every week for two hours to reduce marine growth, slime formation, etc.

The details on these and other equipment and process modifications are discussed in the eight annual and development reports by Stearns-Roger Corporation submitted to OSW.

3. Freeport Plant Conversion to 17 Effects

The original 12-effect plant started operation in April 1961. Stearns-Roger Corporation, who operated the plant from that date, demonstrated the large scale operability and reliability of the vertical-tube falling-film multiple effect evaporator desalting process. Production of the first billion gallons of product water was attained on May 21, 1965. The plant, by this time, was capable of producing in excess of the million gallon-per-day rated capacity, utilizing only two-thirds of the installed evaporation heat transfer surface (Sixth Annual Report by Stearns-Roger Corporation, p. 1).

The feasibility of further improvements in operation and cost was studied by Stearns-Roger who submitted the results in a document entitled "Feasibility Study — A Study to Increase the Efficiency of the Freeport Demonstration Plant" (September 30, 1965). Five alternatives involving 17-, 21-, and 24-effect evaporator systems were studied and reported in this document. OSW selected the 17-effect alternative involving extension of the 12-effect plant to a 17-effect plant by the addition of a 5-effect Stearns-Roger module. This alternative involved no presoftening of the seawater, use of the available temperature range (275°F to 103°F), and only minimal changes in the existing equipment and operating procedures to improve efficiency and reduce costs (p. 3, "Feasibility Study", op. cit.). The following process improvements and technological developments were predicted to result from the 17-effect evaporator (p. 8 & 9, "Feasibility Study"):

- The demonstration of the performance of a greatly simplified and reduced capital cost multiple effect installation.
- Thermal economy of the process will be improved by a factor of approximately 40% with an increase of only 14% in total installed heat transfer surface, while maintaining or improving the production rate.
- A major improvement in the heat rejection condenser approach to the available cooling water temperature, utilizing power condenser technology and equipment of standard manufacture.
- Heat transfer correlations to flow conditions and to lower overall temperature differences in various tube lengths, including full control of recirculation within individual effects. Coefficients exceeding 900 Btu/hr °F are expected.
- Heat transfer and vapor handling can be evaluated at the lowest temperatures and pressures yet attempted in the multiple effect process, thereby demonstrating the ability of the process to make maximum utilization of low grade heat.

A comparison of the 12- and 17-effect plant design parameters is presented in Table 2.4. Discussion on the actual performance of the 17-effect plant is given in Chaper 9.A, entitled "Evaluation of Overall Performance".

Table 2.4

Design Comparison, 12- and 17-Effect Plants

DESIGN VARIABLES	12-EFFECT PLANT	17-EFFECT PLANT
Number of Heat Exchangers	27	19
Installed Heat Transfer Surface (Square Feet)		
Evaporative	64,550	57,500
Preheating	37,760	36,900
Heat Rejection	4,190	2,500
TOTAL	104,500	96,900
Number of Installed Process Pumps	32	25
Nominal Capacity, Millions of Gallons per day, net	1.0	1.0
Maximum Brine Temperature, °F	250	265
Brine Flashing Range, °F	132	162
Heat Rejection Temperature, °F	115	100
Gain Ratio, Net (LBS water per LB heating steam)	10.5	13.8
Extraction Ratio (LBS water extracted per LBS sea water feed)	0.71	0.68
Sea Water Supply Rate, LBS/HR	490,000	510,000
Power Demand, Kilowatts (including intake pumps)	360	450

E. SELECTION OF EQUIPMENT AND MATERIALS

For the original 12-effect plant the main criterion in the selection of equipment and materials of construction was that of keeping capital and operating costs down to the minimum necessary for efficient, long term continuous operation. This is expressed in the following quote from the "Preliminary Description of W. L. Badger LTV Demonstration Plant", Report 450 by Badger and Associates to OSW in July 1959:

> "The selection and proportioning of equipment will be dictated wherever possible by the desire to obtain optimum costs for the basic plant. Equipment designs will be changed only where absolutely necessary to make it possible to test operability of the various alternates. As a result, the production rate, fuel and power consumption and equipment configuration will not be the best obtainable by use of these alternate cycles."

1. Selection of Equipment

The following were the major items of equipment at the Freeport 12-effect plant:

a. Evaporator effects — total 12.

b. Heat exchangers — two types: series 200 for condensate heat recovery and series 300 for bleed-vapor heat recovery. Total number of units 15 and 12, respectively.

c. Flash tanks — total 11, one for each effect, except effect XII.

d. Deaerator/decarbonator.

e. Vacuum system — mechanical vacuum pump plus a steam jet ejector.

f. Clarifier-thickener and classifier.

g. Desuperheater for DOW steam.

h. Barometric condenser.

i. Pumps.

j. Instrumentation.

a. Evaporator Effects — There are many types of evaporator effects in use in industry which can be classified broadly as either the natural circulation type or the forced circulation machines. The natural circulation evaporators utilize the density difference arising from temperature difference for the movement (or circulation) of the liquor that is to be evaporated. In the second type, external energy usually in the form of a pump is utilized to create this circulation. It is obvious that the forced circulation evaporators have higher liquor velocities and hence higher rates of heat transfer, although at an additional cost over that of a natural circulation evaporator. Differences within this broad classification arise from the type of heating element arrangement used, from circulation or once-through passage of the liquor through the heating element, and whether the liquid film is rising or falling through the tubes.

The selection of the type of evaporator to be used in the Freeport Plant was based on the following considerations:

- it had to be low in cost to reduce water costs;
- it had to have high heat transfer rate capability to reduce heat transfer surface requirements for a given production rate;
- it had to be adaptable for use in a multiple effect system which allowed a limited temperature differential as driving force; and
- it had to be capable of handling a 1% to 4% suspension of fine $Mg(OH)_2$ crystals.

Pilot plant experiments by Badger Associates at Wrightsville Beach, N. C., indicated that high heat transfer coefficients could be obtained in a falling-film type LTV (long-tube-vertical) evaporator. A sketch of this evaporator is shown in Figure 2.1. Brine enters the top liquor box and flows downward through the tubes as a thin film. The water vapor produced flows downward concurrently with the thin film and in the vapor head the two are separated. The concentrated liquor (brine blowdown) is pumped to the next effect water box while the vapor is used as the heating medium in the next effect. The falling-film evaporator is a single pass or once-through evaporator. Solids handling was no problem with this type since the positive downflow "washing" action within the tubes ensures removal of solids to the sump. Chicago Bridge & Iron Company, the constructors of the Freeport Plant, incorporated this evaporator in their design.

Detailed "as built" drawings of the 12 evaporator effects are available in OSW as Drawings Nos. 5 through 76. Each effect consists simply of a shell and a tube heat exchanger (also called "heating element") mounted over a vapor-liquid separator called the vapor head. The heating element consisted of a mild steel shell of 5 foot 6 inch I. D. having 2-inch I. D. 24-foot long tubes of different alloys in different effects. Although the number of holes in each effect tubesheets was 560, the installed number of tubes varied from 383 to 535. The diameter of the vapor head varied from 10 feet (Effect I) to 18 feet (Effect XII). Pilot plant heat transfer data ("Specifications 195", p. IV-22) indicated overall coefficients U to be in the range 377 — 850 Btu/hr ft^2 $^\circ$F, the higher values being for the high temperature effects. The active number of tubes (surface area) are determined by the overall coefficient value applicable and the plant capacity desired according to the equation $A = Q/U\Delta T$. Original evaporator heat transfer areas are given in Table 2.5. Later, in 1964 and 1965, it was demonstrated that the rated capacity (1 MGD) could be achieved with only 48,347 and 42,636 ft^2 of evaporative surface as against 62,562 ft^2 of the original surface.

As mentioned earlier, the use of enhanced (double-fluted) surface tubes almost doubled the performance of the evaporator effects. Quantitative data relating to the evaporator operation is given in the Heat Transfer History chapter.

b. Heat Exchangers — Heat exchangers played a critical role in the Freeport Plant. This becomes evident when it is realized that the total amount of heat recovered in these seawater preheaters is approximately 70 million Btu/hr which is about twice the amount of heat (35.7 million Btu/hr) brought in by the prime steam to Effect I.

As stated earlier, Series 200 preheaters recovered heat from the condensate from all the effects. These were two pass, shell and tube heat exchangers having shell diameters ranging from 24 inches to 32 inches I.D. Exchangers 201 through 207 had 8-foot long tubes, and exchangers 208 through 214 had 20-foot long tubes. According to "as built" drawings (Numbers 300-2 through 338) by C. B. & I., HX-201 through 207 had 250 holes in the tubesheet but the actual number of installed tubes varied from 178 to 228. These were 1-inch O.D. tubes on 1 1/4 inch triangular pitch, of different alloys (admiralty, aluminum brass, 90-10 copper-nickel, and carbon steel) in different effects, to test their performance in seawater service. Heat exchangers 208 through 210 had 325 tube holes but the number of installed tubes varied from 224 to 311. The empty holes were plugged with carbon steel plugs. HX-211 through 214 had 460 tube holes and the number of tubes installed varied between 355 and 440. Freeport water cooler, HX-215, had all the 580 holes tubed out.

The vapor condensers, i.e., Series 300 preheaters, heat the incoming seawater with vapor bled from each effect vapor head. There are two longitudinal passes on both tube and shell sides. The heat exchanger shell diameter was 28 inch I.D. Although 343 tube holes were made on 1 1/4 inch triangular pitch, the actual number of 1-inch O.D. tubes installed varied from 265 to 343. Other tube holes were plugged. Tube materials utilized were

Table 2.5

Evaporator Heat Transfer Surface Summary

EFFECT	ORIGINAL TUBES						PRESENT TUBES						
NO.	NO.	I.D. IN.	WALL THICK IN.	MAT.	K*	TOTAL INSIDE SURFACE FT^2	NO.	I.D. IN.	WALL THICK IN.	MAT.	K	TOTAL INSIDE SURFACE FT^2	SURF. INCREASE FT^2
I	490	1.834" (2"O.D.)	.083" (14 Ga)	CS-475 Corten 15	26.0	5648	490	1.87" (2"O.D.)	.065"	Al-Br	58.0	5758	110
II	467	1.87" (2"O.D.)	.065" (16 Ga)	Al-Br	65.0	5487	467	1.87"	.065"	ADM	65.0	5487	0
III	467	1.87" (2"O.D.)	.065" (16 Ga)	Al-Br	58.0	5487	467	1.87"	.065"	Al-Br	58.0	5487	0
IV	383	1.87" (2"O.D.)	.065" (16 Ga)	90-10 Cu-Ni	27.2	4500	383	1.87"	.065"	90-10	27.2	4500	0
V	400	1.834" (2"O.D.)	.083" (14 Ga)	CS	26.0	4611	429	1.87"	.065"	Al-Br	58.0	5041	430
VI	535	1.834" (2"O.D.)	.083" (14 Ga)	CS-520 Corten 15	26.0	6167	535	1.87"	.065"	Al-Br	58.0	6286	119
VII	425	1.87" (2"O.D.)	.065 (16 Ga)	Al-Br	58.0	4994	425	1.87"	.065"	Al-Br	58.0	4994	0
VIII	490	1.834" (2"O.D.)	.083 (14 Ga)	CS-475 Corten 15	26.0	5648	490	1.87"	.065"	Al-Br	58.0	5758	110
IX	467	1.87" (2"O.D.)	.065 (16 Ga)	ADM	65.0	5487	467	1.87"	.065"	ADM	65.0	5487	0
X	425	1.87" (2"O.D.)	.065 (16 Ga)	Al-Br	58.0	4994	425	1.87"	.065"	Al-Br	58.0	4994	0
XI	430	1.834" (2"O.D.)	.083 (14 Ga)	C.S.	26.0	4957	560	1.87"	.065"	Al-Br	58.0	6580	1623
XII	390	1.87" (2"O.D.)	.065 (16 Ga)	ADM	65.0	4582	554	1.87"	.065"	Al-Br	58.0	6510	1928
Totals	5369					62562	5692					66882	4320

NOTE: Total No. of Holes ea. tube sheet 12x560 = 6720
Plugs 2(6720-5692) = 2056
Inside Surface per tube (24' length) 16 ga 11.750 FT^2
Inside Surface per tube (24' length) 14 ga 11.527 FT^2

REF: * 16th Biennial Materials of Construction Report, Chemical Engineering, November 1954.
* National Bureau of Standards, Circular 592.
* Baumeister et al., Marks Mechanical Engineers Handbook, 5th Edition, Mc-Graw-Hill, New York, 1963.

carbon steel, admiralty, and 90-10 cupro-nickel alloy. Because of the high overall heat transfer coefficients in condensing steam-to-water exchangers, heat transfer surface is better utilized in the 300 Series units compared to the 200 Series preheaters.

Heat exchanger 312 was a surface condenser to condense the vapor from the last effect (Effect XII) of the evaporator. Proper operation of this unit is especially important since the overall plant temperature driving force ΔT (first effect condensing steam temperature - final effect vapor temperature) is partially controlled by the efficiency of heat transfer in this unit, and the smaller the plant ΔT, the smaller is the plant capacity, according to the basic equation $W = UA\Delta T/\lambda$. Sketches of HX-200 and -300 preheaters are presented in Figure 4.28 in the Heat Transfer History chapter. Surface condenser HX-312 is shown in Figure 2.5. The two-pass and four-pass tube-side flow patterns obtainable with this unit are shown in Figure 4.3 in the Heat Transfer History chapter. Although pump work is increased in two-pass mode there is a simultaneous increase in heat transfer efficiency.

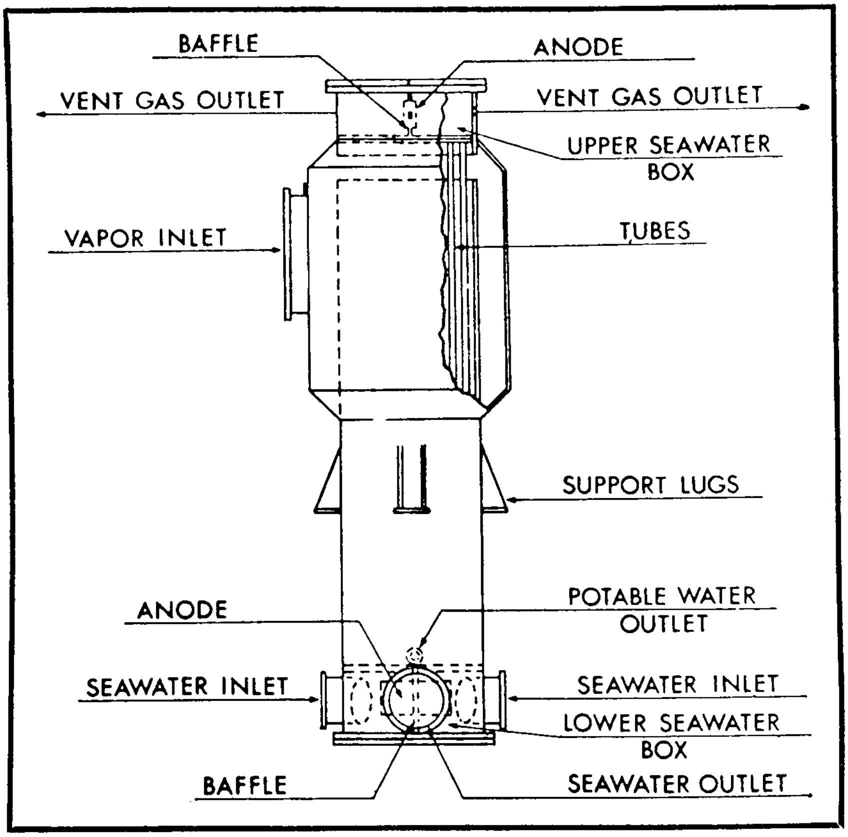


Figure 2.5. Surface Condenser, HX-312

Heat exchangers 212, 213, 214, 311 and 312 which were exposed to non-deaerated seawater were provided with cathodic protection against corrosion by the use of DOW Type 24W magnesium anodes and 90-10 cupro-nickel cladding at least 15% of the metal thickness of tubesheets and water boxes. Tables 2.6, 2.7, and 2.8 provide the specific design conditions for Series 200 and 300 exchangers (taken from pp. IV-26 and 27 of "Specifications 195" by OSW). The actual performance data of the exchangers are discussed in the Heat Transfer History chapter.

c. Flash Tanks — The original purpose of the 11 flash tanks associated with the twelve effect system is stated in the "Specifications 195" document on page IV-18 thus:

> "The condensate flash system will not normally be operated, and is included primarily to aid in disposal of contaminated condensate when operating under other than warranted conditions. This flash system must be capable of handling half of the total plant production, flashing the condensate in stages to each effect of the evaporator below that from which the condensate is first diverted into the flash system. Since the primary purpose is to segregate contaminated condensate, the use of the heating elements as flash chambers is not permissible. The condensate flash system will be tested by artificially contaminating first condensate and diverting this plus the condensate from effects II through VI to the flash system when operating at a total production rate of one million gallons per stream day. When tested in this manner the vapor from each flash tank shall contain no more than 0.5 percent entrained liquid."

The specifications for the tanks on page 37 were given by W. L. Badger and Associates.

Table 2.6

Summary Sheet for Series 200 Heat Exchangers

Shell side fluid - distilled water
Tube side fluid - sea water containing 1.0% $Mg(OH)_2$ solids
- operating flow rate - 440,000 lb/hr
Type - shell and tube - two pass tube side - two longitudinal passes shell side
Corrosion allowance - 0.10 inches total on steel shells, tube sheets, water boxes and baffles
Design - see Drawing E-182-501 by W. L. Badger Associates
Design Temperature - 300°F
Tubes - 1" O.D. x 18 ga. (alloy)
x 14 ga. (steel)

H.E. No.	Duty Btu/hr 1000	Tube lgth Ft.	Tube Matl	Tube Side Fluid Temp°F In	Out	Press.-Psia Op'g.	Des.	Drop	Noz Size	Flow lb/hr	Shell Side Fluid Temp.-°F In	Out	Press-Psia Op'g	Des	Drop	Nozzles In	Out
201	366	8	STL	245	246	42	250	1.4	8"	27,745	261	248	40	115	Less than	3	3
202	515	8	ADM	237	238	45	250	0.9	8"	56,342	249	240	40	115	0.5	3	3
203	574	8	AL-BR	229	230	48	250	0.9	8"	85,005	241	234	40	115	"	3	3
204	918	8	90/10	220	222	51	250	0.9	8"	113,523	234	226	40	115	"	4	4
205	975	8	STL	210	212	54	250	1.4	8"	141,938	226	219	40	115	"	4	4
206	914	8	ADM	202	204	58	250	0.9	8"	169,906	218	213	22	115	"	6	6
207	1937	8	AL-BR	192	195	61	250	0.9	8"	197,403	212	205	21	115	"	6	6
208	3138	20	STL	179	186	65	250	3.0	8"	224,350	204	190	21	115	"	8	8
209	3594	20	ADM	165	173	70	250	1.8	8"	251,603	189	175	20	115	0.6	8	8
210	3774	20	AL-BR	150	159	74	250	1.8	8"	278,854	175	161	20	115	0.8	8	8
211	4370	20	STL	134	144	79	250	3.0	8"	305,822	160	146	19	115	0.9	8	8
212	3392	20	90/10	126	134	25	250	1.8	8"	305,822	146	135	18	115	0.9	8	8
213	6642	20	ADM	98	114	28	250	1.8	8"	332,424	135	115	16	115	1.2	8	8
214	5487	20	AL-BR	85	98	30	250	1.8	8"	355,854	115	100	15	115	1.4	8	8

Table 2.7

Summary Sheet for Heat Exchangers 301-311

Shell side fluid - **design basis** - **condensing evaporator vapors**
- for special tests - evaporator blowdown liquor
Tube side fluid - **sea water containing** 1.0% Mg $(OH)_2$ **solids**
- Operating flow rate - 440,000 lb./hr.
Type - **shell and tube** - **two pass tube side**
-two longitudinal passes shell side when used **with evaporator blowdown liquor**
Corrosion Allowance - 0.10 inches total on steel shells, tube sheets, water boxes, and baffles.
Design - see Drawing E-182-502 by W. L. Badger Associates
Design temperature - 300°F
Tubes - 1" O.D. x 20' - 0" long x 18 ga. (alloy)
14 ga. (steel)

			Tube Side Fluid						Shell Side Fluid - Design Basis							
			Temp.-oF		Pressure-Psia						Press-Psia		Nozzles			
H.E. No.	Duty Btu/hr 1000	Tube Mat'l	In	Out	Op'g.	Des.	Drop	Noz. Size	Flow lb/hr	Temp °F	Op'g	Des.	Vap.	Cond.	Vent	Aux(2)
301	2922	STL	238	245	44	250	3.0	8"	3090	250	29.8	90	8"	1"	1"	4"
302	2858	ADM	230	237	47	250	1.8	8"	2008	242	25.9	90	8"	1"	1"	4"
303	2937	AL-BR	222	229	50	250	1.8	8"	3072	234	22.4	90	8"	1"	1"	4"
304	3304	90/10	212	220	53	250	1.8	8"	2435	225	18.9	90	8"	1"	1"	4"
305	2504	STL	204	210	57	250	3.0	8"	2586	215	15.6	90	8"	1"	1"	4"
306	2871	ADM	195	202	60	250	1.8	8"	2947	206	13.0	90	8"	1"	1"	4"
307	2405	AL-BR	186	192	62	250	1.8	8"	2453	196	10.6	90	8"	1"	1"	4"
308	2391	STL	173	179	68	250	3.0	8"	2391	183	8.6	90	10"	1"	1 1/2"	4"
309	2602	ADM	159	165	72	250	1.8	8"	2612	170	6.0	90	10"	1"	1 1/2"	4"
310	2418	AL-BR	144	150	76	250	1.8	8"	2404	154	4.1	90	10"	1"	1 1/2"	4"
311	5204	ADM	114	126	26	250	1.8	8"	5117	135	2.5	90	10"	1"	1 1/2"	4"

Table 2.8

Specifications for Heat Exchanger 312

Duty - 24,102,000 Btu/hour design

Design - See Drawing 503 by W. L. Badger Associates

Special Features - Cathodic protection for each water box section, (5) total

<table>
<tr><th>Shell Side Conditions</th><th>Tube Side Conditions</th></tr>
<tr><td>Fluid - water vapor from XII effect
Pressure - 3.0" Hg abs.
Temperature - 4°F superheat
Flow rate - 23,430 lb/hr.
Non-condensible load - normal - 20 lb/hr. CO_2 + 5 lb/hr air
maximum - 50 lb/hr CO_2 + 5 lb/hr air
Nozzles - Vapor inlet - 42"
Condensate outlet - 3"
Vents - 2-4"</td><td>Fluid - non-deaerated seawater from pumps P-1 P-1 and/or P-2
Pressure - 0-20 psia
Inlet temperature - Design - 88°F
Actual - mean monthly temperature varies 60-88°F
Flow arrangement - 4 pass with one pump operating
2 pass with both pumps operating
Flow rate - 3100-4000 gpm - 4 pass
7100-8000 gpm - 2 pass
Allowable pressure drop - 7.5 psi
Nozzles - pump inlets - 16"
4 pass return - 18"</td></tr>
<tr><td colspan="2">Materials - Tubes - 1" O.D. x 16' x 0" x 18 BWG aluminum brass on 1 1/4" triangular pitch
- Shell - vapor side - steel
- Tube sheets, heads, water boxes - Muntz metal or 15% 90/10 cupro-nickel clad
Corrosion Allowance - 0.10" on steel</td></tr>
</table>

FLASH TANKS (Dished Head, Cone Bottom, 6' Straight Side)

FT- 1	Impure Condensate from I - Vapor to II -	24" dia.
FT- 2	Impure Condensate from II and/or FT-1 - Vapor to III -	24" dia.
FT- 3	Impure Condensate from III and/or FT-2 - Vapor to IV -	24" dia.
FT- 4	Impure Condensate from IV and/or FT-3 - Vapor to V -	24" dia.
FT- 5	Impure Condensate from V and/or FT-4 - Vapor to VI -	24" dia.
FT- 6	Impure Condensate from VI and/or FT-5 - Vapor to VII -	24" dia.
FT- 7	Impure Condensate from VII and/or FT-6 - Vapor to VIII -	24" dia.
FT- 8	Impure Condensate from VIII and/or FT-7 - Vapor to IX -	24" dia.
FT- 9	Impure Condensate from IX and/or FT-8 - Vapor to X -	24" dia.
FT-10	Impure Condensate from X and/or FT-9 - Vapor to XI -	24" dia.
FT-11	Impure Condensate from XI and/or FT-10 - Vapor to XII -	24" dia.

As mentioned earlier, these flash tanks were later used routinely for condensate heat recovery when the Series 200 heat exchangers were removed from service. After an appropriate increase in the condensate piping sizes the flash tanks worked quite efficiently.

d. Deaerator - Decarbonator — The original function of the deaerator was to remove dissolved gases, mainly oxygen, from the seawater feed so that its corrosion potential would be reduced. The oxygen concentration of 7 to 8 ppm in the raw seawater feed was to be reduced to 0.5 ppm. With the abandonment of the $Mg(OH)_2$ sludge scale prevention technique and adoption of the pH control technique, the deaerator had an additional duty of decarbonating the seawater feed of the released CO_2 amounting to about 90 ppm (total). The design outlet concentration of CO_2 in product water was specified as 2 ppm. The OSW document "Specifications 195" gave the following design specifications for the deaerator:

- seawater feed — 440,000 lbs/hr @ 135°F
- stripping steam rate — 410 lb/hr from Effect XI
- diameter of the deaerator tower — 6 feet
- total height of the tower — 43 feet
- maximum design temperature and pressure — 150°F and full vacuum; tower shell material — carbon steel; coating on inside of tower, packing support, flanges and nozzles — 1/8 inch thick epoxy lining or approved equal

- seawater spray nozzle — Saran plastic, Spraco full cone No. 18DF or equal
- mixer for $Mg(OH)_2$ slurry — Lightnin model 110-SE-5, or equal, with monel shaft and propeller designed for 80% motor loading at normal conditions. The motor was to be 5 H. P. 440 volt, 3-phase, 60 cycle, Allis-Chalmers Super-Seal or equal.

A sketch of the deaerator is shown in Figure 2. 6.

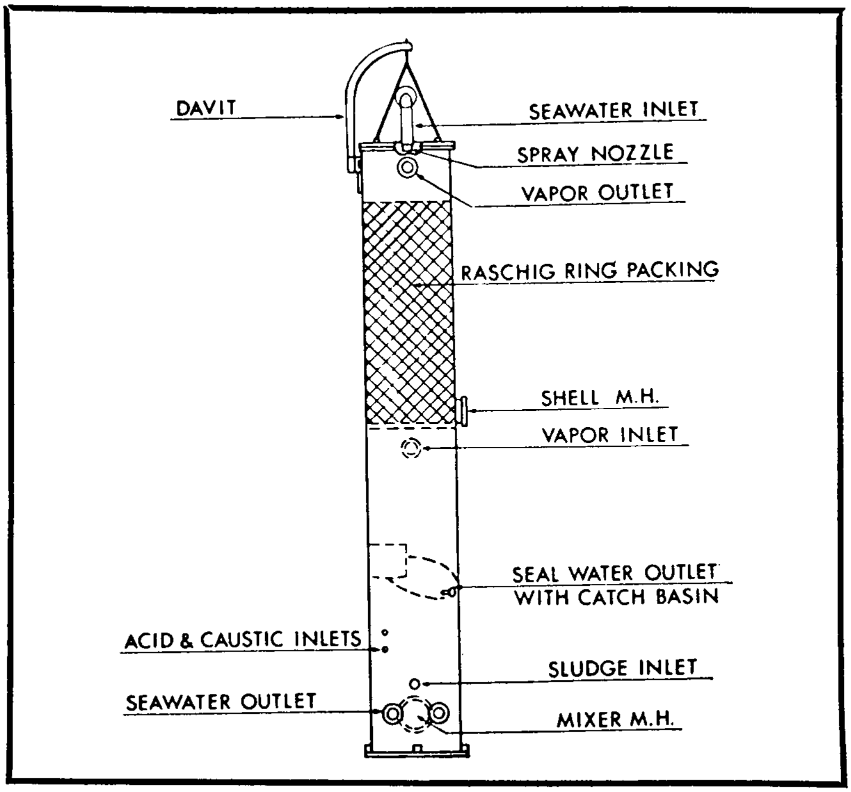


Figure 2.6. Deaerator-Decarbonator

Chicago Bridge & Iron Co. followed these specifications in the design and construction of the deaerator (Drawing Numbers 79 and 80) and they also specified the packed height of the tower as 16 feet with 3 inch ceramic raschig rings as the packing. Operation of the deaerator assumed double significance with the CO_2 removal function imposed on it. Alkaline scale ($CaCO_3$, $Mg(OH)_2$) prevention was entirely dependent on efficient CO_2 desorption from the seawater feed. In addition, severe corrosion and capacity decrease result from CO_2. An exhaustive study was therefore undertaken by Stearns-Roger to determine the optimum operating conditions for the deaerator (Second Annual Report, i.e., OSW R&D No. 100, p. III-4). This study was continued right up to the final Development Run 16. The following conclusions were drawn from this work:

- Proper operation of the vacuum maintenance system is necessary to maintain a low tower pressure to allow for flashing of the feed. A separate vacuum system for the deaerator-decarbonator should be considered.

- The preheat circuit prior to the deaerator-decarbonator must be properly designed to provide the necessary heating of the seawater feed for flashing to occur at the inlet of the deaerator-decarbonator. Feeding the process from the return leg of Heat Exchanger 312 is an aid to tower performance.

- The portion of the process under a vacuum must be leak-tight. Leaks are detrimental to deaerator-decarbonator performance in two ways:
    - Leaks increase the vapor load that must be handled by the vacuum system and can overload the system such that the tower pressure rises.
    - Leaks introduce carbon dioxide and oxygen into the system and can decrease the partial pressure difference between the stripping steam and the seawater in the packed section.

- Optimum stripping steam rate for normal seawater is 1 pound per 1000 pounds of seawater.

- Minimum degree of seawater feed superheat (relative to deaerator pressure) is 0.5 to 1.0°F.

- Vent steam rate should be such as to allow for flashing feed condition.

- 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. Under these conditions, caustic neutralization can be eliminated.

- Catalyzed sodium sulfite is successful in scavenging residual oxygen from the deaerator brine. However, removal of the residual oxygen does not significantly affect indirectly measured plant corrosion rates.

- An acid mix chamber is not required to obtain satisfactory mixing of the acid (sulfuric) and the seawater feed to the deaerator.

- The deaerator effluent alkalinity up to 30 ppm equivalent $CaCO_3$ does not cause scaling or venting problems.

Two important OSW R&D reports useful in this area are:

- No. 158 — "Desorption of CO_2 and O_2 from Seawater" by M. W. Kellogg, and

- No. 314 — "Deaerators for Desalination Plants" by Oak Ridge National Laboratory.

e. <u>Vacuum System</u> — A vacuum system is required to maintain pressures lower than atmospheric in the final condenser and all effects having brine saturation temperatures lower than about 212°F. Plant capacity depends on the temperature driving force T_1-T_L where T_1 is the first effect steam chest temperature and T_L is the brine boiling point in the last effect. The brine saturation temperatures in all the effects are determined by the absolute pressures maintained in each effect vapor head. When the last effect is maintained at the lowest possible vacuum, maximum driving force T_1-T_L is obtained and hence maximum capacity for a given evaporator with a fixed T_1. Noncondensible gases from sub-atmospheric pressure effects have to be removed with the help of the vacuum system. Efficient operation of the vacuum producing equipment is therefore very essential for smooth functioning of a multiple effect desalination plant.

The original vacuum system at Freeport included a barometric condenser, a vacuum pump, and a steam jet ejector to operate as a spare and/or auxiliary vacuum source to the vacuum pump ("Specifications Number 195", p. IV-64). Noncondensibles, mainly air and CO_2, and water vapor from the deaerator and the final condenser were to be cooled by raw seawater in the barometric condenser. The vacuum pump was to remove the noncondensibles saturated with water vapor at the cooling water temperature.

The barometric condenser was to be of the disc-flow type having a tall pipe 35 feet long. The vacuum pump, of the reciprocating dry type, was to be designed for continuous full load duty using an electric motor drive. The number of stages to be used in the steam jet ejector was left to the discretion of the constructor. The following operating conditions were specified in the original design:

- Barometric Condenser

  Flow rates: CO_2 — 50 lb/hr at 133°F
  Air — 22.5 lb/hr at 133°F
  H_2O vapor — 470 lb/hr at 133°F
  H_2O vapor — 220 lb/hr at 115°F

  Duty: 745,000 Btu/hr

- Vacuum Pump

  Noncondensibles: CO_2 — 50 lb/hr (430 SCFH)
  Air — 22.5 lb/hr (295 SCFH)

  Water vapor: Corresponding to saturation at the outlet temperature of gas from the barometric condenser.

  Cooling water: 100 GPM based on a 15°F rise in temperature. Maximum design temperature of inlet seawater was to be 88°F.

  Steam conditions: 160 psig and 530°F for operation of the steam jet ejectors. The steam pressure might fluctuate by ± 10%.

An absolute pressure of 3 inches of mercury was to be maintained in the final twelfth effect.

f. Clarifier-Thickener and Classifier — In the original design, the purpose of the clarifier-thickener was to recover $Mg(OH)_2$ crystals from the brine blowdown of the plant. This unit was to be designed for the blowdown rate of 120,000 lb/hr at 120°F and containing 13.8% total dissolved solids in addition to 4% by weight of $Mg(OH)_2$ in suspension. The clarifier-thickener bottom discharge was to contain 15 to 18% $Mg(OH)_2$ by weight. The approximate dimensions of this unit were given as 30 feet diameter by 14 feet liquid side depth with 1 foot freeboard. With the abandonment of the $Mg(OH)_2$ scale prevention technique this unit was put out of service until the major plant modification to 17 effects when it was used as a silt settling tank.

The classifier was to remove mill scale and other large and heavy particles from the recycle magnesium hydroxide slurry, plus incidental classification of $Mg(OH)_2$ crystals according to particle size.

g. Desuperheater — The prime steam supplied by Dow Chemical Company was to be at 160 psig and 530°F so that it was superheated by about 159°F. The steam condensing temperature required in the first effect chest was 261°F. The extent of superheat was much more (i.e., 269°F) with respect to the first effect steam chest conditions. The superheated steam has very low coefficient of heat transfer so that it is inefficient as a heat transfer agent. Saturated steam, on the other hand, gives very high steamside coefficients of heat transfer. The desuperheater's function, therefore, is to remove the superheat from the DOW steam and supply saturated steam to the first effect steam chest. No specifications were found for this unit in the "Specifications Number 195" document. However, the desuperheating was to be accomplished by spraying condensate from the first effect.

h. Pumps — There were seven items of services to be performed by the pumps at the Freeport Plant according to the original design:

- condensing water
- seawater feed
- blowdown
- condensate
- flashed condensate
- product water
- seal water

The condensing water pumps, designated as P-1 and P-2, were to pump raw seawater from an intake pit to the final condenser, HX-312. These pumps were to be of the vertical propeller type. Shaft guide bearings were to be of rubber and were to be water lubricated. The pumps were to be of all iron construction except for the propeller which was to be fabricated from Ni-Resist. Flax was to be used as the packing material. Normal flow was stated as 3500 GPM whereas the design value was specified as 4000 GPM for a differential head of 22 feet. Discharge line size was to be 16 inches.

The seawater feed pumps, P-3, P-4, and P-5, were to pump seawater to the first effect. One pump, P-3, would take suction from the discharge of the condensing water pump and discharge into the deaerator. The other two pumps, P-4 and P-5 in parallel, would pump the seawater feed from the deaerator through a series of heat exchangers to Effect I. $Mg(OH)_2$ seeds were to be introduced into the deaerator sump to produce a 1% slurry of $Mg(OH)_2$ which was then the suction of P-4 and P-5. The following design conditions were stipulated for these pumps:

- Continuous full load duty.
- Maximum diameter impellers were not to be used. Their design was to allow a later increase in capacity of at least 15% at the design head.
- All pumps were to be fitted with shaft sleeves.
- Stuffing box seals with lantern rings were to be furnished on all pumps. Fresh seawater was to be used as the sealing fluid. The stuffing box was to be designed so that a mechanical seal could be added at a later date.
- These pumps should be end suction, top vertical discharge. Close coupled pumps were not acceptable.
- The pumps were to be fabricated from cast iron and steel with the exception of those items specifically mentioned in the Seawater Feed Pump Summary sheet given below as Table 2.9.

Table 2.9

Seawater Feed Pump Summary Sheet

Pump No.	P-3	P-4, P-5*
1. Process Conditions		
Fluid pumped	Fresh seawater	Fresh seawater with 1% $Mg(OH)_2$ slurry
Fluid temperature - ^{o}F	88 maximum	134
GPM (normal)	955	855
GPM (design)	1150	1050
Differential head (ft.)	55	275
Suct.-Disch. line size	8" - 8"	8" - 8"
Pressure over suction	Atmos.	5.2" Hg. Abs.
NPSH (available)	Flooded	3'
2. Materials of Construction		
Impeller	IB Ni Resist	IB Ni Resist
Wear Rings	Monel	Monel
Lantern Ring	Monel	Monel
Shaft Sleeve	Monel	Monel
Packing	Flax	Flax
Gland nuts	Monel	Monel

* NOTE: It is expected that under certain conditions, P-4 and P-5 will be operated in series. Therefore, the pump casing and other parts in contact with the fluid pumped shall be designed to withstand the discharge of P-4 as the suction to P-5.

The blowdown pumps, P-11 through P-22, were required for transferring concentrated liquor (blowdown) from one evaporator effect to the next. The general design conditions listed earlier for the seawater feed pumps were also applicable to these pumps. Table 2.10 gives the flow conditions for which the blowdown pumps were to be designed.

Thirteen condensate pumps, designated P-31 through P-43, were required to remove condensate (product water) from the steam chest of each effect, the 300 Series heat exchangers and the final condenser. The pumps were to discharge into a header servicing the shell side of the 200 Series exchangers which heat the seawater feed. The general requirements of these pumps were the same as for the blowdown pumps. However, these pumps were to handle potable water and hence their design was to be such that no oil or other outside contaminants could get into the product water. These pumps were to be of all iron construction. Flow conditions for these pumps are presented in Table 2.11.

Table 2.10

Blowdown Pump Summary

Pump for Effect No.	I	II	III	IV
Pump No.	P-11	P-12	P-13	P-14
Forward Feed Conditions				
lb/hr.	411,400	382,700	354,200	325,800
GPM (Normal)	850	790	725	660
GPM (Design)	1000	910	840	760
Sp. Gr. at Opr. Temp.*	0.968	0.973	0.980	0.987
Viscosity - CP. *	0.30	0.30	0.30	0.32
Dissolved Solids - Wt.%	3.24	4.02	4.35	4.73
Suspended Solids - Wt.%	1.07	1.15	1.24	1.35
Design Diff. Head. - Ft.	45	45	45	45
Common Conditions				
Temperature - °F	251	243	235	226
NPSH Available	3.0	3.0	3.0	3.0
Suct.-Disch. Line Size	12-8	12-8	12-8	12-8
Vapor Head Press - " Hg.abs.	60.73	52.70	45.56	38.51
Min. Static Head Above Suct.	4.5	4.5	4.5	4.5

*Does not consider $Mg(ON)_2$ slurry.

BLOWDOWN PUMP SUMMARY (SHEET - NO. 2)

Pump for Effect No.	V	VI	VII	VIII
Pump No.	P-15	P-16	P-17	P-18
Forward Feed Conditions				
lb/hr.	297,800	270,300	243,400	216,100
GPM (Normal)	600	550	480	425
GPM (Design)	700	640	550	490
Sp. Gr. at Opr. Temp. *	0.994	1.000	1.010	1.020
Viscosity - CP. *	0.34	0.35	0.37	0.39
Dissolved Solids - Wt. %	5.18	5.7	6.34	7.1
Suspended Solids - Wt. %	1.48	1.63	1.81	2.04
Design Diff. Head. - Ft.	45	45	45	45
Common Conditions				
Temperature - °F	216	208	198	185
NPSH Available	3.0	3.0	3.0	3.0
Suct.-Disch. Line Size	12-8	12-8	12-8	12-8
Vapor Head Press - " Hg.abs.	31.70	26.53	21.59	16.34
Min. Static Head Above Suct.	4.5	4.5	4.5	4.5

Table 2.10
(cont'd)

Pump for Effect No.	IX	X	XI	XII
Pump No.	P-19	P-20	P-21	P-22
Forward Feed Conditions				
lb/hr.	188,900	161,900	134,900	114,800
GPM (Normal)	365	310	250	210
GPM (Design)	420	360	290	250
Sp. Gr. at Opr. Temp. *	1.032	1.048	1.069	1.087
Viscosity - CP. *	0.43	0.52	0.62	0.78
Dissolved Solids - Wt. %	8.1	9.5	11.4	13.4
Suspended Solids - Wt. %	2.33	2.72	3.26	3.83
Design Diff. Head. - Ft.	45	45	45	45
Common Conditions				
Temp - °F	172	157	138	119
NPSH Avail.	3.0	3.0	3.0	3.0
Suct. - Disch. Line Size	12-8	12-8	12-8	12-8
Vapor Head Press - " Hg.abs.	12.20	8.35	5.17	2.99
Min. Static Head Above Suct.	4.5	4.5	4.5	4.5

Table 2.11

Condensate Pump Summary Sheet

Conditions Common to All Pumps

GPM (Normal) - varies from 47 to 61 GPM

GPM (Design) - 65 GPM

Head, ft. - 70 (differential)

NPSH (available) - 5 ft.

Fluid pumped - condensate

Suction - Discharge line size 3" - 2"

Conditions Specific to Each Pump

Pump No.	P-31	P-32	P-33	P-34	P-35	P-36	P-37	P-38	P-39	P-40	P-41	P-42	P-43
Temp. - °F	261	250	242	234	225	215	206	196	183	170	154	135	115
Sp. Gr.	0.94	0.94	0.95	0.95	0.95	0.96	0.96	0.97	0.97	0.98	0.98	0.99	0.99
Viscosity - Cp.	0.21	0.22	0.23	0.24	0.26	0.28	0.30	0.31	0.34	0.37	0.42	0.49	0.59

Pump P-44 was provided in the original design for transferring condensate from the flash system to the split tank. The flash system was to be in operation intermittently, such as during operation at greater than design production rate, etc. The amount of condensate in the flash system would vary and therefore the flow to the pump would also vary considerably depending upon the immediate operating conditions.

Design flow conditions were stated for this pump as follows:

condensate temperature — 134°F
flash condensate flow — 60 to 360 GPM (normal); 400 GPM design.
differential head — 60 feet
suction-discharge line size — 6" - 4"
pressure over suction — 5.2" Hg abs.
N.P.S.H. available — 3 feet

Product water pumps P-50 and P-51 were to transfer product water from the split tank to the two users — Dow Chemical Company and the city of Freeport. Normally the pumps were to operate at the same time but they were to be sized such that either of the two could transfer the water requirements of both users. Flow conditions for these pumps were identical and given as:

- fluid temperature — 100°F;
- normal flow — 425 GPM;
- design flow — 750 GPM;
- head — 190 feet normal, 200 feet design;
- suction-discharge line size — 8"-8"; and
- suction - flooded.

Pump P-47 was specified in the original Badger design to act as a separate fresh seawater source for sealing certain pump glands, level indicator legs, etc. The general design requirements were the same as for the previously mentioned pumps. Flow conditions were given as 135°F (maximum) fresh seawater, 50 GPM design, head 175 feet with suction pressure as 10 feet minimum. The pump was to be fabricated from cast iron and steel with the exception of the following items:

- impeller — type IB Ni-Resist;
- wear rings, lantern ring, shaft sleeve, gland bolts — monel; and
- packing — flax.

2. Selection of Materials

The materials of construction for various types of equipment in the Freeport Plant were originally selected on the basis of experience in chemical industry in general but with the major emphasis on keeping the capital costs down. The main problem with selection of cheap materials of construction in a desalting plant, is of course, the extreme corrosiveness of hot seawater. High fluid velocities and turbulence complicate the problem further. Since one of the primary objectives of the Freeport Plant was the testing of cheap plant construction materials under large-scale operating conditions the original design incorporated a variety of materials in the numerous items of equipment utilized in the plant. The following excerpt from the paper[1] presented by Chirico and Dunn of Chicago Bridge & Iron Company explains the overall plan of materials selection:

> "The Freeport Plant is almost totally constructed of carbon steel materials except for those portions exposed to the raw seawater. The latter equipment has been fabricated from cupro-nickel material. Magnesium anodes are installed to provide cathodic protection in critical sections where dissimilarity of metals exists. Although seawater is alkaline, the removal of occluded and dissolved gases, mostly air, is necessary to permit the use of inexpensive material of construction. Tubing material in the evaporator effects has been alternated, using both ferrous and nonferrous metals under different temperature and blowdown concentration conditions. It is anticipated that a sufficient history on corrosion data will be obtained to provide for optimum tubing material selection in future installations. Test

---

(1) "Seawater Conversion Plant — Freeport, Texas", by Chirico and Dunn of CB&I, presented at ASCE Convention, Phoenix, Arizona (1961).

> spools, furnished by the International Nickel Company, have been installed throughout the entire system, including the vapor, distillate and blowdown liquor lines. All seawater feed passes through the deaerator prior to entering any of the evaporator effects. The writers are confident that carbon steel materials will prove to be durable as utilized in the design."

A brief listing of the materials of construction utilized for various parts of all the major equipment items of the original Freeport Plant is given in Table 2.12.

F. LEVEL OF PROCESS CONTROL

1. Instrumentation

The purpose of instrumentation of the Freeport Plant has been briefly stated in "Specifications Number 195", page IV-195 as "... for operating control of the plant and for the extraction of test data on this partially experimental plant." On the same page it has been further stated that: "Maximum emphasis has been made on uninterrupted plant operation, system reliability, convenient operation, and accuracy of measurement..." From a functional point of view the following services are rendered by plant instrumentation:

- Measurement of process variables to indicate the state of operation of the plant and quality of product.
- Control of variables within close limits to optimize operations and thus minimize water costs.
- Indication (transmission for remote observation and control) and recording for comparison and corrective action.
- Obtaining accurate heat and material balance data.
- Smooth functioning of plant at the desired and preset conditions; protection of equipment; and, finally, some basic instrumentation is essential for the elimination of constant operator attention.

Table 2.12

Materials of Construction Utilized in the Original Freeport Plant

Equipment	Materials of Construction
Evaporator effects	
• bodies	Low carbon steel A-285-C-FBX
• top water box lining	None initially
• tubesheet	Low carbon steel
• heating element tubes	Low carbon steel, admiralty, aluminum brass, 90-10 cupro-nickel with not less than 1.25% iron, corten
Preheat exchangers	
• shell	A-285, Grade 'C' Flange quality steel
• tubes	Carbon steel, 90-10 cupro-nickel, admiralty, aluminum brass
• pass partitions	Carbon steel HX-212, 213, 214, 311 and 312 to have muntz metal or cupro-nickel cladding on tube sheets and water boxes.
Final condenser, HX-312	
• shell	Carbon steel
• tubes	Aluminum brass
• tube sheets and water boxes	Hortonclad A-265 A-285-C-FBX Backing 1/8" 70-30 CuNi clad
Deaerator	
• shell	Carbon steel
• spray nozzle	Saran plastic
• packing	Ceramic raschig rings The vessel, packing support, nozzles and flanges to be coated with 1/8" thick Epoxy lining.
Pumps	
• condensing water pumps	All iron construction except IB Ni-Resist impeller.
• seawater feed pump	Impeller-Ni-Resist; wear ring, lantern ring, gland nuts and shaft sleeve of monel.
• blowdown	Cast iron and steel; wear ring, etc., monel.
• condensate	Same as above
• flashed condensate	Same as above
• product water	Same as above
• seal water	Same as above
Piping	
• seawater feed piping	Class A, A.S.T.M. A-53 Grade B
• condensate	Class B, black butt-welded, or seamless carbon steel ("Specifications -195", p. IV-155)
• vapor or steam	A-283-C, A-53, A-7
• blowdown	Class A
• sulfuric acid, caustic 50%	Class D, black butt-welded
• seawater underground	Class F, A.S.T.M. 58, Grade A with coal tar enamel coating.

The instrumentation flowsheet for the original Freeport design by W. L. Badger Associates is shown in Drawing Number 511. The general instrument schedule for the original plant is given in Table 2.13. A list of panel mounted instruments is given in Table 2.14. High accuracy instrumentation of selected effects was carried out in 1965 to enable accurate measurement of two phase pressure drops and concomitant temperature drops in the heating element tubes as predicted by Dukler's theory (see "Heat Transfer History", Chapter 4). Figure 2.4 shows the typical instrumentation of an evaporator effect. Actual measurements did confirm, especially in the low temperature effects, the presence of significant temperature drops associated with the brine-vapor pressure drops.

2. Process Control

The multiple-effect falling-film desalination process, as employed at the Freeport Plant, is essentially simple to control from the standpoint of maintaining the quantity and quality of production. The overall process operation at Freeport is shown in block diagram form in Figure 2.7. Raw seawater feed is first regeneratively preheated in a series of heat exchangers so that upon acidification and flow through the deaerator-decarbonator most of its bicarbonate alkalinity is decomposed and CO_2 is vented through the vacuum system. The deaerated and decarbonated seawater feed is further preheated to bring it close to the saturation temperature of Effect I into which it is finally fed. Prime steam from an external source is supplied to the steam chest of this effect to drive the process. Vapor, brine and condensate flow through the plant from the high temperature end (first effect) to the low temperature end (last effect). The quantity of vapor transferred between effects is approximately the same; the brine quantity diminishes while the condensate quantity continually increases. At the cold end of the plant, the concentrated brine is discarded, vapors from the last effect are condensed using raw seawater, and all the condensate from the plant, which is product water, is pumped to storage. There are three basic fluid paths in the Freeport LTV process — seawater or brine; steam or vapor; and, condensate. These paths, along with their main controlling instrumentation, are given in the self-explanatory block diagrams in Figures 2.8, 2.9, and 2.10.

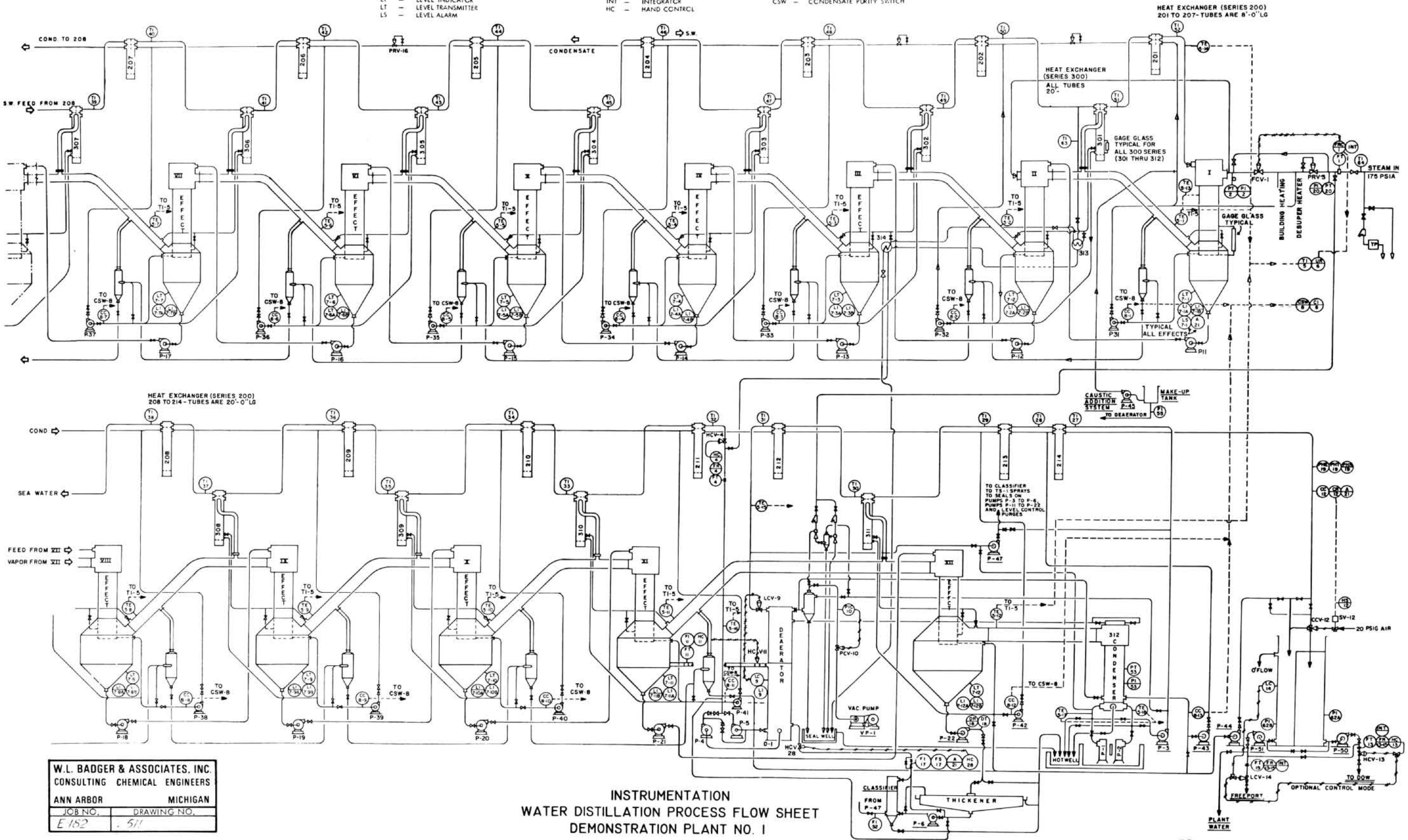
In the instrumentation design of an LTV plant an economic balance has to be struck between the minimum instrumentation costs and minimu operating manpower requirements since product water costs must be kept down to an absolute minimum. Because of the experimental nature of the Freeport

INSTRUMENTATION LEGEND

TE — TEMPERATURE ELEMENT
FR — FLOW RECORDER
CI — CONDUCTIVITY INDICATOR
UR — HEAT TRANSFER APPARENT COEFFICIENT RECORDER
TI — TEMPERATURE INDICATER
LC — LEVEL CONTROLLER
CRy — CONDUCTIVITY RELAY
SV — SOLENOID VALVE
FI — FLOW INDICATOR
LI — LEVEL INDICATOR
LT — LEVEL TRANSMITTER
LS — LEVEL ALARM

PT — PRESSURE TRANSMITTER
PRV — PRESSURE REGULATOR VALVE
PI — PRESSURE INDICATOR
PHT — pH TRANSMITTER
PHE — pH ELEMENT
PHR — pH RECORDER
A — ANNUNCIATOR ALARM
FT — FLOW TRANSMITTER
FRC — FLOW RECORDER - CONTROLLER
INT — INTEGRATOR
HC — HAND CONTROL

FCV — FLOW CONTROL VALVE
HCV — HAND CONTROL VALVE
HS — HAND SWITCH
CC — CONDUCTIVITY CELL
CCV — CONDUCTIVITY CONTROL VALVE
PIC — PRESSURE INDICATING CONTROL
DT — DENSITY TRANSMITTER
DR — DENSITY RECORDER
FS — FLOW SWITCH
CSW — CONDENSATE PURITY SWITCH



INSTRUMENTATION
WATER DISTILLATION PROCESS FLOW SHEET
DEMONSTRATION PLANT NO. 1

FREEPORT-TEXAS

W.L. BADGER & ASSOCIATES, INC.	
CONSULTING	CHEMICAL ENGINEERS
ANN ARBOR	MICHIGAN
JOB NO.	DRAWING NO.
E 182	511

Table 2.13

General Instrument Schedule — No. 1 Demonstration Plant

GENERAL INSTRUMENT SCHEDULE - NO. 1 DEMONSTRATION PLANT

Item	Service	Equivalent to Mfg. - Mod.	Input Span	Output/ Scale
PT-1	Steam Feed	Foxboro 13 A	100" W.C.	3-15 psi
PT-2	No 1 Chest	Foxboro 42	60 psig	3-15 psig
PT-4	Sea water feed	Foxboro 13 A	100" W.C.	3-15 psi
TE- 5-1 to 18	Vapor-Liquor	Foxboro Dynatherm	0-350°F	--
LT-7-1 to 12	Evaporator levels	Taylor 200 TD 111	200" W.C.	3-15 psi
CC-8-1 to 13	Condensate conductivity	Industrial Inst. CEL-101	0-300 ppm (NaCl)	--
LC-9	Deaerator	Foxboro 13 FA	0-200"	3-15 psi
PIC-10	Pump Vacuum	Taylor 162 RA 135	0-8" Hg abs.	0-8
PT-11	Deaerator Vapor	Republic D/P	0-2" W.C.	3-15 psi
CC-12	Dow water diversion	Industrial Inst. CEL-101	0-300 ppm	--
SV-12	Dow water diversion	Automatic Switch Co. No. 8314	110 VAC	--
FT-13	Dow water	Foxboro 13 A	100" W.C.	3-15 psi
LC-14	Freeport water	Foxboro 13 FA	100" W.C.	3-15 psi
FT-15	Freeport water	Foxboro 13	100" W.C.	3-15 psi
FI-17	Sludge feed	Fischer-Porter 10 A 1152	5-50 gpm	5-50
DT-18	Brine output (Specific grav)	Fisher-Gov. Densitrol	1.00-1.10	1.00-1.10
pHE-19	Product pH	Beckman	3-11	--
PT-20	Dow steam	Foxboro 42	0-200 psig	3-15 psi
HC-28	Clarifier Underflow	Foxboro 52	Manual	3-15 psi
PT-53	Condenser vacuum	Foxboro 42	0-30" hg vac.	3-15 psi
FI-52	Water to classifier	Fischer-Porter	0.1-10 gpm	0.1-10
FI-55	Caustic to Deaerator	Fischer-Porter	0.2-2gph	0.2-2

GENERAL INSTRUMENT SCHEDULE - NO. 1 DEMONSTRATION PLANT (Cont.)

Item	Element Material	Connections	Remarks
PT-1	St. Stl.	½" NPT	Include bypass piping
PT-2	Bronze	¼"NPT	
FT-4	Monel	½"NPT	Include bypass piping and purges
TE-5-1	Brass Wells	1" NPT	See separate tabulation
LT-7-1	Brass	¼" NPT	Include purges
CC-8-1	Brass	1¼" NPT	Carbon electrode
LC-9	Monel	3"-150 lb. ASA	In conjunction with a reduced span valve positioner
PIC-10	Brass	½" NPT	
FT-11	Carb. Stl.	½" NPT	
CC-12	Brass	1¼" NPT	Carbon Electrode
SV-12	Brass	¼" NPT	3-way
FT-13	St.Stl.	½" NPT	
LC-14	St.Stl.	3"-150 lb.·ASA	In conjunction with a reduced span valve positioner.
FT-15	St. Stl.	½" NPT	
FI-17	Monel (Carb. Stl.)	2" x 2½"-150 lb. ASA	Low flow switch to A-21
DT-18	C.I.	2"-125 lb. ASA	2 nd. pen of FR-4.
pHE-19	St. Stl.	½" NPT	Temp range 50-150°F
PT-20	Bronze	¼" NPT	
HC-28	Bronze	1/8" NPT	Field loading station
PT-53	Bronze	¼" NPT	
FT-52	C.I.	1"-150 lb ASA	Rotameter
FI-55	Monel	½" NPT	Rotameter

Table 2.14

Panel Instruments for No. 1 Demonstration Plant

Item	Service	Equivalent to Mfg. - Mod.	Input	Chart Scale	Units	Remarks
FRC-1	Steam Feed	Foxboro Mod. 40	3-15 psig	0-40 sq. rt.	1000 pph	Auto-Manual Station, Integrator
PI -2	No.I Chest	Ashcroft 1377 - 6"	3-15 psig	60	psig	
FR-4	Sea Water Feed	Foxboro Mod. 40	3-15 psig	0-60 sq.rt.	10^4 pph	2nd. pen DR-18 Case mounted manual station (HC-4
TI-5	Vapor Liquor	Foxboro Dynalog Push button	Resistance Bulb	0-350	°F	18 pts.
UR-6	Heat Trasfer Apparent Coef	Dynalog BTU	Resistance Bulb-Electronic	0-800	BTu/hr.°ft^2	Uses 13 of TI-5 pts.
LI-7-1 to 12	Evaporator Levels	Ashcroft 1377 A-6"	3-15 psig	100-0	% of 200"	0= 15 psig - 100 = 3 psig
LS-7-1 to 12	Evaporator Levels	Ashcroft 1079-1165-XED	3-15 psig	0-100	%	Hi and low alarm on A-21 Mounted at rear of panel
CI-8	Product Purity	Foxboro Dynalog	Conductivity Cells	0-300	ppm NaCl	
CSw-8	Product Purity	Foxboro 1099	Conductivity Cells	No. 1 to 15		15 position selector for CI-8
CRy-12	Dow Water Diversion	Ind. Ins. RE-H3	Conductivity Cell	0-300	ppm NaCl	With HS-12 manual switch
FR-13,15	Dow-Free-port Water	Foxboro Mod. 40	3-15 psig	0-60 sq. rt.	10^4 pph	2nd Pen FR-15 Integrators on FR-13 FR-15 Case mounted manual station (HC-13)
pHR-19	Condensate	Foxboro Dynalog	25-0-25 mv.	3-11	pH	Low alarm contacts to A-21 Pneumatic loading station
P-1-20	Dow Steam	Ashcroft 1377-6"	3-15	0-200	psig	
PI - 53	Condenser	Ashcroft 1377-6"	3-15	0-30	in. Hg vacuum	
PI - 54	Inst. air	Ashcroft 1377-6"	0-150	0-150	psig	
FI-11	Deaer. Vapor.	Ashcroft 1377-6"	3-15	0-100 sq.rt.		Manual station (HC-11)
A-21	Panel	Panalarm	120 v. AC	Engraved	3	With flasher,outside horn and acknowledge button
A-22	Panel	Panalarm	120 V.AC	Engraved	3	
Am-23	Sea water Feed	GE AB-18 4 1/2"	3 wire	0-35	amps	Current transformer located in switch cubicle
Am-24	Dow Water	GE AB-18 4 1/2"	3 wire	0-75	amps	Current transformer located in switch cubicle
Am-25	Freeport Water	GE AB-18 4 1/2"	3 wire	0-75	amps	Current transformer located in switch cubicle
Am-26	Thickener	GE AB-18 4 1/2"	3 wire	0-2.5	amps	Current transformer located in switch cubicle

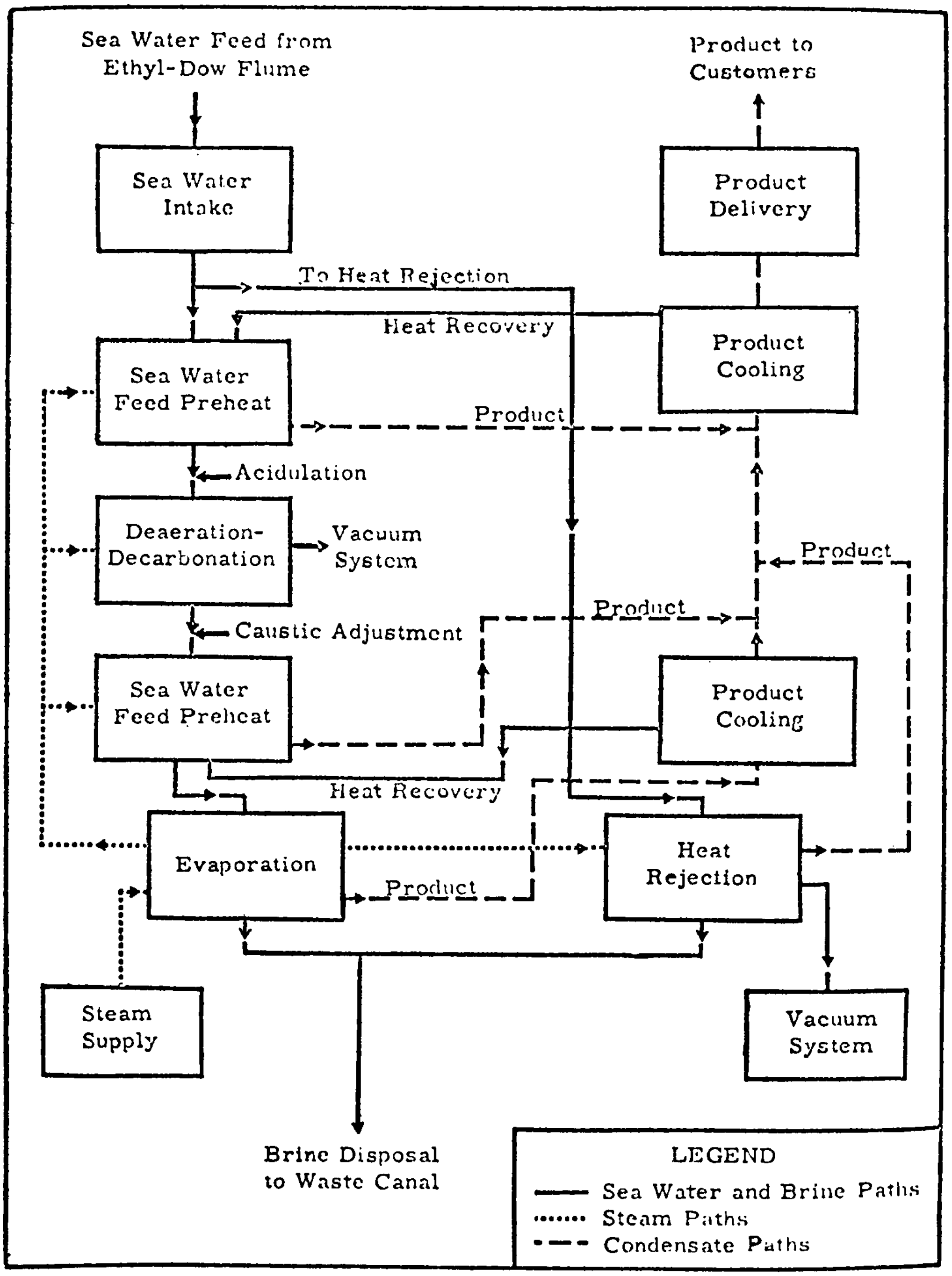


Figure 2.7. Freeport ME-LTV Process Block Diagram

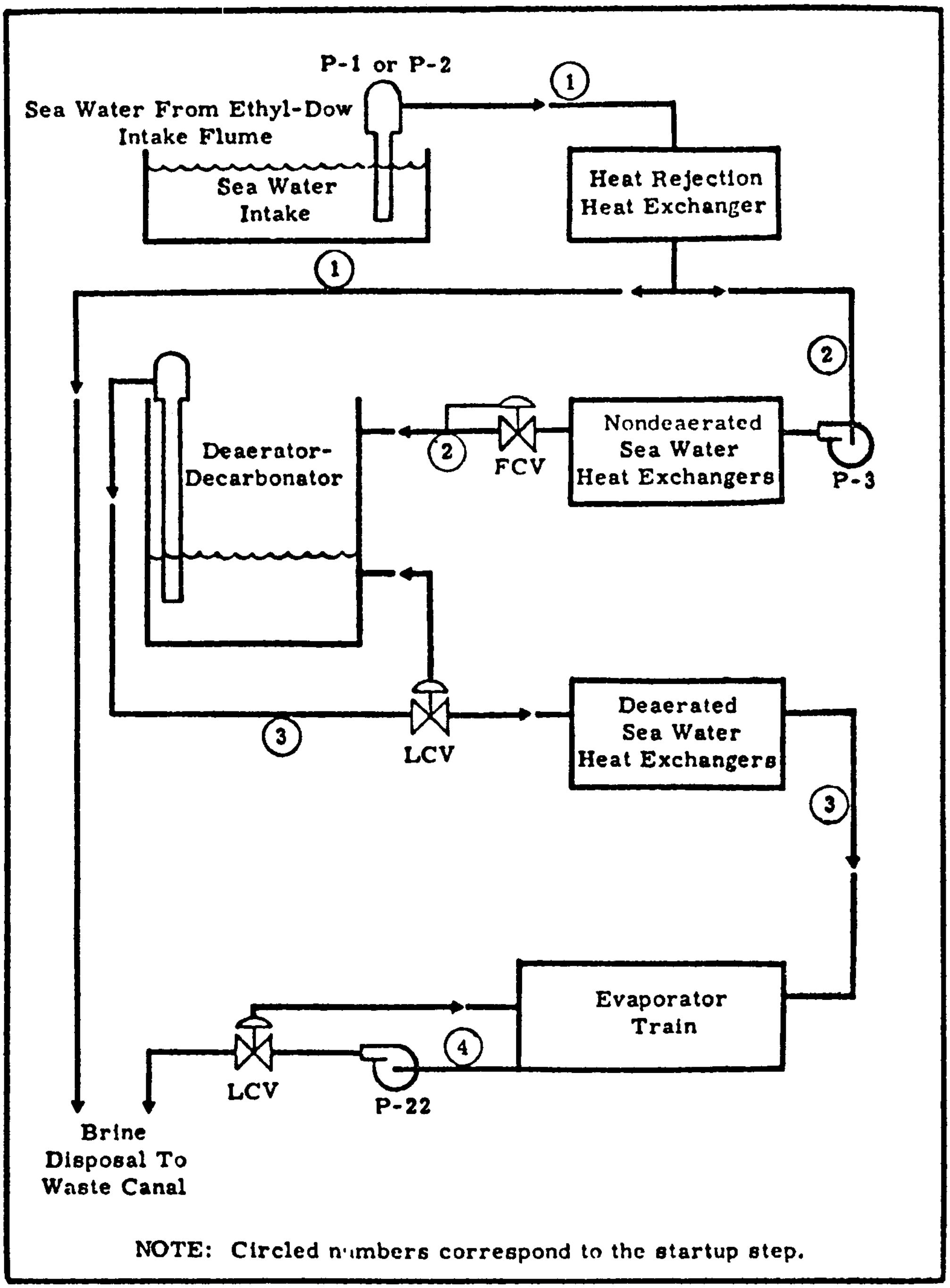


Figure 2.8. Seawater Feed and Brine Cycle, Block Diagram

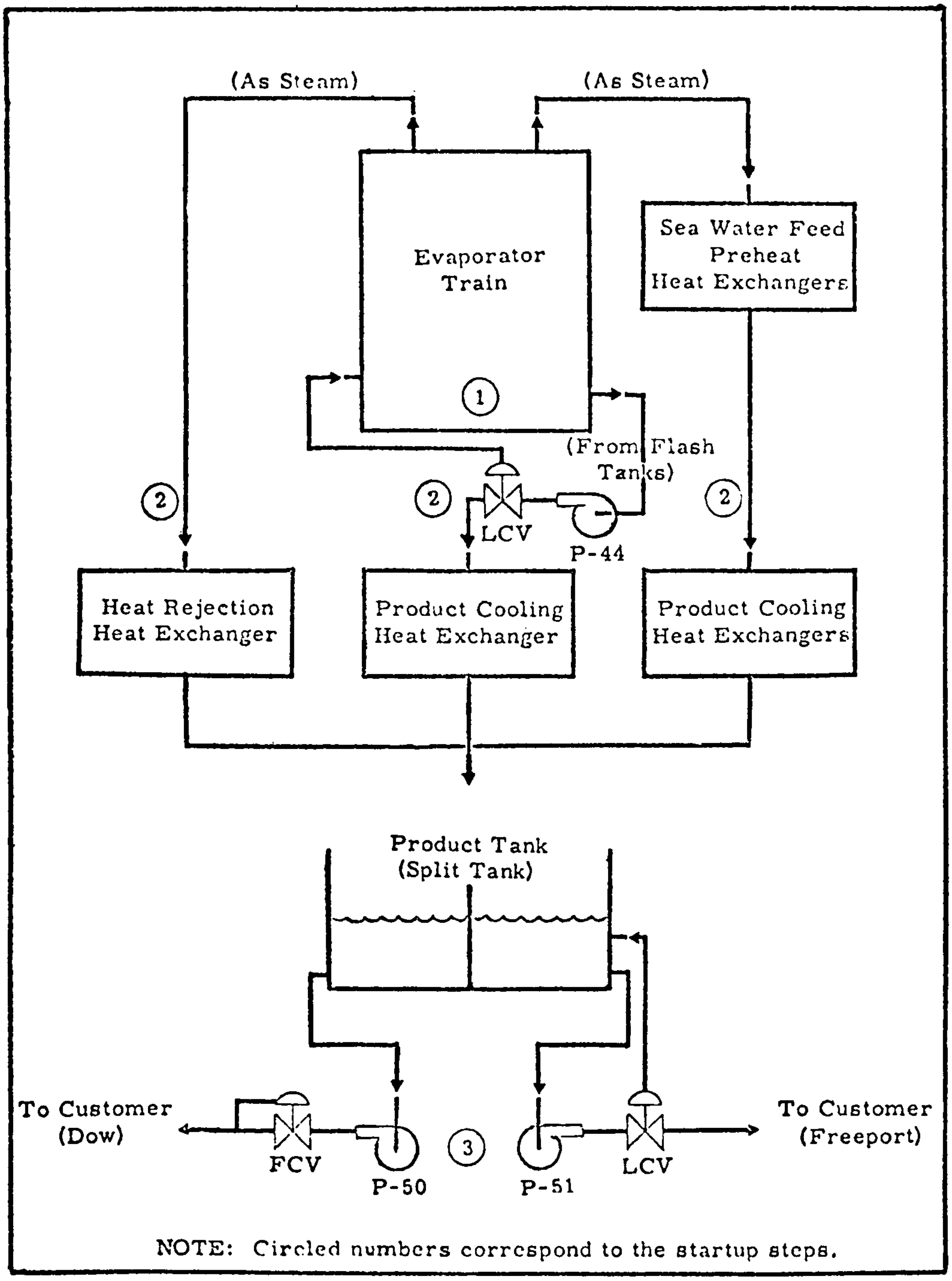


Figure 2.9. Condensate Paths, Block Diagram

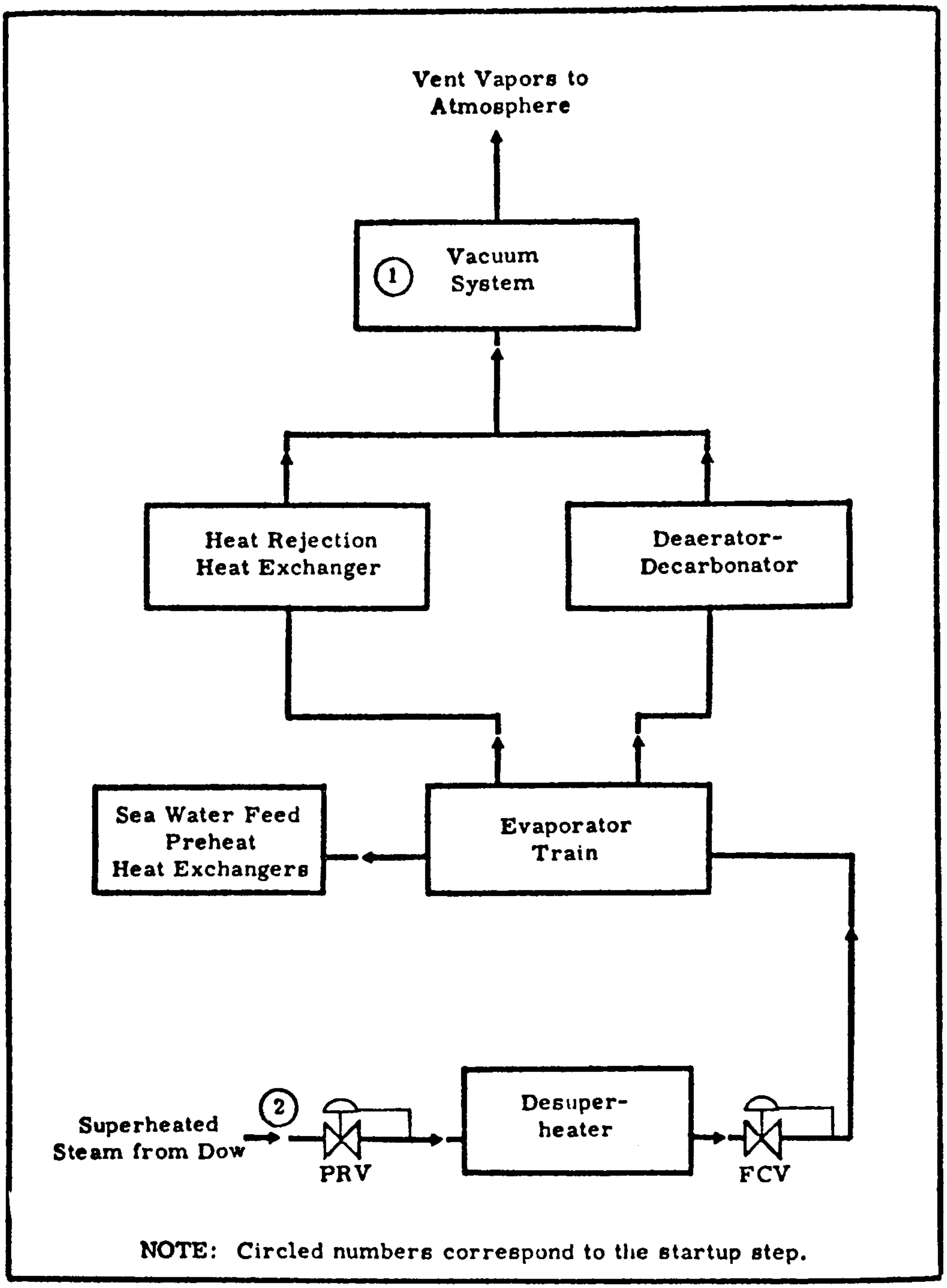


Figure 2.10. Steam Paths, Block Diagram

Plant, the instrumentation and process control were not optimized. Control loops involving the following equipment were important for smooth and efficient operation of the Freeport Plant:

a. Deaerator-Decarbonator

b. Effect I

c. Evaporator train

d. Cold end of the plant

a. Deaerator-Decarbonator Control — The deaerator-decarbonator (DA-DC) was the most heavily instrumented unit at Freeport. A sketch of this unit showing the main control instrumentation is shown in Figure 2.11. The DA-DC has two functions — to remove dissolved oxygen and free CO_2 from the seawater feed to reduce corrosion and alkaline scale problems. Seawater feed is preheated and acidified just ahead of the unit and then sprayed over the packing inside the DA-DC tower. Dissolved O_2 and CO_2 are stripped from the downflowing seawater by means of the stripping steam introduced at the bottom of the tower packing. The deaerated-decarbonated seawater accumulates in the sump which is then pumped to the exchangers for further preheating. Catalyzed sodium sulfite can be added to this stream to remove the residual oxygen while caustic soda can be added to adjust the seawater pH as required. For efficient operation of the DA-DC and the whole plant, the following four controls are essential:

- SWF Control — Two choices are possible in the control of seawater feed to the DA-DC unit. For stable operation of the deaerator only, flow control of brine to the deaerator and level control of the brine out of the unit would be chosen. However, for smooth operation of the total plant it is sometimes more desirable to flow control the brine outflow and level control the brine inflow to the deaerator. The latter was initially chosen for operation by the plant constructors, Chicago Bridge and Iron Company (see Drawing No. 3, "As-Built"). To avoid cavitation damage to brine outflow pumps P-4 and/or P-5, the control system was later changed to level controlled outflow and flow controlled brine inflow as shown in Figure 2.8 (from Fifth Annual Report, page 222, by Stearns-Roger).

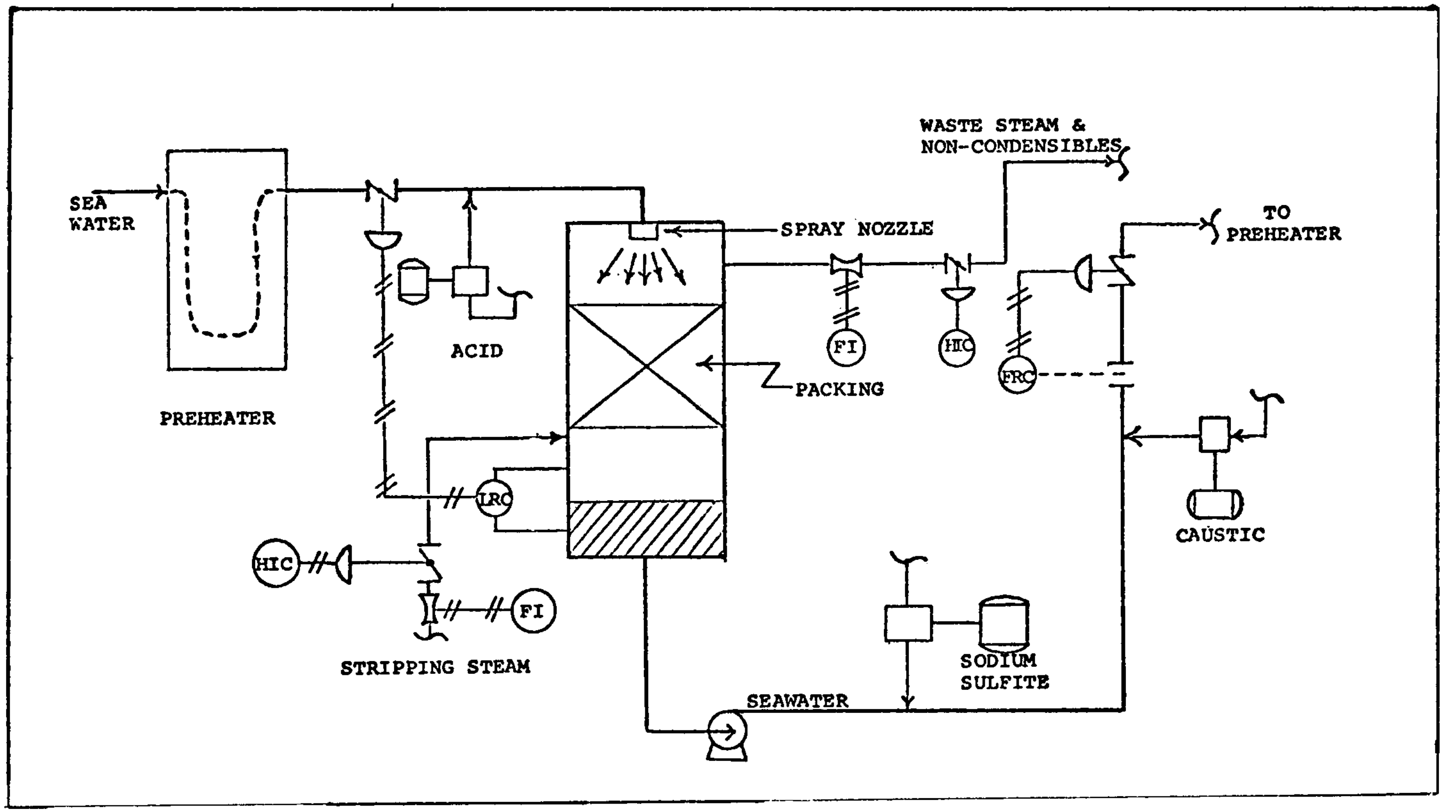


Figure 2.11. Deaerator-Decarbonator Control Instrumentation

- Steam Flows — Steam for stripping the seawater feed of its dissolved gases is taken from a low temperature effect such as the eleventh effect in the original plant. Its flow rate has to be controlled in relation to the seawater feed rate and its alkalinity which is approximately constant. Original design provided for hand control of this stripping steam inflow to the DA-DC tower at about 1 pound per 1000 pounds of SWF.

  The next higher level of instrumentation would require the use of automatic flow ratio controllers if significant fluctuation of SWF rate exists.

  Since flashing of seawater feed in the DA-DC tower is essential for efficient stripping operation, temperature control of the steam flow has been proposed. In this mode of control, the seawater temperature drop across the DA-DC tower would be sensed; a controller would reset the position of the valve in the deaerator vapor outlet line (see Figure 2.11) to control this temperature drop. This system has been used on a hand control basis to maintain the specified degree of flashdown ($1.0 \pm 0.2^{o}F$) in the deaerator at Freeport.[1]

- Acid Addition — For proper decomposition of the bicarbonates in SWF it is essential to control the sulfuric acid addition rate. An excess of acid in SWF could lead to severe corrosion of most steel equipment while scaling can result from inadequate rate. Initially the acid rate was hand controlled to yield seawater inlet pH around 4 to 4.5. Since indicated pH was found to fluctuate considerably the control point was later shifted to seawater outflow from the DA-DC with pH control point being 5.6 to 6.0. An improvement in this control loop would be the replacement of the present manual adjustment of acid pump stroke with an automatic controller. Instrumentation for this conversion was actually purchased in 1969 but installation was deferred due to plant shutdown.

---

(1) Campbell, K. S., "LTV Process Control", a paper presented at the 2nd Conference on Instrumentation for Desalination Plants, San Diego, February, 1969.

- Other DA-DC Controls — Two other control loops around the DA-DC are of interest. Initial design included caustic neutralization loop to bring SWF outflow from the DA-DC to neutral range. During the final development runs it was shown that by controlling the outlet pH between 5.6 to 6.0 caustic addition could be eliminated without scaling, corrosion or venting problems in the first effect (see e.g., Eighth Annual Report, p. 87).

  Sodium sulfite addition to the brine outflow from the deaerator was a hand controlled operation to chemically reduce (scavenge) the residual oxygen to 0.05 - 0.1 ppm. Development Report Number 9 by Stearns-Roger has the following comment on page 14: "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."

b. Control of Effect I — Control of the evaporating section of an ME-FF plant is fairly straight forward. The vacuum system and the heat rejection condenser performance fix the conditions at the cold end of the plant while steam temperature and flow rate fix conditions at the high temperature end for a given seawater feed rate. The intermediate effects assume steady state conditions automatically when no process problems or restrictions exist.

For lowest water costs there is an economic driving force to reduce the SWF rate to the minimum since this leads to savings in pumping power, acid, steam, and other chemicals. Similar driving force exists to raise the first effect temperatures. However, two scaling restraints exist against which the plant operations must be optimized. At the low temperature end of the plant gypsum ($CaSO_4$, $2H_2O$) scaling occurs if brine is so concentrated as to exceed a concentration factor of about 3.0. Anhydrite ($CaSO_4$) scaling at about 268° F with normal seawater limits the first effect high temperature operation without pretreatment.

Once the optimum temperature and related parameters are selected for the first effect a number of control schemes are possible for smooth and efficient operation. A common approach is to bring the steam in on flow control. Since scaling and plant capacity are determined by temperature of brine in the first effect sump, the steam flow to this effect can also be put under sump temperature control as shown in Figure 2.12.

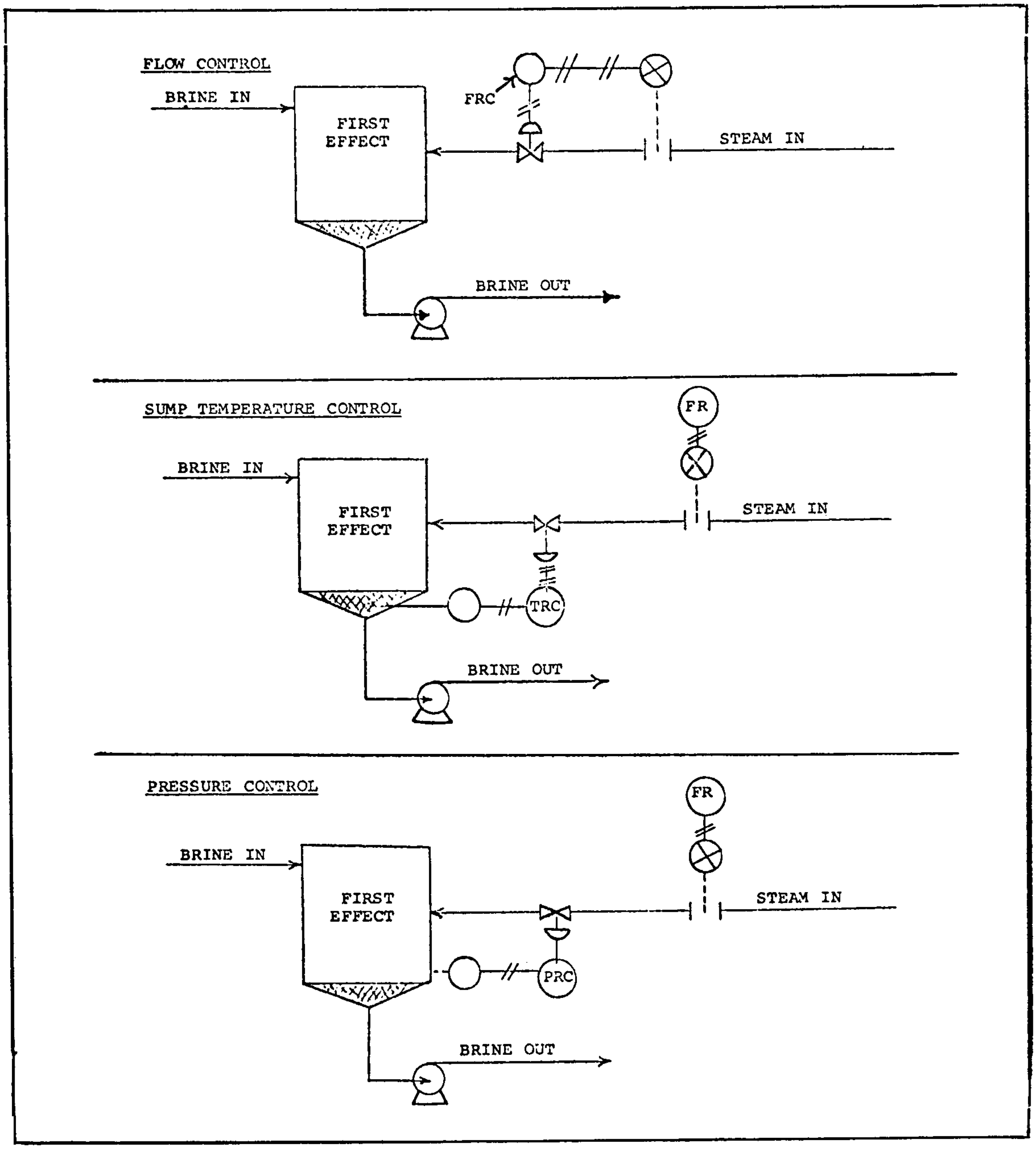


Figure 2.12. Instrumentation Sketches

There are several disadvantages to this control scheme. There is overcorrection due to process lag; if brine flow decreases, scaling may occur as a result of overconcentration, the only remedy being to continually change the temperature set point on the controller as the brine flow fluctuates. Still another problem is that of instrument span, which may be required to be 80°F with an absolute accuracy of ± 0.5°F!

The third method of first effect prime steam control is based on steam chest pressure control (Figure 2.12). The steam rate is adjusted to maintain a constant, preset chest condensing pressure. For a plant operating close to its maximum rate, chest pressure control serves to regulate not only the sump temperature but also the tube temperature, both of which affect scaling potential. At the Freeport Plant this control scheme has been effective in maintaining smooth plant operations close to the maximum plant operating temperature. The chest pressure controller acts quickly and smoothly to adjust the plant operations for changes in heat balance requirements caused by such factors as internal vent adjustments or heavy rains.

This scheme, however, is not entirely satisfactory. The pressure set point on the controller has to be manually changed every time the brine rate is cut for reduced throughputs or for any other reason. The pressure span may be as much as 45 psia to 10 psia and an absolute accuracy of 0.25 psia may be required at the 45 psia level.

All of the three schemes discussed above have a common disadvantage insofar as they maintain the steam flow irrespective of changes in the brine rate. Scaling could result in this scheme when brine rate drops to low values. The best control scheme would involve SWF to steam ratio control with (scaling threshold) temperature override to protect the system from the scaling problems.

c. <u>Evaporator Train Control</u> — As mentioned earlier, the intermediate effects reach steady state conditions by themselves when the first and last effect operations are maintained steady. Therefore, the only instrumentation required on these effects is that needed for usual process monitoring purposes. Intereffect steam flow is self-regulatory; the vent rates are adjusted manually at start-up. Condensate flow is driven by the intereffect pressure and regulated by the size of the condensate transfer lines. Brine transfer in the series forward feed scheme can be done with or without control instrumentation. In the common "feed forward" mode the brine is pumped from the sump of one effect into the top water box of the next effect. The pump discharge is throttled to regulate the level in the sump. Too high a level leads to brine droplet carryover and product contamination while too low a level can cause cavitation damage to the pump. Investment in regular level controllers quickly pays out in terms of savings in operating manpower.

The second mode is called "sump-to-sump" method of brine transfer. Here brine flow takes place under the driving force of the inter-effect pressure difference while the size of the transfer line also regulates the flow rate. Brine from the sump of a given effect is recycled to the top water box of the same effect so that sump level is unaffected. Neither level control nor operator attention is necessary for maintaining a stable, smooth operation at design conditions.

d. Control of the Cold End of the Plant — The cold end of the 17-effect plant is shown in Figure 2.13. Only two instrumental control loops were involved. These were applied to level control the brine and product water discharge rates. Flow of cooling water and steam to ejectors was on manual control. For optimum operations, the brine feed rate to the whole plant had to be minimized, but not to the level where gypsum scaling would occur in the cold end of the plant. The last effect brine concentration was periodically checked by hand refractometer. If a concentration factor was found to exceed the scaling threshold (about 3.0), then the brine feed rate to the plant was reset to maintain scale free operations. An instrumental control loop could, of course, be designed using a continuous indicating/recording refractometer or a chloride ion electrode as a sensing element. One main drawback of such a system is that it takes about an hour for a change in total dissolved solids to appear at the cold end after the brine rate is adjusted.

Finally, the total automation of the plant is possible if considerable fluctuation in water demand exists. The Freeport Plant showed a rapid response in attaining 80% of the rated capacity in about 30 minutes from the initial 25%. A surge tank, with level control to initiate changes in production, could stabilize the operation of such a plant. For steady water demands, the present level of process control discussed above can be considered adequate with the exception of the pH adjustment loop on the deaerator and ratio control of seawater feed to heating steam which should be controlled instrumentally rather than manually.

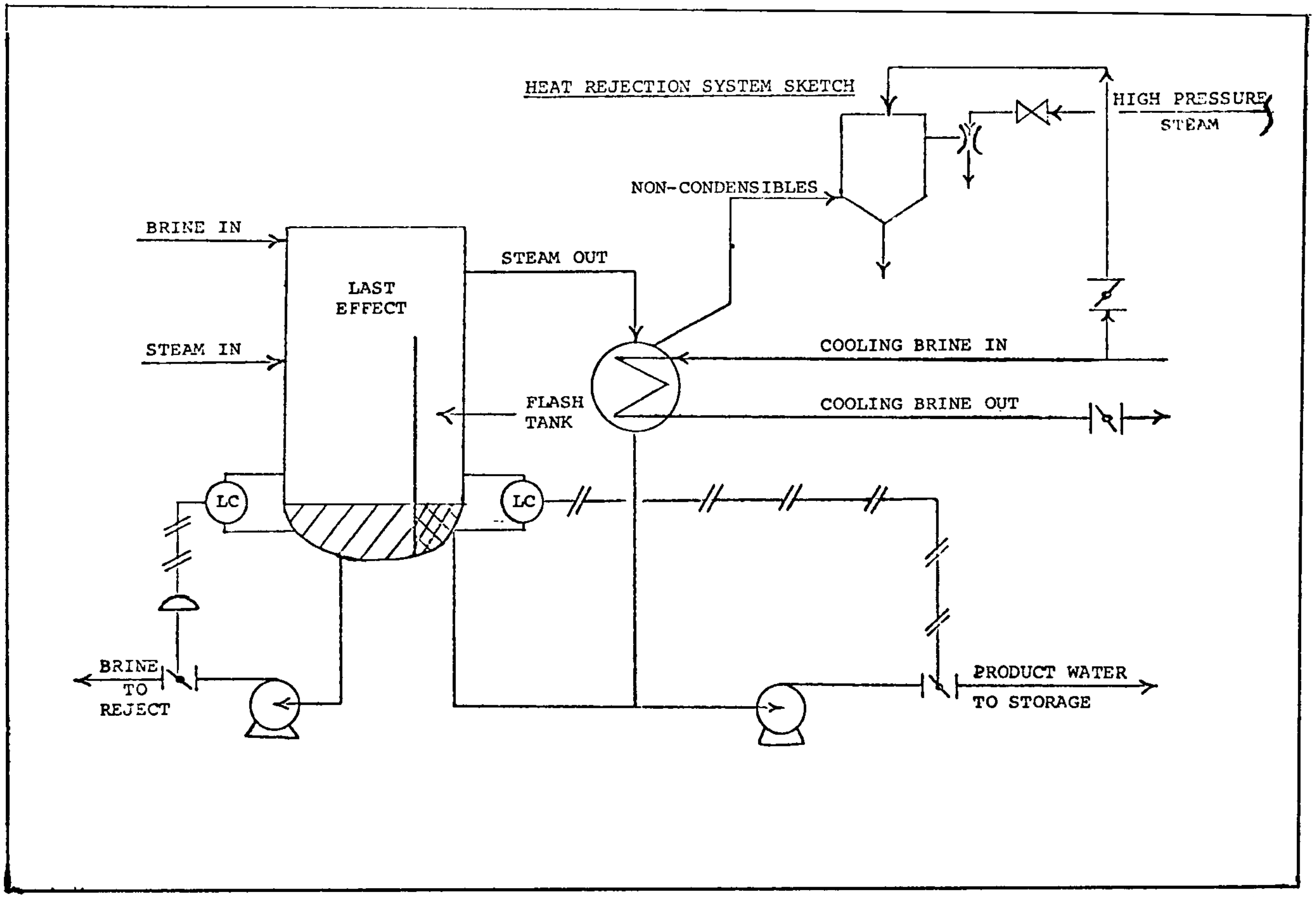


Figure 2.13. Heat Rejection System Sketch

Chapter 3

START-UP

An essential milestone in the design, construction, and utilization of a complex project, or any large endeavor that is expected to have an operating function, is the preparation for and the initial test operation. The importance of this milestone holds true even though the new facility is "essentially a copy" of an operating design. There are invariably, in addition to errors and mistakes of human origin, new subsystem designs or new innovations at some process point that need, require, and demand the final attention of debugging before being accepted for routine use. Such is the experience of technological growth, and the Demonstration Plant, Freeport, Texas, was no exception.

The prime construction contractor, Chicago Bridge and Iron Company, utilized their prior experiences when preparing the "Operating Instructions for United States Government, Department of Interior, Office of Saline Water, Demonstration Plant No. 1, Freeport, Texas". This documented the instructions and procedures for the inspection, preparation, and initial start-up of this plant.

The CB&I "Operating Instructions" were comprehensive and complete. The section on inspection before start-up included checking each item of rotating equipment for proper installation and alignment, direction of rotation, initial maintenance, and applicable safety features. Before the initial start, the instrument subsystem was carefully inspected to verify the correctness of the installation, that the air system was clean and tight, and that all meters were calibrated and functioning properly. Detailed instructions were included for the hydrostatic and vacuum testing (as required) of the evaporators and related flash tanks, pumps, and piping. Since x-ray inspection of welds was not included in these Instructions, it has been presumed this precaution was a construction item. Air pressure testing was conducted to locate, for elimination, any leaks found in the system. Air leaks into the process equipment would, and did, cause problems.

Subsystems requiring specific preparation before the initial start-up were identified and such preparation was defined. For example, the clarifier-thickener had to be loaded with a 2% slurry of $Mg(OH)_2$ in seawater prior to operation.

The status of the plant had to be established before the initial start-up. A shutdown during routine operation would naturally leave the various valves and controls in a specific setting — open or closed, as the case may be. After construction and testing for proper functioning, any valve and/or control could be left in either position. It is necessary to check each one to be certain each is in the correct position to facilitate the initial start-up, or the status of the plant must be established as a prelude for the start-up procedures.

The sequence of steps to be taken to place the plant in the operating status was prepared. Accompanying such a checklist was one for the shutdown of the plant, both under routine or normal situations and for emergencies. Also, a section was included outlining situations in which alternate methods of operation were possible.

The "Operating Instructions" contained information on specific items of equipment, certain diagrams and charts that were useful during operations, and a section of safety suggestions. Further information, and certainly in more detail, can be obtained by examining the subject document.

The construction of the first Demonstration Plant was formally completed on April 7, 1961 (First Annual Report, page I-1). Inspection of the plant and preparation for the initial start-up followed. The first problem encountered upon undertaking the specific start-up sequence occurred with the clarifier-thickener. Silt built up rapidly in the slurry tank and $CaCO_3$ formed on the feedwater control valve. These occurances caused the prompt abandonment of this method for scale control. The sulphuric acid treatment process was adopted with satisfactory results.

Early operational problems, not necessarily in sequence of occurence, included scaling in Effects X and XII (20 tubes were found plugged at the first inspection); and new impellers were needed on 6 brine pumps to increase their capacity in order to reduce the brine out concentration factor from 4 to 3. As could be expected whenever new equipment is operated for the first time, there were numerous equipment difficulties with various items. Specific items and difficulties were not identified in the annual reports; however, it is reasonable to expect these problems might well include impeller and shaft problems, air and seawater leakage, and corrosion of parts exposed to salt water. Major trouble areas did encompass such matters as brine carryover, improper equipment applications and corrosion.

During the plant inspection prior to the first official production run, heavy scaling of $CaSO_4$ was found in Effect XII with progressively reduced amounts in Effects XI and X. Much of this scale was removed by circulating cold seawater.

The proper adjustment of the electrical circuit breakers was made as operating experience was gained — a normal course of events leading to the correct setting of protection equipment. Leaks occurred at various points, necessitating inspections and corrective actions. An early determination was made to remove and replace carbon-steel tube plugs because of extensive corrosion.

April 1961 saw the first start-up of the Demonstration Plant, and by September, the contractor and operating crew were settling into their operating routine when Hurricane Carla hit the Texas Coast. As a result, the operating crew rapidly matured in the knowledge of the various plant subsystems and operating procedures. The initial recovery start-up occurred after just 5 days of emergency repair and clean-up. In fact, the plant had to be operated manually since very little of the instrumentation system was functional. Four days later, the only available and operating main feed pump shorted out, necessitating the plant shutdown. This intervening period of twelve days permitted the completion of the minimum essential repair work for sustained operations. Although failures of equipment due to Hurricane Carla did occur at later dates, these events did not pose catastrophic possibilities inasmuch as the maintenance effort had restored most of the subsystems to nearly full operational capability, and, in due time, did recondition and restore all affected parts.

Chapter 4

HEAT TRANSFER HISTORY

In a plant utilizing a multiple effect falling-film LTV evaporator for desalination of seawater, heat transfer occupies a central role. This becomes obvious when it is realized that out of 35 major items of equipment at Freeport (12-effect plant), 34 items had heat transfer as their main function.[1] The production capacity of such an evaporative plant is directly related to the rates of heat transfer in all the major items of equipment. Higher heat transfer rates lead not only to increased production rates of water but also to decreased unit costs of water. Considerable effort was, therefore, applied at the Freeport Plant over its eight-year period of operation to obtain information relating to long-term heat transfer rates and the effects of design and process variables on these rates. Before the presentation of these investigations and the results obtained, it is essential to undertand the theory underlying heat transfer in such ME-LTV falling-film plants. This chapter, therefore, has been divided into the following subsections:

A. Theory and Original Design
B. History of Overall Coefficients by Effect
C. Effect of Distribution Devices on U
D. Effect of Process and Design Variables on U
E. Types of Surface Used and Performance
F. Variation of U with Tube Type
G. Effect of Non-condensibles Accumulation and Venting Rates on U
H. Preheater Performance
I. Heat Rejection Condenser Performance

---

(1) Dykstra, D. I., Chem. Eng. Prog., 61, No. 7, 80 (1965)

A. THEORY AND ORIGINAL DESIGN

1. Mathematical Model of the Mechanism of Heat Transfer

As stated earlier, heat transfer is the main function of most of the major items of equipment in an evaporative desalination plant of the Freeport type. Of the various such units, however, the evaporator effects play the most important role. The mechanism of heat transfer is unique in these effects because of the phase change and physical property variation down the entire length of the tubes. A simple sketch of the heat transfer zones involved is shown in Figure 4.1.

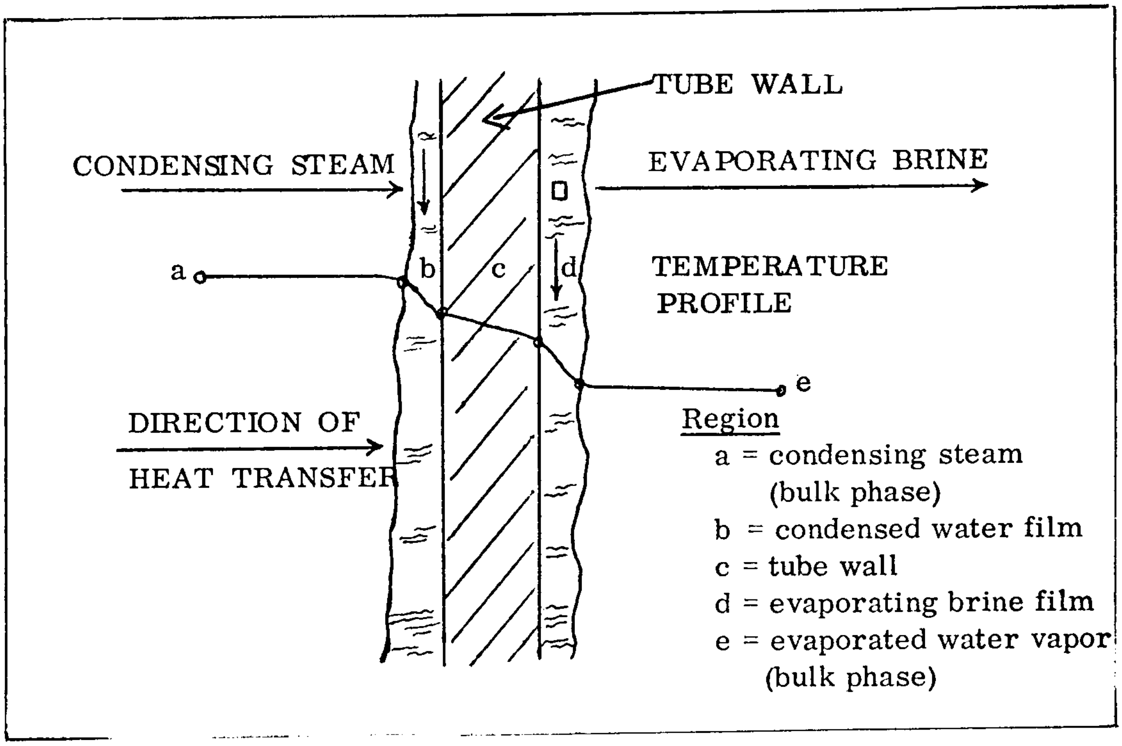


Figure 4.1. Schematic Model of Heat Transfer

Neglecting the non-condensibles and scaling effects, there are five main regions involved in the heat transport route as marked in Figure 4.1. On the outside of the tubes is region a consisting of the bulk phase of steam or vapor from the previous high temperature effect. The water film b is the result of continuous condensation of this steam on the outside of the tubes. The tube wall forms the third region c in the heat flux route. Seawater or brine from the previous effect (in feed forward mode) enters the top of the tubes through a liquid distribution device (triangular notch weir or porcelain nozzle) and flows down by gravity as a film on the inside surface of the tube. Heat transported to this region d causes a proportional quantity of water to evaporate, leading to the formation of bulk vapor phase e inside the tube. Under steady state conditions, temporal variations of temperature, pressure, concentration, etc., in these zones are minimal, but the spatial variations are significant. For example, the thickness of the water film in zone b is continuously increasing downward as more steam keeps condensing along the whole length of the tube. The brine film d, however, decreases in thickness as it proceeds to the exit end of the tube because of continual depletion of its water content due to evaporation. Spatial temperature variation is brought about by the two-phase pressure drop down the tube so that the temperature driving force is less at the tube top than at the bottom exit. High vapor generation rates, low absolute pressures, and small I.D. long tubes lead to large pressure drops and hence reduced temperature driving forces. But increased vapor rates are not totally undesirable since increased interfacial shear reduces the film thickness and thus tends to increase the heat transfer coefficient. Therefore, there is spatial variation of temperature, pressure, fluid properties, and dimensionless numbers like Reynolds Number, Prandtl Number, etc. Because of the existence of temperature profile across the heat transfer zones as shown in Figure 4.1, there is additional variation of conditions across the zones at any given distance from the tube top. Prediction of accurate heat transfer rates is extremely difficult due to this complex situation. Prengle, Dukler, and Crump, Inc., under an OSW contract, developed the following mathematical model and predictive technique (OSW R&D Progress Report Number 74). The model is based on the following assumptions:

- Thermal resistance of the bulk steamside zone a is negligible.
- Continuous filmwise condensation exists on the steamside.
- Interfacial shear on the steamside is negligible.
- On the brine side, water evaporation takes place from the brine film interface. A nucleate boiling mechanism does not exist at the low flux rates and ΔT's obtained in such plants.
- Bulk phase resistance of the vapor core inside the tube is negligible.

The principal index of heat transfer performance of an equipment is U, the overall coefficient of heat transfer which is connected to the individual coefficients of the three remaining zones b, c, and d as follows:

$$U = \frac{1}{\frac{1}{h_b} + \frac{1}{h_c} + \frac{1}{h_d}} \qquad (4.1)$$

where

h_b = film coefficient for the steamside water film

h_c = individual coefficient for the tube wall, equal to thermal conductivity of tube material divided by the tube wall thickness

h_d = coefficient for the brine film

The prediction of U, therefore, can be made once h_b and h_d can be predicted as a function of the design and process variables like Reynolds Number, Prandtl Number, interfacial shear β, etc.

The heat flux (heat transferred per unit time per unit area) at any position, y - distant from the wall can be written as:

$$q'_y = -(k_L + E_H C_{pL} \rho_L) \frac{dT}{dy} \qquad (4.2)$$

where k_L, E_H, C_{pL}, ρ_L represent the molecular thermal conductivity, eddy thermal conductivity, specific heat, and density of the liquid at the position y. The local temperature gradient is $\frac{dT}{dy}$. This equation can be rearranged and integrated to give the temperature drop in either film b or d (see Figure 4.1):

$$\int_{Tw}^{Tm} dt = -\int_{o}^{m} \frac{q'_y dy}{(k_L + E_H C_{pL} \rho_L)} \qquad (4.3)$$

$$Tm - Tw = -q' \int_{o}^{m} \frac{dy}{(k_L + E_H C_{pL} \rho_L)} \qquad (4.4)$$

where m is the thickness of the liquid film at a given distance from the tube top, and Tm is the temperature of the liquid film interface with vapor, and Tw that of the tube wall. Under steady-state conditions with no sensible heating across the thin film, q'_y is independent of y, so the subscript y has been dropped and q taken out of the integral sign in Equation 4.4.

Now the individual heat transfer coefficient h for this film can be written as:

$$h = - \frac{q'}{Tm - Tw} \tag{4.5}$$

so that on substitution (for q') in Equation 4.4, the heat transfer coefficient can be obtained as:

$$h = \frac{1}{\int_{o}^{m} \frac{dy}{(k_L + E_H C_{pL} \rho_L)}} \tag{4.6}$$

In order to solve this equation for h, it is necessary to have equations that will give variation of E_H with position y, the variation of m, the film thickness, with liquid rates; fluid properties and the interfacial shear β . Two equations are proposed by Dukler (OSW R&D Report Number 74) to predict E_H:

i. for the region close to tube wall:

$$E_H = n^2uy \left[1-\exp(\frac{-n^2uy\rho}{\mu})\right] \tag{4.7}$$

ii. and for the fully developed turbulent region away from the tube wall:

$$E_H = P\,(du/dy)^3 \,/\, (d^2u/dy^2)^2 \tag{4.8}$$

where n and P are constants, u is the local velocity at a distance y from the tube wall, and u is the liquid viscosity. The two regions are divided at film thickness b given by:

$$b = \frac{26u}{\sqrt{\tau w \, \rho g_c}} \tag{4.9}$$

where τw is the shear stress due to fluid friction at the tube wall. The velocity distribution u = u(y) is obtained by Dukler[1] from numerical integration of film momentum balance non-linear equations. The derivatives du/dy and d^2u/dy^2 are also obtained in numerical form. Values of E_H can be calculated once this velocity distribution data is available.

The liquid film thickness m is obtained from the material balance equation:

$$W_L = \rho_L \int_o^m (D-2y)udy \tag{4.10}$$

where W_L is the local liquid flow rate and D is the internal diameter of the tube. For a given flow rate and velocity distribution, only one value of m will satisfy Equation 4.10. This is obtained by trial and error. The liquid Reynolds Number is easily obtained as:

$$Re_L = \frac{4W_L}{\pi D \mu} \tag{4.11}$$

Because of the fact that the velocity distribution in the film is affected by interfacial shear arising from concurrent vapor flow over the film, its thickness m depends both on the Reynolds Number, Re_L, and the dimensionless interfacial shear β given by the following equation:

$$B = \frac{(dP/dL)_{TP}Dg\ 1/3}{4\rho_L\ 1/3 - \mu\ 2/3} \tag{4.12}$$

where $(dP/dL)_{TP}$ represents the axial pressure gradient due to frictional energy loss in two phase flow. Increasing Re_L increases the film thickness while increasing β has the reverse effect.

Since both m and E_H can now be obtained as functions of position y, the integral in Equation 4.6 can be evaluated to yield the local value of the individual or film transfer coefficient h. In OSW R&D Progress Report Number 74, Figures 6 through 9 and 11 through 14, values of this coefficient have been plotted as functions of Reynolds Number with interfacial shear β as a parameter for selected values of Prandtl Number, Pr_L. It is seen from these figures that increasing β has a much more pronounced effect on the heat transfer coefficient than that of Reynolds Number.

---

(1) Dukler, A. E., <u>Chem. Eng. Prog.</u>, Symposium Series 56, No. 30, 1, (1960)

The discussion given above was limited to local coefficients; i.e., at a certain point down the length of the tube. Since the observed heat flux q is based on the entire tube area, it is essential to compute h_{avg}, the average coefficient. (This is necessitated by the fact that due to continuous evaporation, or condensation, the film thickness is variable, as are the Reynolds, Prandtl, and β numbers.) Making use of enthalpy balance on a differential length of tube and then integrating from L = 0 to L = L (tube length), one obtains:

$$\int_{Re_{in}}^{Re_{out}} \frac{dRe_{Lx}}{dh_x} = \int_0^L \frac{4\Delta T_x}{\mu\ L\ \lambda}\, dL \qquad (4.13)$$

where λ is the enthalpy of vaporization, and ΔT_x is the temperature driving force in this differential tube length dL. For the entire tube, one can write:

$$\frac{\Delta\ Re_L}{h_{avg}} = \frac{4\Delta T_{avg} L}{\lambda\ \mu_L} \qquad (4.14)$$

where $\Delta\ Re_L$ is the change in Reynolds Number over length L. Finally, combining Equations 4.13 and 4.14, the following equation is obtained for the average coefficient:

$$h_{avg}\psi_{avg} = \frac{Re_L \displaystyle\int_0^L \frac{\Delta Tx\ dL}{\Delta T_{avg} L}}{\displaystyle\int_{Re_{in}}^{Re_{out}} \frac{Re_{Lx}}{h_x}} \qquad (4.15)$$

where

$$\psi = \left(\frac{\mu_L^2}{\rho_L^2 g K_L^3}\right)^{1/3} \qquad (4.16)$$

For obtaining h_x variation with Re_{Lx} (local Reynolds Number), the change in β must be known which means essentially that the two-phase pressure drop must be determined. At low pressures (i.e., below 1 atm.), there is no adequate method for predicting this two-phase pressure drop. Dukler[1] utilized empirical correction factors to obtain tentative correlations. (A number of correlations are also listed in "Saline Water Conversion Engineering Data Book, Supplement Number 1, Oct. 1966, published by U.S. Department of Interior.) The average overall coefficient, U_T, is obtained by substituting steamside and brineside h_{avg} values in Equation 4.1.

2. Comparison of Theory with Experimental Data

For logical comparison, the basis must be common to the compared data. Where both U and ΔT_o vary over the length of the tube, the following equation gives the correct heat transfer rate:

$$q = \int_0^L U \Delta T_o a_d dL \tag{4.17}$$

where a_d is the heat transfer area per unit tube length in zone d (Figure 4.1), and ΔT_o is the overall temperature driving force at any distance down the tube top. For average U_d over the entire tube length:

$$q = U_d \int_0^L \Delta T_o a_d dL \tag{4.18}$$

This U_d can be compared directly with the theoretical average U_T to check the validity of the theoretical correlation. Prengle, Dukler, and Crump, Inc., reported[1] agreement within 5% of the Wrightsville Beach Pilot Plant data.

3. Effect of Tubeside Pressure Drop

Prengle et al reported[1] the effect of tubeside pressure drop on the temperature driving force ΔT. For high brine temperatures (above approximately 200°F), a pressure drop of even 0.8 inch Hg in the tube did not show significant influence on ΔT, whereas at low temperatures (and, of course, low pressures) of approximately 125°F, the ΔT at the tube top was found to be reduced to 20 percent of the ΔT at tube bottom for a pressure drop of only 0.2 in Hg!

---

(1) Prengle, Dukler, Crump, Inc., OSW R&D Report Number 74.

As stated earlier, the large vapor flow rate (due to high specific volume of vapor at low pressure) reduces ΔT but increases U so that to obtain optimum heat flux (q/A), the product $U \Delta T$ must be optimized. Under ideal conditions, each effect tube diameter, length, spacing, etc., must be optimized to yield the best value of heat flux. From a practical standpoint, at least two types of tubes should be used — for high temperature effects small diameter but long tubes; whereas, for low temperature/low pressure effects, short tubes with a large I. D. would reduce the excessive pressure drop.

4. Important Conclusions

The following important conclusions were drawn by Prengle et al[1] on the basis of the theory as given above:

- The ability to generalize the predictive technique depends on generalizing the method for predicting pressure drops in two phase flow.

- The overall coefficients do not drop off drastically at lower operating temperatures as was originally concluded (OSW Progress Reports Numbers 26 and 456). These were apparent coefficients based on the ΔT at the tube exit, which is greater than the effective ΔT over the entire tube length. Decreased ΔT results from the pressure drop necessary to move the vapors out of the tubes.

  An important modification to the present design concept is suggested by the theory — namely, that the tube diameters and lengths should not be the same in all effects but should be varied so as to maximize the $U \Delta T$ product. Once a reliable pressure drop correlation is available, it is possible to mathematically establish this maximum for a given operating pressure level, fluid properties, tube size, etc.

- Calculation of steamside and brineside individual coefficients indicate that the steamside resistance is of the same order or even larger than the brineside resistance. This suggests that, for improvement in U, a grating type (e.g., fluted type)

---

(1) Prengle, Dukler, Crump, Inc., OSW R&D Report Number 74.

surface may be used, tube spacings may be decreased, and baffles (or dummy tubes) may be used on shellside to increase vapor shear β .

- Entrainment can be serious in small diameter tubes handling a large vapor load. Entrainment affects heat transfer rates, pressure drops, and scaling.

- The heat transfer coefficient is dependent on the Reynolds Number, Prandtl Number, dimensionless interfacial shear β , and the heat transfer group:

$$\psi = \left[\frac{\mu_L^2}{\rho_L^2 g k^3} \right]^{1/3}$$

  Variations in liquid flow rates (i.e., Re_L) do not affect the coefficient as much as those in interfacial shear β .

5. Original Design

W. L. Badger Associates, Inc. (Ann Arbor, Michigan) provided the initial design specifications for the Freeport Plant which until 1967 consisted of 12 evaporator effects. Tube size of 2 inch O.D. and 24 foot length was selected because the Wrightsville Beach pilot plant data (OSW Report Number 26) was available for only this tube size and it indicated satisfactory values of the overall coefficient U. Partial heat transfer data from these pilot plant tests is given in Table 4.1. Effect number column indicates that the particular run most closely approximated the conditions that were expected in this effect of the demonstration plant ("Specifications Number 195", page IV-22).

B. HISTORY OF OVERALL COEFFICIENTS BY EFFECT

Before presenting the effectwise history of the overall heat transfer coefficients the calculations performed to check the consistency and reliability of the overall and effectwise heat and material balance data are given below. Plant performance data, including the overall coefficients, are derived from this heat and material balance data. For random check the Development Runs 9-2 and 16-3 were selected to represent 12- and 17-effect plant operation respectively.

Table 4.1

Heat Transfer Test Data for Conditions of Demonstration Plant Operation

Run No.	Effect No.	Temp. °F. Feed	Temp. °F. Sat-Vap	ΔT, °F	U	ΔP, in. Hg.
LWCH-2	I	242.2	249.5	8.8	721	2.76
LWCC-2	II	249.5	241.3	6.2	741	6.47
LWCF-2	IV	230.5	222.4	6.8	727	5.80
LWCE-2	VI	213.7	201.6	7.9	666	5.14
LWCD-2	VIII	189.7	176.7	10.1	532	---
LWCC-2	X	162.7	151.0	11.0	467	1.95
LWCB-1	XI	153.2	141.0	11.7	407	---
LWCA-3	XII	140.8	119.7	12.8	377	1.87

Notes:

1). Units of U are Btu/hrft^2 °F. It is based on inside tube area and ΔT=condensing vapor temp. - tube exit brine temp.

2). ΔP is pressure difference between distributor plate (above orifice) and vapor head.

1. Overall Heat and Material Balance: Development Run Number 9-2

Figures for this run were taken from page 102a, Fourth Annual Report by Stearns-Roger (OSW R&D Progress Report Number 171). Operating time of the run was 2931 hours.

a. Overall Material Balance —

Streams In, lb/hr			Streams Out, lb/hr		
i.	Inlet seawater	2,152,511	i.	Waste seawater	1,066,664
ii.	Dow steam	34,000	ii.	Concentrated brine	105,297
iii.	Sulfuric acid Caustic soda	less than 75 lb/hr	iii.	Product water:	
				to Freeport	93,680
				to Dow Company	285,210
				to Plant	20,960
			iv.	Vents	770
	Total	2,186,586		Total	1,572,581

Obviously there is an error in reporting the inlet seawater rate or waste seawater (page 102a, Fourth Annual Report). An alternative balance can be written between the inlet seawater to the deaerator plus the Dow steam, and the brine blowdown, product water, and vents:

$$492,000 + 34,000 = 399,850 + 123,450 + 770 + 1930$$
$$526,000 = 526,000$$

Therefore, this material balance is satisfactory. There is definitely some error in the reported values of seawater intake (total) or waste seawater from heat rejection condenser HX-312.

b. Overall Heat Balance — Since the total seawater rate and the waste seawater rate are not correctly stated only an approximate check can be attempted on the overall enthalpy balance.

Heat In, Btu/hr	Heat Out, Btu/hr
i. Seawater feed 492,000 x 47.51 =23,374,920	i. Brine blowdown 123,450 x 78.9 = 9,740,205
ii. Dow steam 34,000 x 1279.55=43,504,700	ii. Product water to Dow (285,210 x 71.2) =20,306,952 to Freeport (93,680 x 62.8) = 5,883,104 to Plant (20,960 x 62.8) = 1,316,288
	iii. Vents from Effects II & III 874,650 from Deaerator 1,171,065 from HX-312 1,002,330
	iv. Losses from Evaporator effects 1,224,700
	v. Heat rejected by HX-312 24,545,600
Total 66,879,620	Total 66,064,894
Difference (In - Out) = 814,726	

This is only 1.21% of the total heat input, so the overall heat balance can be considered satisfactory.

c. Effectwise Heat and Material Balance (Run Number 9) —

Effect I

Material Balance

Streams In, lb/hr			Streams Out, lb/hr		
i.	Seawater feed	493,000	i.	Brine blowdown	458,955
ii.	Dow steam (with desuperheating water)	38,700	ii.	Vapor to Effect II	34,045
			iii.	Condensate	37,720
			iv.	Vent	980
	Total	531,700		Total	531,700

Therefore, the material balance is satisfactory.

Heat Balance

Heat In, Btu/hr			Heat Out, Btu/hr		
i.	SWF 493,000 x 224.1 =	110,481,300	i.	Brine 458,955 x 226.1 =	103,769,725
ii.	Dow steam 38,700 x 1132.6 =	43,831,620	ii.	Vapor out 34,045 x 1169.4 =	39,812,223
			iii.	Condensate 37,720 x 244.9 =	9,237,628
			iv.	Vent 980 x 1172.5 =	1,149,050
			v.	Losses =	321,300
	Total	154,312,920		Total	154,289,926

The difference being only 22,994 Btu/hr, the heat balance can be considered satisfactory.

Effect X

Material Balance

Streams In, lb/hr		Streams Out, lb/hr	
i. Brine from Effect IX	204,060	i. Brine to Effect XI	175,150
ii. Vapor from Effect IX	26,255	ii. Vapor to Effect XI	28,910
		iii. Condensate	26,175
		iv. Vent	80
Total	230,315	Total	230,315

Heat Balance

Heat In, Btu/hr		Heat Out, Btu/hr	
i. Brine in 204,060 x 128.2 =	26,160,492	i. Brine out 175,150 x 111.7 =	19,564,255
ii. Vapor in 26,255 x 1134.8 =	29,794,174	ii. Vapor out 28,910 x 1127.3 =	32,590,243
		iii. Condensate 26,175 x 139.5 =	3,651,412
		iv. Vent 80 x 1134.8 =	90,784
		v. Losses =	56,100
Total	55,954,666	Total	55,952,794

The difference between the heat in and out is only 1,872 Btu/hr so the heat balance can be considered satisfactory.

Since the overall heat and material balance and effectwise balance for Effects I and X were found to be satisfactory, it is assumed that the remaining effects would yield similar results.

d. Overall Heat Transfer Coefficients (U) — The apparent value of the coefficient U is obtained from the equation:

$$U = \frac{Q_E}{A \Delta T}$$

where Q_E is the heat transfer rate (Btu/hr) across area A (ft^2) and ΔT is the difference in temperature of the condensing vapor and the brine temperature at tube exit. Correct average U can be calculated if the temperature history of brine inside the tube is known. If this brine temperature variation is assumed as uniformly decreasing from top to bottom of the tube, then logarithmic mean of the tube entrance and exit ΔT's can be substituted in the above equation to obtain average U corrected for the ΔT variation due to pressure drop. The data for the following calculation is taken from Table 5-10 (page 113) of the Fourth Annual Report.

Effect I

Heat transferred, Q_E = enthalpy of steam in - enthalpy of condensate - enthalpy of vent stream - enthalpy loss from evaporator surface

= 43,831,620 - 9,237,628 - 1,149,050 - 321,300

= 33,123,642 Btu/hr

(Reported value is 33,446,100 Btu/hr on page 102a, Figure 5-12, Fourth Annual Report)

$$U_a = \frac{33,123,642}{(5031)(275.8 - 267.0)}$$

$$= \underline{\underline{748.2 \text{ Btu/hr ft}^2 \text{ }^oF}}$$

The reported value of U_a is 706.6 which is lower than the value calculated above because of incorrect Q_E and ΔT (9.7°F instead of 8.8) values.

Effect VI

$$U_a = \frac{26,425,100}{(4000)\ (223.6 - 214.0)}$$

$$= \underline{\underline{688.2 \text{ Btu/hr ft}^2\ {}^{o}\text{F}}}$$

Here, Q_E value was taken as reported in Table 5.12 on page 117 of the Fourth Annual Report. The value of U reported is 736.5 which is definitely in error. Using the Stearns-Roger reported values for Q_E, A, and Δ T, this computation yields a U of 564.6 Btu/hr ft^2 ^{o}F.

Effect X

$$U_a = \frac{26,053,100}{(4025)\ (171.6 - 156.0)}$$

$$= \underline{\underline{414.9 \text{ Btu/hr ft}^2\ {}^{o}\text{F}}}$$

Reported value is 458.9 which is again incorrect when using all the data as reported — the computed value being 348.0 Btu/hr ft^2 ^{o}F.

Effect XI

$$U_a = \frac{25,037,500}{(4780)\ (153.0 - 139.1)}$$

$$= \underline{\underline{376.8 \text{ Btu/hr ft}^2\ {}^{o}\text{F}}}$$

The reported value is 418.5 which should be 333.6 using all data as reported by Stearns-Roger.

Effect XII

$$U_a = \frac{22,985,800}{(6276)\ (137.3 - 121.3)}$$

$$= \underline{\underline{228.9 \text{ Btu/hr ft}^2\ {}^{o}\text{F}}}$$

The reported value is 254.6 which should be 211.7 when $\Delta T = 17.3^{\circ}F$ and the Q_E and A values are the same as above.

Values of U_a for other effects are as follows:

Effect II $27,275,700/(4000)(266.1 - 257.4) = 783.8$ Btu/hr ft^2 $^{\circ}$F
Effect III $26,990,500/(4000)(256.5 - 247.0) = 710.3$ Btu/hr ft^2 $^{\circ}$F
Effect IV $26,534,700/(4000)(245.5 - 236.2) = 713.3$ Btu/hr ft^2 $^{\circ}$F
Effect V $26,512,800/(4000)(235.0 - 224.8) = 649.8$ Btu/hr ft^2 $^{\circ}$F
Effect VII $26,832,100/(4000)(211.9 - 201.9) = 670.8$ Btu/hr ft^2 $^{\circ}$F
Effect VIII $29,009,600/(4000)(200.0 - 188.3) = 619.9$ Btu/hr ft^2 $^{\circ}$F
Effect IX $27,183,700/(4000)(186.9 - 172.6) = 475.2$ Btu/hr ft^2 $^{\circ}$F

These U_a values are plotted in Figure 4.2 against the brine (sump) temperature. The U_a values not only show the correct trend (i.e., increase) with increasing brine temperature but also a reasonable agreement with the Wrightsville Beach pilot plant data reported in OSW R&D Progress Report Numbers 26 and 456.

2. Overall Heat and Material Balance: Development Run Number 16-3

Figures for this material balance were taken from the Run Number 16-3 H & M Balance Sheet, page 93, of the FY-1969 Eighth Annual Report by Stearns-Roger Corporation. This run was started on 8 March, 1969, and completed on 2 May, 1969, with the total operating time of 960 hours. Figure 4.3 depicts a simplified input-output flow for a 17-effect plant.

a. Overall Material Balance —

Streams In, lb/hr			Streams Out, lb/hr		
i.	Dow steam	25,800	i.	Brine blowdown	144,700
ii.	Seawater feed	473,000	ii.	Product water	349,800
iii.	Acid and alkali	less than 100 lb/hr	iii.	Vent gases	
				from Effect IV	200
				from Deaerator	1,600
				from Effect XII	300
				From HX-318	100
	Total	498,800		Total	496,700

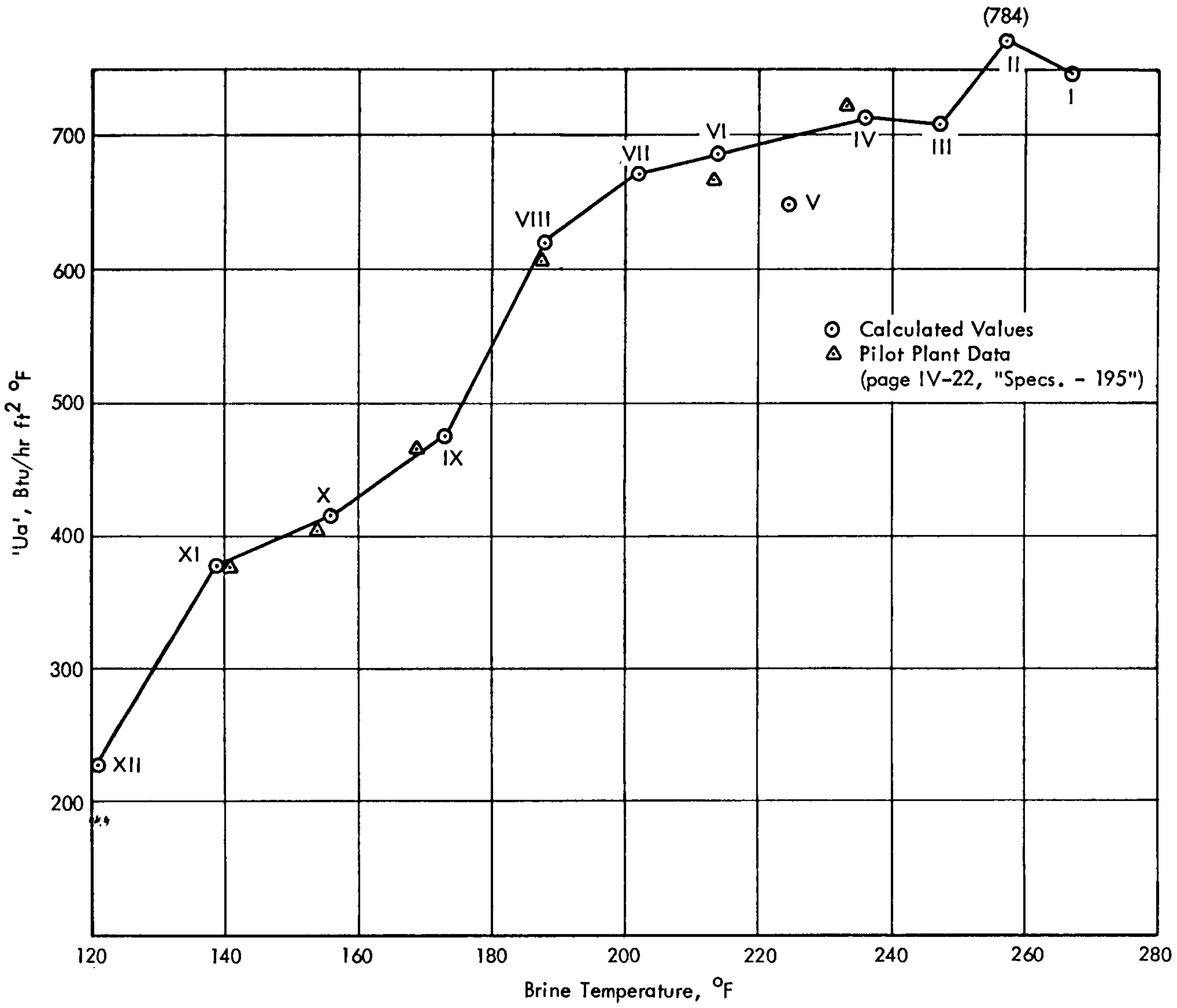


Figure 4.2. Overall Coefficients for Development Run Number 9

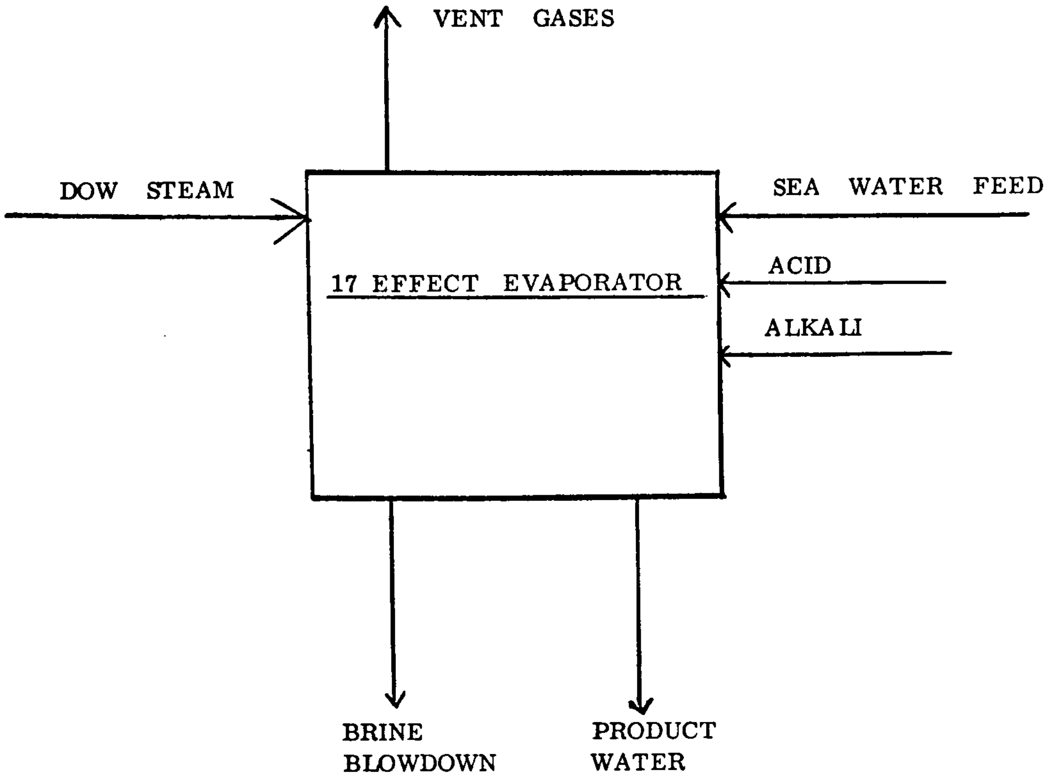


Figure 4.3. Net Flows Entering and Leaving the Total Plant

Continuing from page 87, the difference = in - out = 2100 lb/hr

$$= \frac{2100 \times 100}{498,800}$$

$$= 0.42\%$$

Hence, the overall material balance can be considered good.

b. Overall Enthalpy Balance —

Enthalpy In, Btu/hr		Enthalpy Out, Btu/hr	
i. Dow steam 25,800 x 1256 =	32,404,800	i. Brine blowdown 144,700 x 61.4 =	8,884,580
ii. Seawater 473,000 x 39.7 =	18,778,100	ii. Product water 349,800 x 58.2 =	20,358,360
		iii. Vent streams	
		Effect IV	232,660
		Deaerator	1,773,120
		Effect XII	340,470
		HX-318	110,370
		iv. Enthalpy loss to atmosphere	4,400,000
		v. Heat rejected in HX-318	15,290,000
Total	51,182,900	Total	51,389,560

$$\text{Percent difference} = \frac{206,660 \times 100}{51,182,900}$$

$$= 0.403$$

Therefore, the enthalpy balance can be considered satisfactory.

c. Effectwise Heat and Material Balance —

- H & M Balance for Effect I. Figure 4.4 depicts simplified input-output flows for Effect I.

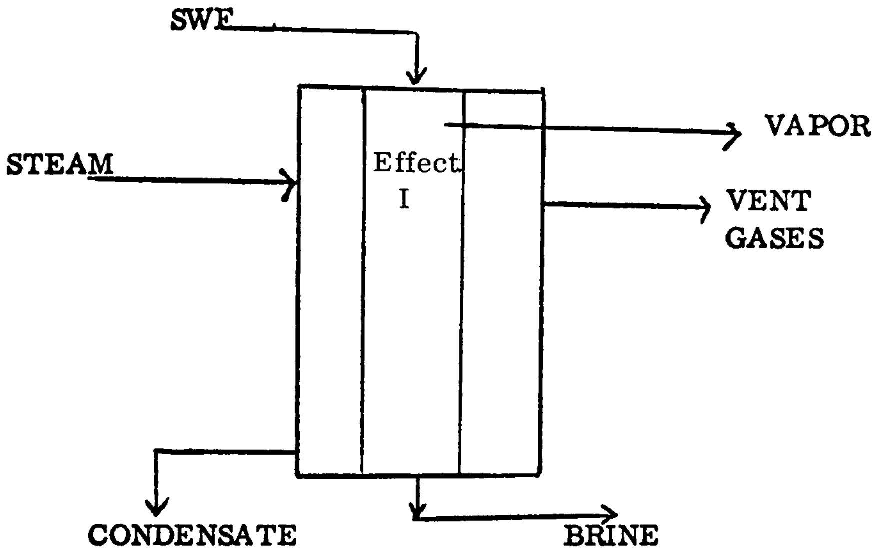


Figure 4.4. Flows Entering and Leaving Effect I

Material Balance. Since the steamside and brineside streams do not come into physical contact their material balances can be written separately.

For steam side: (on lb/hr basis)

steam in = condensate out + vent gases
28,900 = 28,400 + 500
= 28,900

For brine side:

seawater feed in = brine out + vapor out
472,000 = 446,500 + 25,700
= 472,200

Therefore, material balance is satisfactory on both steam and brine sides of Effect I.

Enthalpy Balance

Streams In, Btu/hr	Streams Out, Btu/hr
i. Seawater 472,000 x 226.9 = 107,096,800	i. Brine 446,500 x 228 = 101,802,000
ii. Steam 28,900 x 1127.3 = 32,578,970	ii. Condensate 28,400 x 243.5 = 6,915,400
	iii. Vapor 25,700 x 1169 = 30,043,300
	iv. Vent 500 x 1172.5 = 586,250
	v. Heat loss = 400,000
Total 139,675,770	Total 139,746,950

$$\text{Percent difference} = \frac{\text{enthalpy in - enthalpy out}}{\text{enthalpy in}} \times 100$$

$$= \frac{71,180 \times 100}{140,976,270}$$

$$= 0.05$$

Hence, the enthalpy balance can be considered very good. To obtain the overall heat transfer coefficient U, the rate of heat transferred, Q_E, must be known. This can be obtained either from salt balance or enthalpy balance.

Enthalpy balance for steam side gives:

$$Q_E = H \text{ steam} - H \text{ condensate} - H \text{ vent} - H \text{ loss} \qquad (4.19)$$

where H stands for total enthalpy gain or loss per hour. Substituting values from above gives,

$$Q_E = 32,578,970 - 6,915,400 - 586,250 - 400,000$$
$$= 24,677,320 \text{ Btu/hr}$$

Part of this heat goes to preheat the seawater feed to its saturation temperature: 472,000 x 0.98 x (266 - 264.7) = 601,328 Btu/hr. Therefore, the enthalpy available for evaporation is 24,075,992 Btu/hr. Overall coefficient U is then obtained as:

$$U = \frac{Q_E}{A \triangle T} \tag{4.20}$$

$$= \frac{24,677,320}{(4070)\ (275.5 - 266.0)}$$

$$= 638.2 \approx \underline{\underline{638}}\ \text{Btu/hr ft}^2\ {}^{\circ}\text{F}$$

The reported value of U (referenced H&M Balance Sheet) is 650 Btu/hr ft^2 $^{\circ}$F. The difference between the reported value and the above calculation is 12 Btu/hr ft^2 $^{\circ}$F or 1.88% based upon the calculated value.

Now on a salt-balance basis Q_E is obtained from rate of evaporation less evaporation resulting from flashing of incoming brine. The evaporation from n'th effect is:

$$E_n = L_{n-1} - L_n \tag{4.21}$$

and

$$L_n = SWF \times \frac{Co}{C_n} \tag{4.22}$$

where L_n is blowdown from n'th effect having chlorinity (or salinity) C_n, and SWF is the seawater feed rate to the first effect with chlorinity equal to Co. Substituting the reported values for Run 16-3, the rate of evaporation from the first effect is obtained:

$$E_1 = L_o - L_1$$

$$= SWF - \frac{SWF \times Co}{C_n}$$

$$= 472,000 \left[1 - \frac{2.26}{2.39}\right]$$

$$= 25,677 \text{ lb/hr}$$

The total enthalpy of vaporization in this stream is 25,677 x 935.8 = 24,028,536 Btu/hr. First effect feed is subcooled so additional heat must be transferred to bring it up to the saturation temperature. This is equal to 601,328 Btu/hr as calculated above. Therefore Q_E = 24,629,864 Btu/hr.

Finally, U is obtained on a salt balance basis as:

$$U = \frac{24,629,864}{4070 \times 9.5}$$

$$= 637 \text{ Btu/hr ft}^2 \text{ }^{o}\text{F}$$

This is 2 percent less than the reported value of 650 Btu/hr ft^2 °F, but agrees closely with the enthalpy balance computed by Equation 4.2. This value is within the reported accuracy of data. It is to be noted that this U is based on the assumption of isothermal (i.e., constant ΔT) conditions down the whole length of the tube.

- H & M Balance for Effect II. Typical stream flows for Effect II are given in Figure 4.5.

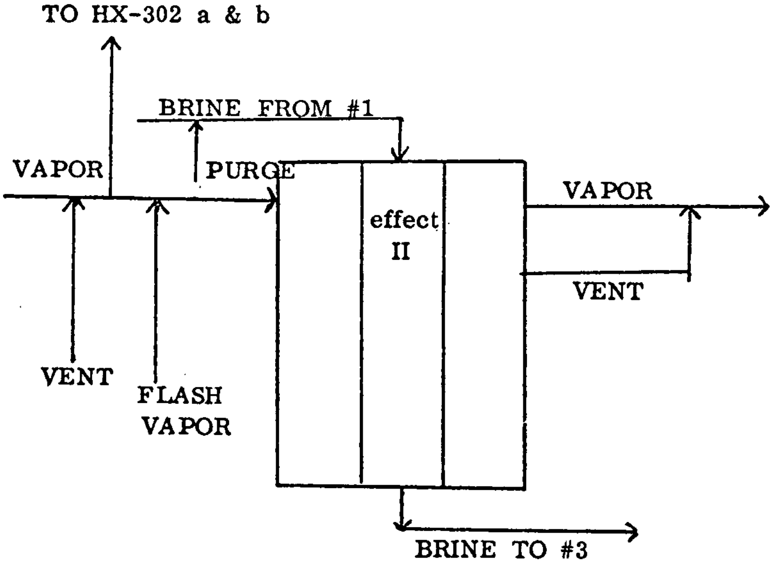


Figure 4.5. Typical Net Flows In and Out of an Effect

Material Balance

Streams In, lb/hr			Streams Out, lb/hr		
i.	Brine from Effect I	446,500	i.	Brine blowdown	422,400
ii.	Purge water	900	ii.	Vapor	24,800
iii.	Vapor from Effect I	25,700	iii.	Condensate	26,100
iv.	Vent	500	iv.	Vent	400
v.	Flash vapor	300			
	Total	473,900		Total	473,700

The difference is 200 lb/hr; hence the material balance is good.

Enthalpy Balance

Enthalpy In, Btu/hr			Enthalpy Out, Btu/hr		
i.	Brine 446,500 x 228	101,802,000	i.	Brine 422,400 x 218.9	92,463,360
ii.	Purge water 900 x 50	45,000	ii.	Vapor 24,800 x 1166	28,916,800
iii.	Vapor 25,700 x 1169	30,043,300	iii.	Condensate 26,100 x 233.2	6,086,520
iv.	Vent 500 x 1172.5	586,250	iv.	Vent 400 x 1169	467,600
v.	Flash vapor 300 x 1169	350,700	v.	Heat losses	400,000
			vi.	Heat transferred in preheater HX-302a & b	4,500,000
	Total	132,827,250		Total	132,834,280

The difference is only 7030 Btu/hr so the enthalpy balance is excellent.

Heat transferred to evaporating brine, Q_E, is obtained as:

$$
\begin{aligned}
Q_E &= \text{H vapor} - \text{H condensate} - \text{H vent} - \text{H loss} - \text{H preheat} \\
&= 30{,}980{,}250 - 6{,}086{,}520 - 467{,}600 - 400{,}000 - 4{,}500{,}000 \\
&= 19{,}526{,}130 \text{ Btu/hr}
\end{aligned}
$$

$$
\begin{aligned}
U &= \frac{19{,}526{,}130}{3660 \times (265.2 - 256.9)} \\
&= 642 \text{ Btu/hr ft}^2 \ {}^{\circ}\text{F}
\end{aligned}
$$

This differs from the reported value of 630 Btu/hr ft^2 $^{\circ}$F by approximately 2%.

- H & M Balance for Effect III.

Material Balance

Streams In, lb/hr			Streams Out, lb/hr		
i.	Brine from Effect II	422,400	i.	Brine blowdown	401,200
ii.	Purge water	900	ii.	Vapor	22,100
iii.	Vapor	24,800	iii.	Condensate	25,300
iv.	Vent	400	iv.	Vent	400
v.	Flash vapor	500			
	Total	449,000		Total	449,000

Hence, material balance is excellent.

Enthalpy Balance

Enthalpy In, Btu/hr			Enthalpy Out, Btu/hr		
i.	Brine 422,400 x 218.9 =	92,463,360	i.	Brine 401,200 x 210.4 =	84,412,480
ii.	Purge water 900 x 50 =	45,000	ii.	Vapor 22,100 x 1163.3 =	25,708,930
iii.	Vapor 24,800 x 1166 =	28,916,800	iii.	Condensate 25,300 x 224.1 =	5,669,730
iv.	Vent 400 x 1169 =	467,600	iv.	Vent 400 x 1166 =	466,400
v.	Flash vapor 500 x 1166 =	583,000	v.	Heat transferred in Preheater HX-303a & b =	5,700,000
	Total	122,475,760		Total	122,357,540

The percentage difference is 0.096%, so enthalpy balance is excellent.

Now,

$$
\begin{aligned}
Q_E &= \text{H vapor} - \text{H condensate} - \text{H vent} - \text{H loss} - \text{H preheat} \\
&= 29,967,400 - 5,669,730 - 466,400 - 400,000 - 5,700,000 \\
&= 17,731,270 \text{ Btu/hr}
\end{aligned}
$$

Hence,

$$
\begin{aligned}
U &= \frac{17,731,270}{3960 \times (256.1 - 248.7)} \\
&= 605 \text{ Btu/hr ft}^2 \ {}^{\circ}\text{F}
\end{aligned}
$$

The reported value of U is 620 Btu/hr ft^2 ^{o}F, which is approximately 2.5 percent higher than the value calculated above.

- H & M Balance for Effect IV. The material balance closed excellent for this effect with input and output both being equal to 425,100 lb/hr.

Enthalpy Balance

Enthalpy In, Btu/hr		Enthalpy Out, Btu/hr	
i. Brine 401,200 x 210.4 =	84,412,480	i. Brine 378,700 x 201.1 =	76,156,570
ii. Purge water 900 x 500 =	45,000	ii. Vapor 23,400 x 1160.1 =	27,146,340
iii. Vapor 22,600 x 1163.3 =	26,290,580	iii. Condensate 23,000 x 215.9 =	4,965,700
iv. Vent 400 x 1166 =	466,400	iv. Vent	none
v. Flash vapor 200 x 1163.1 =	232,660	v. Heat losses	300,000
		vi. Heat transferred in Preheater HX-304	2,600,000
Total	111,169,460	Total	11,168,610

Enthalpy balance is again balanced to within 850 Btu/hr so that it can be considered excellent.

Now,

$$Q_E = 26,989,640 - 4,965,700 - 300,000 - 2,600,000$$
$$= 19,123,940 \text{ Btu/hr}$$

Hence,

$$U = \frac{19,123,940}{3950 \times (247.9 - 239.6)}$$

= 583 Btu/hr ft^2 ^{o}F as compared to the reported value of 590 Btu/hr ft^2 ^{o}F.

- H & M Balance for Effect V. The material balance closes with no error for this effect.

Enthalpy Balance

Enthalpy In, Btu/hr		Enthalpy Out, Btu/hr	
i. Brine 378,700 x 201.1 =	76,156,570	i. Brine 356,900 x 192.6 =	68,738,940
ii. Purge water 900 x 50 =	45,000	ii. Vapor 22,500 x 1157.2 =	26,037,000
iii. Vapor 23,400 x 1160.1 =	27,146,340	iii. Condensate 24,000 x 206.8 =	4,963,200
iv. Vent	None	iv. Vent 400 x 1160.1 =	464,044
v. Flash vapor 10000 x 1160.1 =	1,160,100	v. Heat losses	300,000
		vi. Heat transferred in Preheater HX-305	3,800,000
Total	104,508,010	Total	104,303,184

The difference in enthalpies in and out is approximately 0.2%. The heat transferred across the tube surface in the effect is:

$$Q_E = 28,306,440 - 4,963,200 - 464,044 - 300,000 - 3,800,000$$

$$= 18,779,196 \text{ Btu/hr}$$

Hence,

$$U = \frac{18,779,196}{3960 \times (238.8 - 231.4)}$$

$$= 640.8 \approx 641 \text{ Btu/hr ft}^2 \ {}^{o}\text{F}$$

This is 1.4 percent lower than the reported value of 650 Btu/hr ft^2 ^{o}F.

The heat and material balances for the remaining effects were not checked in detail since both the overall heat and material balance and individual balances for the first five effects have been satisfactory. Values of U were, however, calculated for all the effects based on the reported values of the heat transfer rate Q_E. These and other related data are given in tabular form later in this report.

d. Salt Balance Basis for Calculation of U — The accuracy of the calculated value of the overall coefficient U depends directly on the accuracy of the heat transfer rate Q_E, and on the ΔT temperature driving force. There are two methods by which Q_E can be calculated — one involving salt balance, and the other involving detailed heat and material balance data. The results obtained by applying both of these methods to data from Effect I, Run 16.3, given above, indicate an error of less than 0.2% although normally it is stated that salt balance values are accurate to within 10 to 15%. The low error obtained in salt balance method might be just a coincidence or a result of mutual cancellation of errors. Therefore, the following calculation is done on Effect II to check this.

- Salt Balance Basis. Heat transfer rate Q_E will be obtained from the evaporation rate which is given by equations 4.21 and 4.22.

$$E_n = L_{n-1} - \frac{(SWF)\,(Co)}{C_n}$$

$$= 446{,}500 - \frac{(472{,}000)\,(2.26)}{2.52}$$

$$= 23{,}198 \text{ lb/hr}$$

Part of this vapor comes from flashing of the high temperature inlet brine. This is given by:

$$\frac{(446{,}500)(0.982)\,(266.0 - 256.9)}{940.8} = 4241 \text{ lb/hr}$$

Hence, the heat transferred, Q_E, is given by:

$$
\begin{aligned}
Q_E &= (23{,}198 - 4241)\ (940.8) \\
&= (18{,}957)\ (940.8) \\
&= 17{,}834{,}746 \text{ Btu/hr}
\end{aligned}
$$

Therefore,

$$
\begin{aligned}
U &= \frac{17{,}834{,}746}{(3660)\ (265.2 - 256.9)} \\
&= 587 \text{ Btu/hr ft}^2\ {}^{\circ}\text{F}
\end{aligned}
$$

From detailed heat and material balance, Q_E was calculated to be 19,526,130 Btu/hr and U = 642 Btu/hr ft^2 $^{\circ}$F (see Effect II calculations). Thus the salt balance value is 8.6% lower. The calculated U value is also sensitive to the accuracy in determining ΔT. This ΔT measurement was reported (FY-1969 Report) to be accurate to $\pm$ 0.5°F, which is 5% of the average 10°F driving force in the effects. Therefore, even if Q_E values were accurate, the U values would be accurate to $\pm$ 5% only. Detailed heat and material balance and accurate ΔT values are essential to obtain accurate overall coefficients.

3. Presentation of Overall Coefficient Data

Since the detailed calculations for Development Runs 9-2 and 16-3 above indicate satisfactory general agreement with the calculated values, it is assumed that the remaining reported data has similar consistency and reliability.

As discussed earlier in the theory section, the heat transfer coefficient in a multiple-effect falling-film evaporator is a function of several design and operating variables. If an analysis is to be made of these coefficients for a given effect under different conditions (in several runs) then information of these variables and parameters should be included for each set of conditions. A sample of such background data is given in Table 4.2, entitled "Heat Transfer Performance — Development Run 16.3". Considerable effort was involved in collecting and calculating the presented data from widely scattered sources in Stearns-Roger operational reports on the 17-effect plant. A summary of the

Table 4.2

Heat Transfer Performance - Development Run 16.3

17-Effect Evaporator

PERFORMANCE SUMMARY

Net performance ratio : 12.56 (DOW STEAM)/11.21 (SAT. STEAM)
Extraction Ratio : 0.685
Overall plant 'ΔT', °F : 203.5 (Steam temperature - SWF temperature)
'ΔT' actually utilized, %: 91

MATERIAL BALANCE SUMMARY

Feed		Products	
Seawater	473,000 lb/hr	Water	349,800 (gross) lb/hr
DOW steam	25,800 lb/hr	Vent Steam	2,200 lb/hr
TOTAL	498,800 lb/hr	Brine Blowdown	144,700 lb/hr
		TOTAL	498,800 lb/hr

EFFECT NO.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	COMMENTS
Brine T/P/Conc.	266/38.7/2.39	256.9/33.1/2.52	248.7/28.8/2.64	239.6/24.4/2.80	231.4/21.0/2.96	223.4/18.1/3.13	223.4/18.1/3.13	204.7/12.5/3.55	195.7/10.3/3.79	184.3/8.1/4.05	172.8/6.2/4.38	164.8/4.9/4.72	153.3/3.9/5.08	140.4/2.8/5.53	126.5/1.95/6.07	113.1/1.34/6.65	98.3/0.915/7.38	T in °F, P in PSIA, C in WT%
Q_E, MM Btu/hr	25.1	19.9	18.1	18.1	19.0	17.6	17.1	16.6	16.8	15.7	17.1	16.5	15.5	15.4	14.5	14.4	12.7	Heat transferred/tubes
Area, A_o, ft^2	4070	3660	3960	3960	3960	3970	3970	3660	3640	3960	3970	4430	1720	4140	4200	3610	3840	Heat transfer area
'ΔT', °F	9.5	8.3	7.4	8.3	7.4	7.1	8.2	8.7	8.1	10.4	10.5	6.9	7.9	11.7	12.7	12.1	13.5	Temp. driving force
'U' reported	650	630	620	590	650	620	530	520	570	390	410	540	1140	320	270	330	250	Overall heat transfer coefficient
'U' calculated	649	655	617	585	648	624	525	521	570	381	410	540	1141	318	272	330	245	
Re_{Li}/Re_{Lo}	21260/20150	--	--	--	14060/13230	--	--	11850/11120	--	--	--	--	--	--	--	--	1936/1762	Brine entrance & exit re. no.
Prandtl Number	1.40	1.57	--	--	1.74	--	--	1.87	--	2.34	--	--	--	--	--	--	4.99	Brine Prandtl Number
Brine Rate/Tube	1457 lb/hr	1538	1344	1281	1205	1130	1066	1088	1025	883	823(1000?)	853(1000?)	1489(1000?)	573(1000?)	518	691	714	Brine feed rate/tube
Vapor Rate/Tube	860.6 ft^3/hr	1068.7	1001.9	1234.8	1378.6	1467.8	1663.8	2108.5	2488.3	2742.7	3746.7	4819.2	10814.4	6113.7	7974.4	15961.8	23537.2	Vapor leaving tube
Venting Mode	CASCADE	CASCADE	CASCADE	TO ATMOSPHERE	CASCADE	CASCADE	CASCADE	CASCADE	CASCADE	CASCADE	CASCADE	TO PRE-COND.	CASCADE	CASCADE	CASCADE	TO DEAERATOR	TO HX-318	
Brine Feed Mode	FEED FORWARD	FEED FORWARD	FEED FORWARD	FEED FORWARD	FEED FORWARD	FEED FORWARD	FEED FORWARD	FEED FORWARD	FEED FORWARD	FEED FORWARD	SUMP-TO-SUMP	SUMP-TO-SUMP	SUMP-TO-SUMP	SUMP-TO-SUMP	FEED FORWARD	FEED FORWARD	FEED FORWARD	
Brine Distributor	--	--	--	--	S.V.L. Weir	--	--	--	P.P. SVL Weir	--	Porcelain Nozzle	--	Porcelain Nozzle	--	--	--	--	V-Notch Weirs other Effects
Number of Tubes	310 + 5 + 5 + 4	291	315	314	315	316	316	291	290	315	316	282	150	360	366	252	223	
Tube Material/gage	A.B., 304SS, 306SS, T_i	ADM., 16	A.B., 16	Cu.Ni-90/10,16	A.B., 16	A.B., 16	A.B., 16	A.B., 16	Titanium, 25	A.B., 16	A.B., 16	Cu. Ni-90/10,18	C.D.A. 194/16	316 S.S./20	A.B./16	Cu.Ni-90/10,18	Cu.Ni-90/10,18	A.B. = Aluminum Brass, ADM = Admirald, S.S. = Stainless steel
Tube O.D./Length	16 18 16 0.041 2"/24'	2/24	2/24	2/24	2/24	2/24	2/24	2/24	2/24	2/24	2/24	3/20	3 1/8/11	2/22	2/22	2 1/2/22	3/22	
Tube Resistance, 10^6x	93.4	84.6	93.4	208.3	93.4	93.4	93.4	93.4	170.1	93.4	93.4	157	35.9	310	93.4	157	157	Resistance/Area
Tube Resistance, %	6.1	5.3	5.8	12.3	6.1	5.8	5.0	4.9	9.7	3.6	3.8	8.5	4.1	9.9	2.5	5.2	3.9	Total resistance based on 'U'
$\Delta P/\Delta T_p$	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	Pressure drop & corres. temp.
'U' corrected	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	Drop in tubes corrected 'U'
Scaling	--	--	--	--	--	--	--	--	--	--	--	--	--	22 tubes scaled	--	--	--	Based on corrected 'ΔT'

material balance and the performance factors for the entire plant are included in the table. Figure 4.6 shows graphically the U variation with brine temperature for all the effects in Run 16-3.

History of the overall heat transfer coefficient for all the development runs (1 through 16) is presented in tabular form in Table 4.3, entitled "Thermal Performance of the Freeport Evaporator." Development Run No. 1 started on February 6, 1964 and the final Run 16 ended on May 2, 1969. Overall coefficients were not available for the plant operational period prior to 1964, and for Run Numbers 3, 4, and 8. In some cases, U data was given for only selected effects.

For a rough comparison of overall coefficients, data are given in this table for the Freeport Plant observed values, the Wrightsville Beach pilot plant, and the Stearns-Roger values for the 5-effect module.

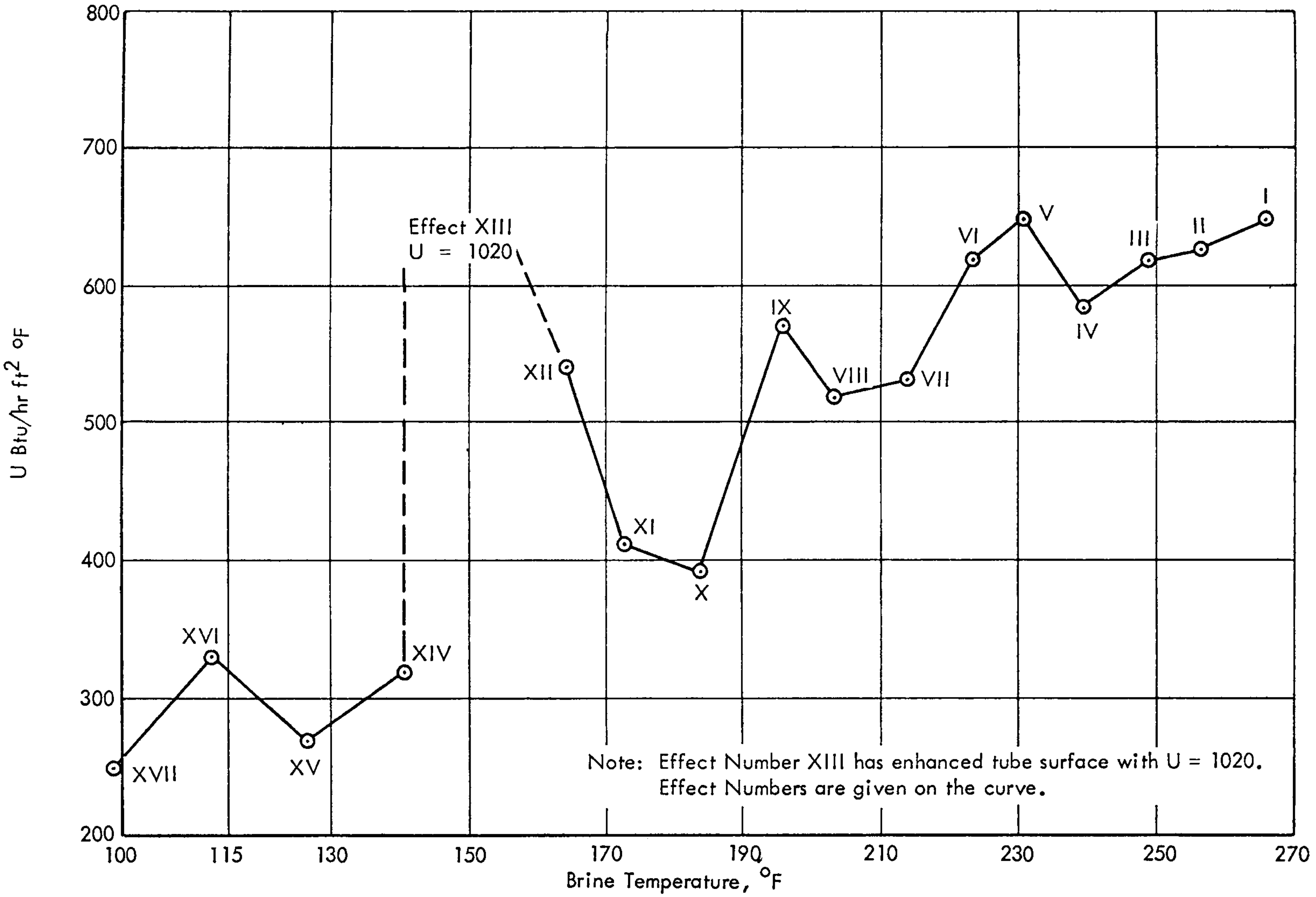
The history of overall coefficient by effect for all the runs is also presented in graphical form in Figures 4.7, 4.8, 4.9, and 4.10.

It is seen from the thermal performance table (Table 4.3) that for the majority of effects there is good agreement between the pilot plant data and the corresponding Freeport effect values. Close agreement between these two sets of data is not expected because the important variables of the Freeport Plant (temperature, pressure, flow rates, etc.) are not relative in magnitude to those in pilot plant operation. Since the pilot plant data has been shown to agree with 5% of the theoretical predictions (OSW R&D Progress Report Number 74), it can be said then that the Freeport data is fairly close, in general, to the theoretical performance expected.

In certain runs some of the effects did show deviations from this general pattern:

- Effect X/Run Number I where U = 330 as against the pilot plant value of U = 467. A low value (β = 5.23) of interfacial shear may partially account for this low value of U. The pilot plant run had β = 7.95.

- Effects XI, XII, and XIV/Runs 2 through 16. Effect XI gave U = 323 in Run 11-1 as against U = 467 for the pilot plant. The Fifth and Sixth Annual Reports by Stearns-Roger (OSW R&D Report Numbers 253 and 440) do not discuss the reasons for this decline but it appears that reduced brine feed rate per



Note: Effect Number XIII has enhanced tube surface with U = 1020.
Effect Numbers are given on the curve.

Figure 4.6. Run Number 16-3, 17-Effect Heat Transfer Rates

Table 4.3

Thermal Performance of the Freeport Evaporator

Development Run Number	Effect No.																	Data Source	Brine Temperature Range Effect 1 - XVII	Comments
	I	II	III	IV	V	VI	VII	VIII	IX	X	XI	XII	XIII	XIV	XV	XVI	XVII			
1	784					607				330	620	619	12 Effects only for Runs 1 to 12 Inclusive					Third Annual Report by Stearns-Roger (1964)	230 - 121 °F	U values not reported for all effects.
2	505					586				411	340	361						Development Report Number 2.	232 - 109 °F	Selected U values only.
3																				U values not reported.
4																				U values not reported.
5-a	666					948				480	411	295						Development Report Number 4 (1966)	240 - 120 °F	Scaling in Effects XI and XII.
6-11	757	684	759	704	633	637	628	472	523	503	395	336						Development Report Number 5 (1967), OSW R&D #297	261 - 145 °F	Scaling in Effect XII (expected due to high concentration factor in XII)
7-b2	664	711	778	778	687	642	643	588	491	393	310	328						Development Report Number 5 (1967), OSW R&D #297	279 - 122 °F	Scaling in the center tubes of Effect I.
8																				Development Report Number 6 contains one U value only
9-2	748	784	710	713	650	688	671	620	475	415	377	229						Fourth Annual Report by Stearns-Roger	267 - 121 °F	U values computed from reported data.
10-3	730	727	651	580	582	655	587	541	455	336	331	229						Fifth Annual Report by Stearns-Roger (1967)	265 - 121 °F	
11-1	604									456	323	347						Sixth Annual Report by Stearns-Roger (1969)	258 - 123 °F	Limited U data only. 3' O.D. tubes in Effect XII.
12-3	712									402	474	304						Sixth Annual Report by Stearns-Roger	261 - 110 °F	3" O.D. tubes in Effect XII
13-4													555	422	407	400	309	New module performance H.M.B. 13-4 by Stearns-Roger	171 - 127 °F	U values reported for the 5-effect module only.
14	730	600	530	750	500	660	600	700	430	360	390	350	530	340	330	430	350	Eighth Annual Report by Stearns-Roger (1970)	268 - 90 °F	Average U values for several sub-runs.
15	670	610	610	650	620	650	610	520	610	400	360	320	1180	410	360	360	330	Eighth Annual Report by Stearns-Roger	269 - 105 °F	Average U values for all series 15 runs.
16	620	620	660	610	570	560	510	540	580	390	380	530	1040	300	270	300	270	Eighth Annual Report by Stearns-Roger	266 - 94 °F	Average U for runs 16-1, 16-2, 16-3.
W.B. Pilot Plant	721	741		727		666		532		467	407	377	542 (smooth tube)	502	465	426	342	Specifications 195 by OSW (1960), p. IV-22. "New Module Performance" H.M.B. 13-4 by Stearns-Roger	242 - 141 °F 162 - 115 °F for Effect 13-17	W.B. = Wrightsville Beach, OSW R&D Reports 26 and 456. This data for rough comparison.

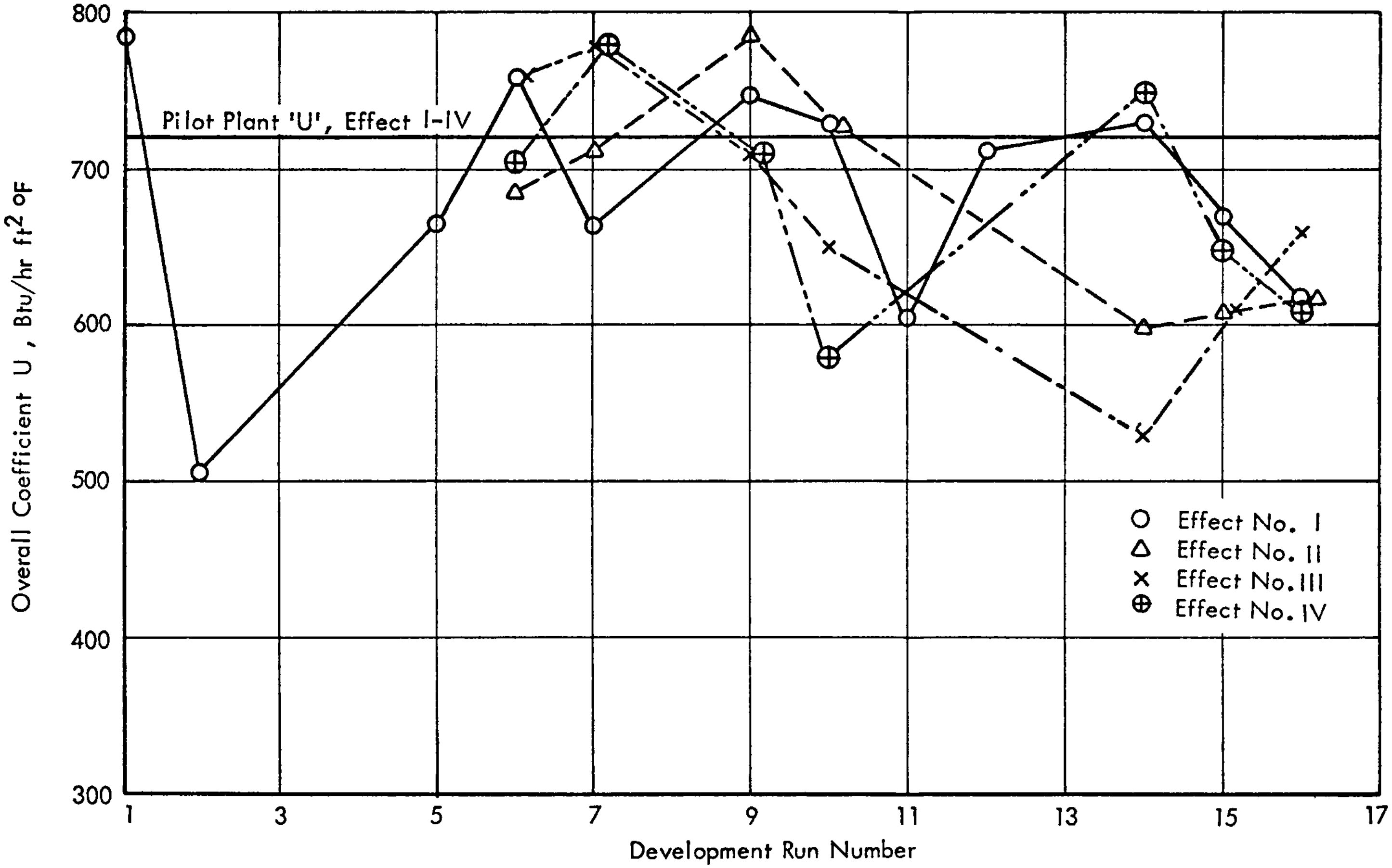


Figure 4.7. History of Overall Coefficients by Effect

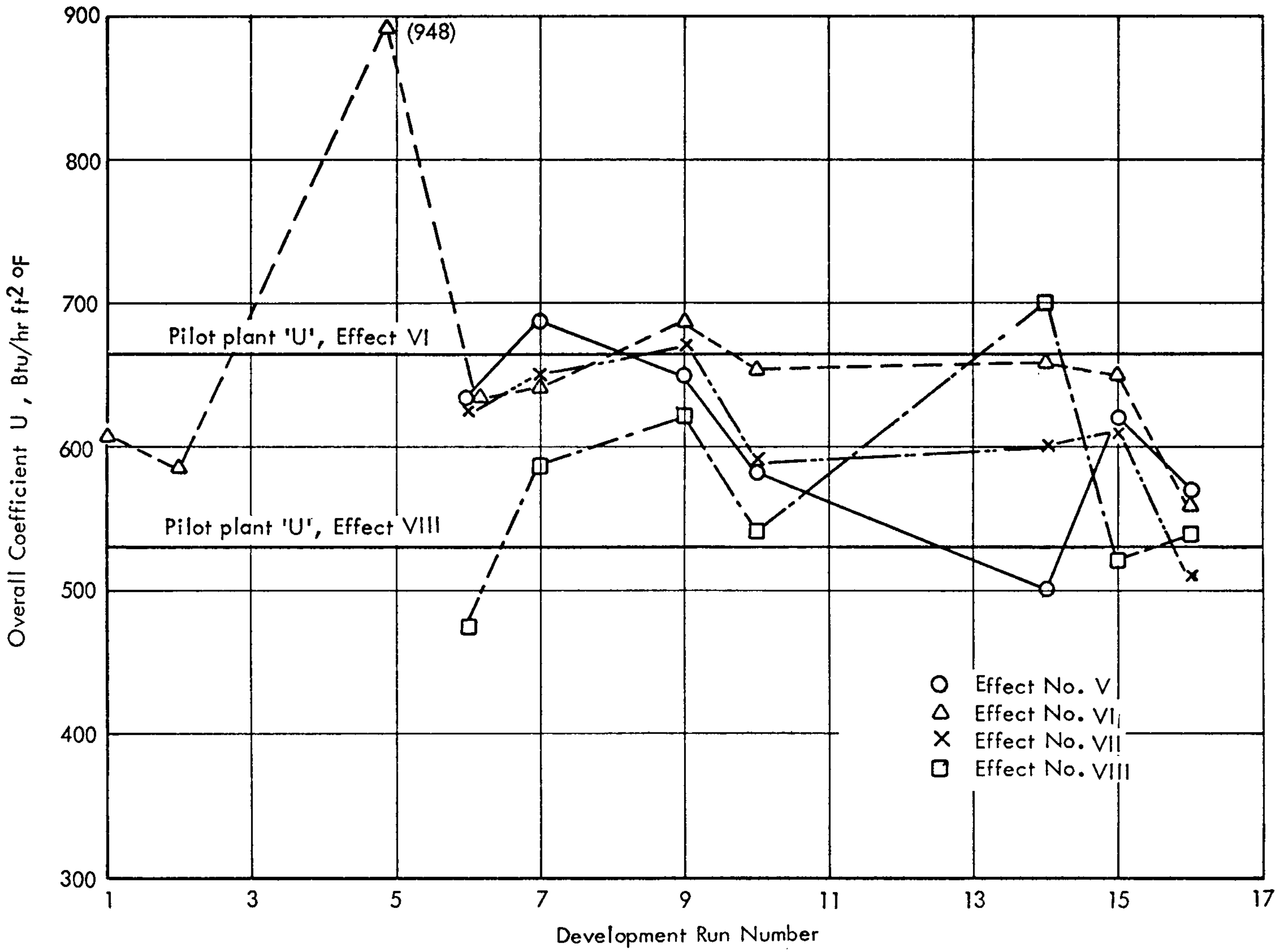


Figure 4.8. History of Overall Coefficients by Effect

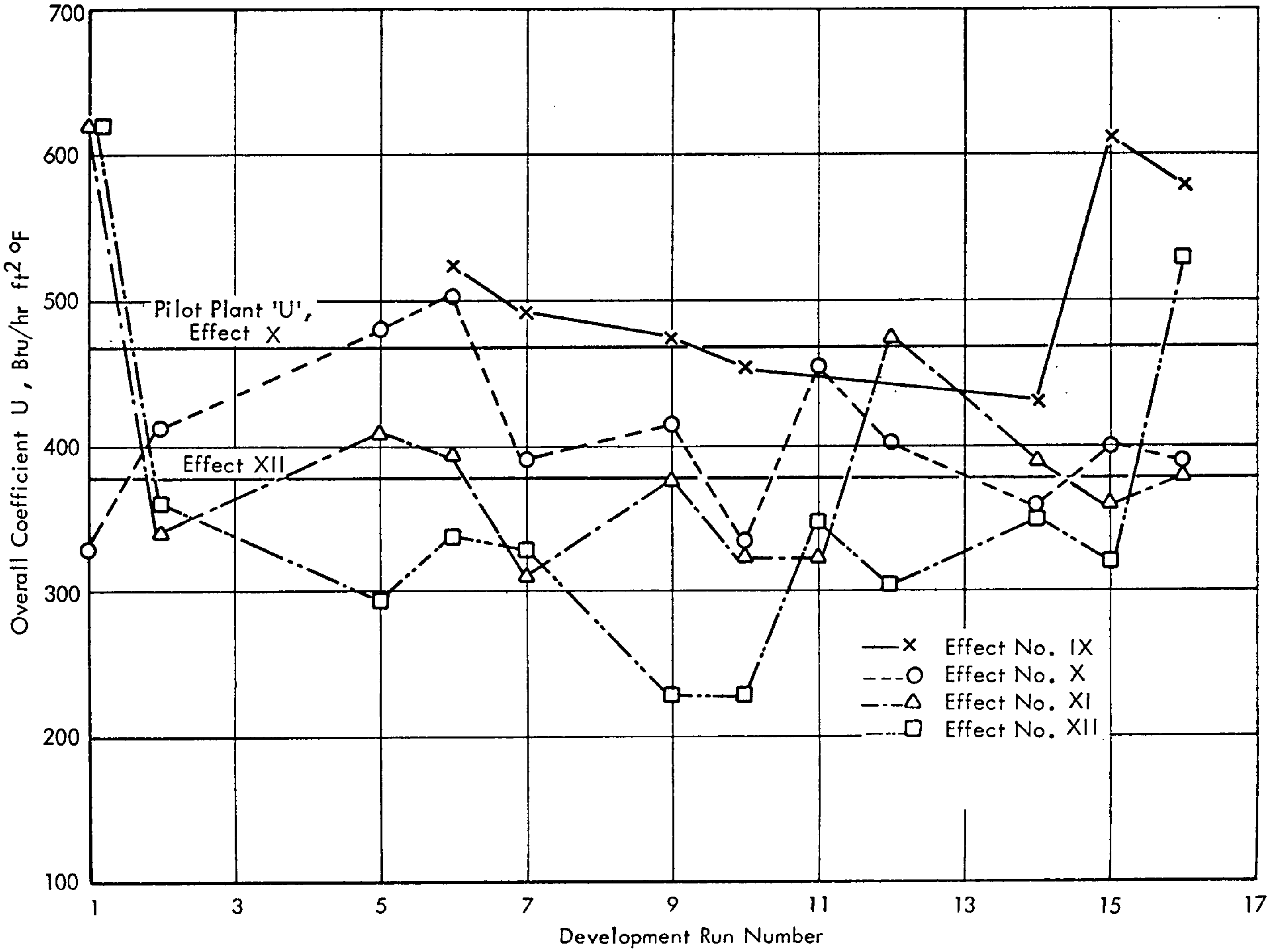


Figure 4.9. History of Overall Coefficients by Effect

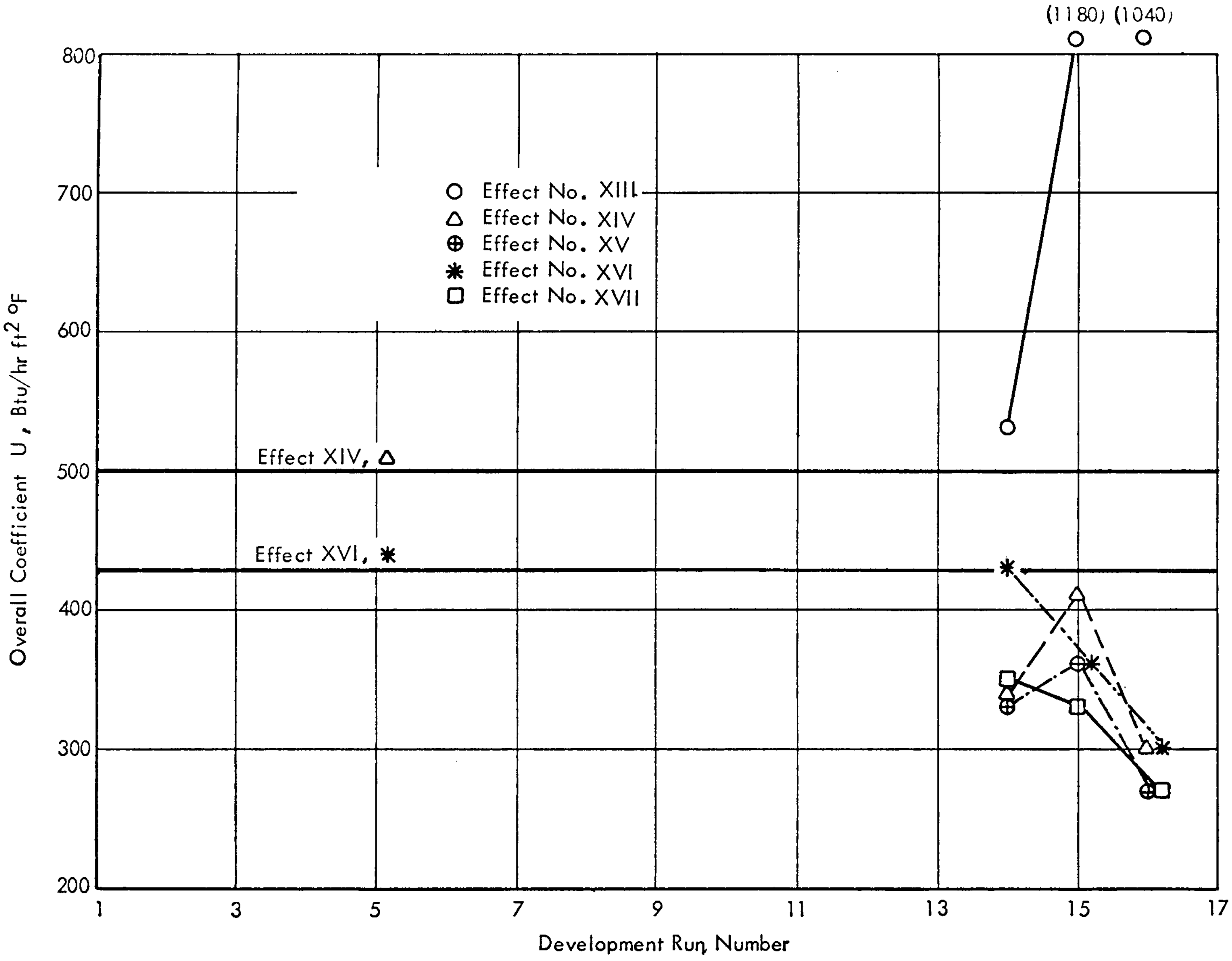


Figure 4.10. History of Overall Coefficients by Effect

tube (670 lb/hr per tube, page 149 OSW R&D Report Number 253) may have contributed to this low performance. Because this effect operates under vacuum air, in-leakage is another possibility.

- Effect XII/ Runs 9-2 and 10-3. A very low value of U = 229 was reached in both runs. Fourth Annual Report (page 118, paragraph 5-104) lists "probable existence of actual liquid levels and/or subcooling of the condensate" as the factors leading to reduced reliability of the heat transfer data. The steam distribution was said to be poor (page 118, paragraph 6-119 in Fifth Annual Report) due to deactivation of many tubes in Effect XII.

- Effect XIV/Run Number 16. A U = 300 was recorded against the design value of 502 in this run. Inadequate steam distribution, in-leakage of non-condensibles and ineffective venting are stated to be the reasons for this low performance. Frequent opening and closing of the effect required in development-type operation perhaps weakens the effective sealing of joints so that under very low (vacuum) pressure conditions existing in this effect, the in-leakage of air is serious enough to adversely affect the heat transfer rates.

Although in some cases, as noted above, the heat transfer performance has been below that expected from pilot plant data, it is to be noted that in quite a few runs the U values were more than expected. This can be seen from the performance table (Table 4.3) and also from the attached graphs (Figures 4.7, 4.8, 4.9, and 4.10). The double-fluted tubes used in Effect XIII in Run Numbers 15 and 16 yielded U values of 1180 and 1040 (Btu/hr ft^2 ^oF), which are less than the design value of 1250 but almost twice as much as the Run Number 14 smooth tube value of 555. Air in-leakage was attributed as the probable cause for the less than expected performance of this effect.

C. EFFECT OF DISTRIBUTION DEVICES ON U

Each effect of a multiple effect system receives in its top water box the entire amount of brine (or seawater) from the previous effect. In the case of the Freeport Plant with its rated capacity as 1 million gallons per day, the seawater feed to the first effect is about 500,000 pounds per hour. In a 12-effect evaporator plant the last effect water box has a feed of about 192,000 pounds per hour so that even though the amount of brine is decreasing from effect to effect, it is still a large quantity to be properly distributed to some 350 to 500 tubes in each effect. If this feed brine is not properly distributed, then the following undesirable effects are obtained:

- Starvation of some or most of the tubes while flooding of the remaining tubes takes place.
- Scaling of starved tubes from solids deposition after the water evaporates off. Since these deposited solids act as nuclei, additional scaling might result even if the normal feed rate were restored. Scaling of tubes drastically reduces the overall coefficient U.
- Flooded tubes short circuiting some of the feed from the waterbox directly (i.e., without being concentrated) to the sump. Capacity and thermal economy are adversely affected.
- Loss in capacity resulting from the uneven feed distribution and reduction in heat transfer rate due to scale formation.

In short, the even distribution of feed brine to all the tubes in the water box of an effect is essential for better capacity and efficiency, and smoother operation of the plant.

1. Original Distribution Devices

As built Drawing Number 7 by Chicago Bridge and Iron Company gives details of the distributor plate used initially at Freeport. Made from carbon steel, the plate had 600 holes of 3/8 inch diameter. The uneven distribution of liquid on this plate is shown in Figures 3-4, 3-5, 3-6, and 3-7 in the Third Annual Report (OSW R&D Report Number 123). Figure 4.11 (Figure 3-6, Third Annual Report) is reproduced here to illustrate the flow pattern existing in normal brine flow. Note the central area of the distribution plate where horizontal brine velocity is very high and its thickness correspondingly low so that most of these central tubes are starved. The peripheral tubes are, on the other hand, being flooded at the same time as seen in Figure 4.11. Actual individual tube outflow measurements (on Effect XI, see Figure 4.12 — OSW R&D Report Number 479 — and Table 4.4) confirmed this pattern.

Since the distribution plates alone were not efficient, V-notch or triangular notch weirs were inserted in each tube of Effects X and XII as shown in Figure 4.13 for testing during Development Run Number 1 (February 1964).

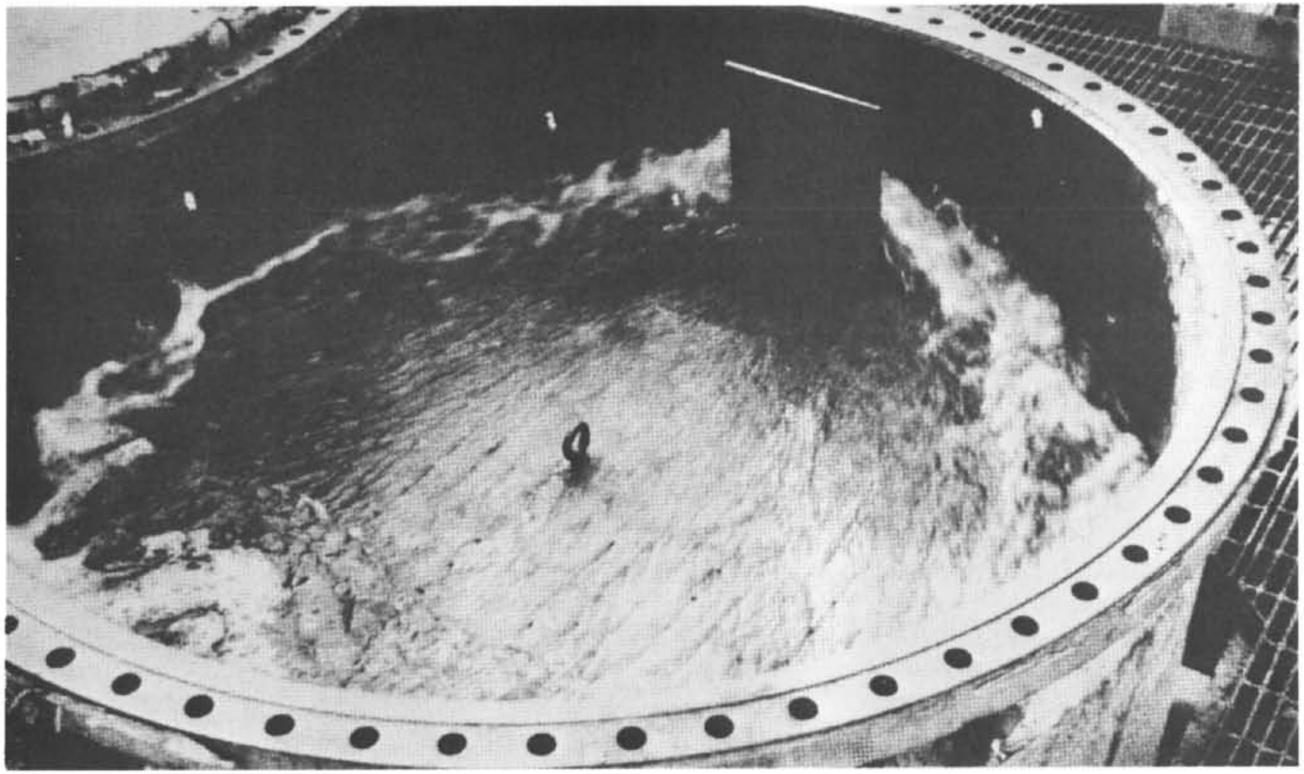


Figure 4.11. Liquid Head Distribution During Normal Flow

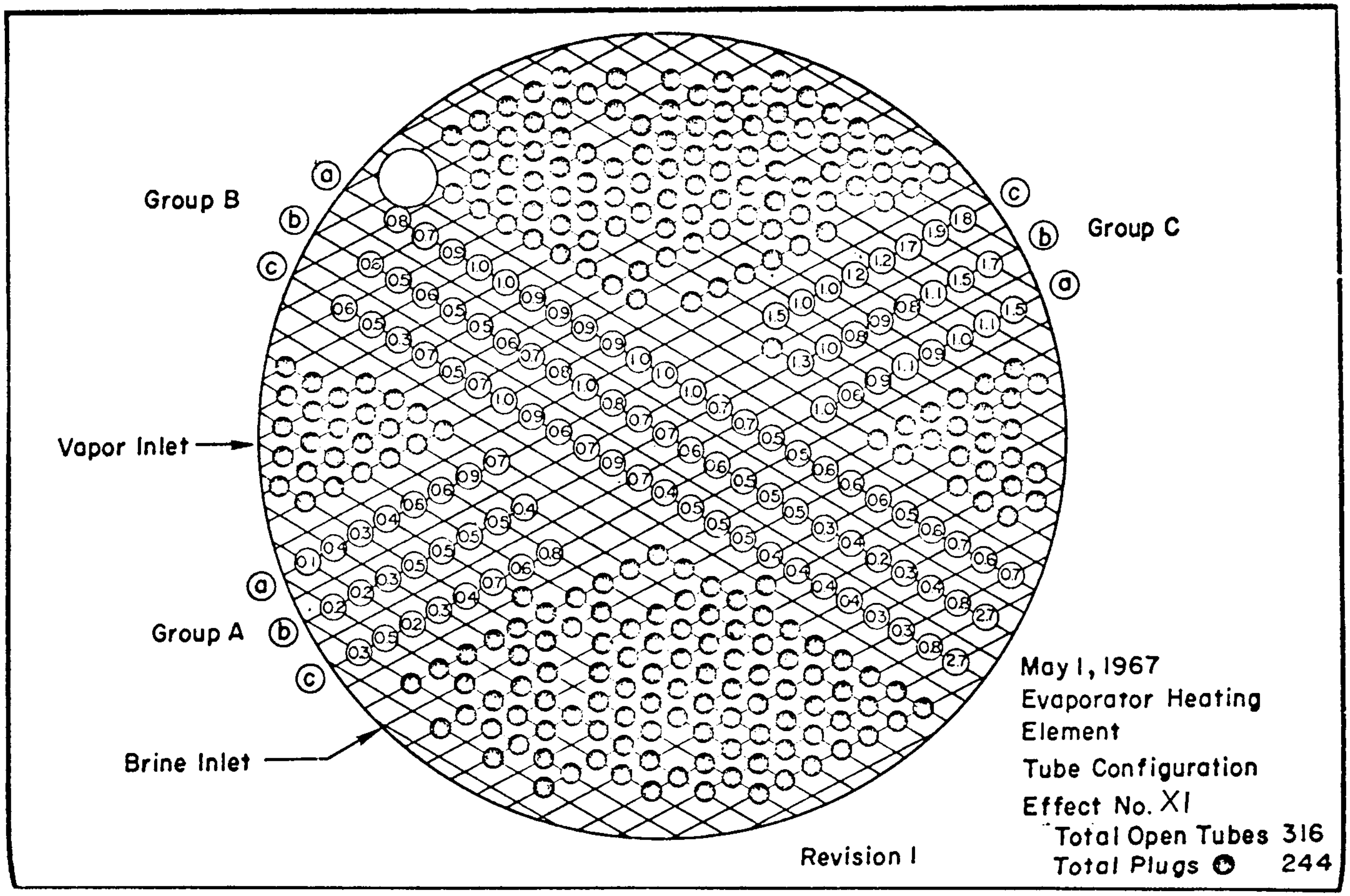
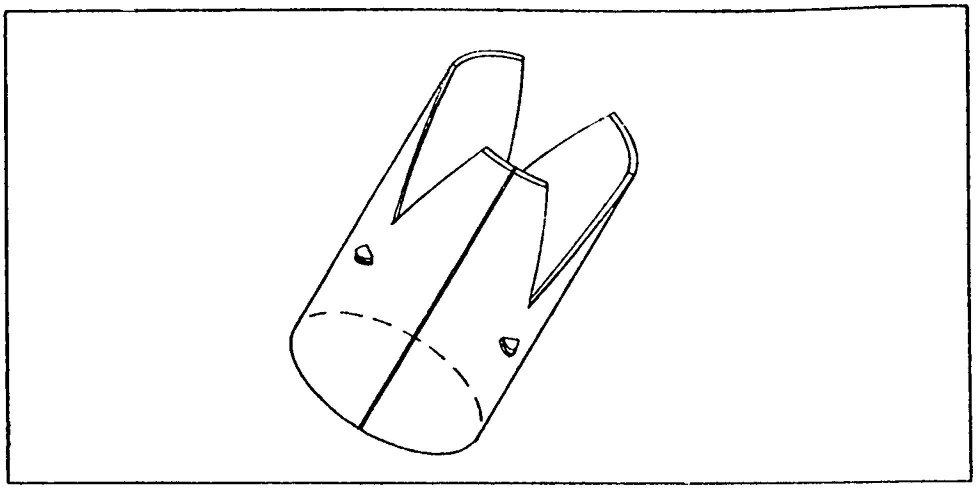


Figure 4.12. Tubes Sampled for Brine Flow Rate Distribution Pattern
Note: The numbers inside the circles represent the ratio of the actual tube flow rate to the average obtained as total brine feed divided by the number of tubes.

Table 4.4

Measurement of Brine Flow Through Individual Tubes in Evaporator XI - Development Run Number 14

DATE	1/4/68		1/10/68	
OPERATING MODE	FEED FORWARD AND MAXIMUM RECYCLE		FEED FORWARD AND MAXIMUM RECYCLE	
Sump Temperatures, $^{\circ}$F				
Effect 10 Sump	182.8		180.7	
Effect 11 Sump	174.6		171.2	
Feed Rates				
To Effect 11 Top Water Box, LBS/HR	300,000		270,000	
Average Tube Feed, LBS/HR/Tube	950		850	
Average Tube Feed, GPM/Tube	1.8		1.7	
RESULTS	NUMBER OF TUBES	PER-CENT	NUMBER OF TUBES	PER-CENT
Less Than 20 Percent of Average Tube Feed	0	0	1	1
20 to 40 Percent of Average Tube Feed	14	12	13	11
40 to 60 Percent of Average Tube Feed	30	25	30	25
60 to 80 Percent of Average Tube Feed	31	26	29	24
80 to 100 Percent of Average Tube Feed	21	18	19	16
100 to 120 Percent of Average Tube Feed	15	13	15	13
More Than 120 Percent of Average Tube Feed	8	6	12	10
TOTALS	119	100	119	100



Brine Distributing Weir

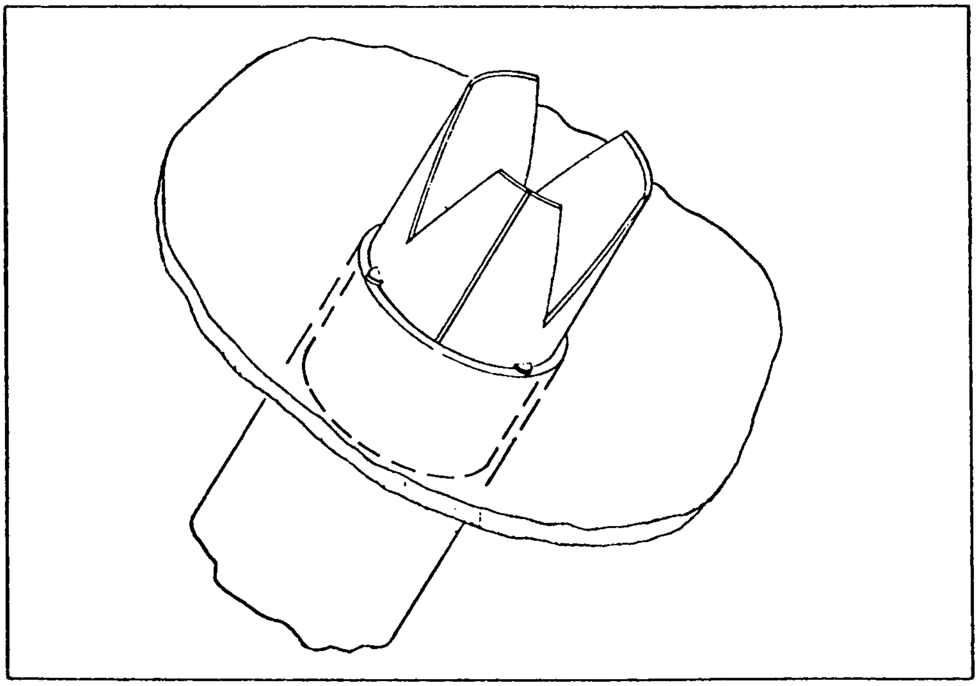


Figure 4.13. Brine Distributing Weir Inserted in Tube

The weirs were fabricated from sheet copper to close tolerances. The theory of the notched weirs is based on the stilling effect of a liquid head on the tube sheet and a sharp reduction of flow through any single notch with a reduction in head. The sharp notch and square edges cause a very low velocity of fluid at or near the bottom of the notch. This draws the stream to the tube metal internal surface rather than spilling to the center. The 1 1/2 inch depth of the notch was expected to provide even distribution of all expected flows without flooding. The design capacity of single notch was reported (Development Report Number 1) as 1.25 gallons per minute at a liquid head of 1 inch, or a capacity of 3.75 gpm per tube. This rate was somewhat greater than the maximum possible flow rate per tube with the then available equipment.

For proper functioning of these weirs the tube sheet must be level, and the height of all the notches above the tube sheet must be the same.

In the Sixth Annual Report by Stearns-Roger, the effectiveness of these V-notch weirs was reported (page 15, paragraph 2-42). The overall coefficient U was found to increase from 616 to 748 Btu/hr ft^2 ^oF when these weirs were installed in Effect I. Houston Research Institute did theoretical calculations for Dow Company's "An Engineering Evaluation of the Long-Tube Vertical Falling-Film Distillation Process" (OSW R&D Report Number 139) to investigate the effect of liquid maldistribution. In each of two cases studied, the overall flow rate was maintained normal. However, for one third of the tubes in the effect the local flow rate was increased; for the second third, held normal; and the final third, decreased. The flow rate change was ± 20% for Case I and ± 50% for Case II. Computer results indicated practically no change for the two cases from the even distribution case. This shows that maldistribution per se does not affect the overall coefficient — what is indeed affected is the possibility of scaling of the tubes. Those tubes which receive less than normal feed rate but have the same rate of heat transfer will produce more concentrated brine (than the tubes receiving the normal feed), thereby increasing their scaling potential. Actual scaling patterns have confirmed this. An increase of the tube feed rate by introducing a recycle of sump brine helps to reduce this scaling potential (OSW R&D Report Number 139, Appendix , page 263).

2. Additional Distribution Devices

Since precision manufacturing and installation are required when using the V-notch weirs, it was desirable to try other liquid distribution devices. Swirl-vane weir (SVL weir) and porcelain spray nozzles were two such devices tested at Freeport. Two types of materials were used for SVL weirs — polypropylene and stainless steel (316L). Polypropylene weirs developed cracks and some popped out of the tubes. The liquid distribution patterns are compared in Table 4.5 for the two types of weirs used in Effect XI and for two modes of brine transfer — feed forward and recycle or sump-to-sump modes. The spray nozzle distribution is seen to be much more even than that of the SVL weir. The spray nozzles, however, require a certain minimum flow rate (and pressure drop) for their operation; hence, where wide fluctuations in flow rates are expected, these nozzles may not be satisfactory. Higher-than-design flow rates are better handled by the spray nozzles since there is no problem of flooding. Effect XIII was equipped with 3 1/8 inch porcelain nozzles (supplied by Knox Porcelain Corporation, Knoxville, Tenn.) prior to Run Number 15 and their operation was satisfactory. The cost of these nozzles was approximately 1/3 the cost of stainless steel SVL weirs.

D. EFFECT OF PROCESS AND DESIGN VARIABLES ON U

The overall coefficient U depends on the individual film coefficients h_b (steam side), h_d (brine side), and the tube wall coefficient $\left(h_c = \frac{k}{x}\right)$. When noncondensible gases form a film on either (or both) steam or brine sides and mineral scaling occurs, their resistances must be added to the other film resistances $\left(\frac{1}{h_b} + \frac{1}{h_d}\right)$. The film coefficients are functions of Reynolds Number, Prandtl Number, interfacial shear β, and properties group ψ. Any (or all) process or design variables which affect the value of these numbers would alter the value of U. The following are the main design or process variables:

a. Tube diameter and length.

b. Tube spacing.

c. Steam temperature and brine temperature.

d. Brine flow rate per tube.

e. Vapor flow rate per tube.

f. Recirculation of sump brine.

g. Viscosity, density, solids concentration, specific heat, thermal conductivity.

Table 4.5

11 Effect Distribution Pattern

SUMMARY

Type Nozzles	SVL	Porcelain Spray
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	% of Tubes in Each Class	
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	4	3
More than 1.3	8	0
	100%	100%

1. Effect of Design Variables on U

Tube diameter and length, tube material, tube spacing, steam and brine temperatures, and properties of brine and steam are the principal design variables that affect the overall coefficient U.

a. Tube Diameter and Length — As discussed earlier (Section A) the tube diameter and length should be optimized for each effect to obtain the maximum heat flux ($q/A = U \Delta T$). At the Freeport Plant all tubes in the evaporator effects were initially 2 inch O.D. and 24 feet in length. Later 2 1/2, 3, and 3 1/8 inch O.D. tubes 22, 20 and 11 feet in length were tested in the low temperature effects (XII to XVII). (See Table 4.2.) The 3 1/8 inch O.D. 11-foot-long tubes were double-fluted, enhanced surface tubes installed in Effect XIII. The effect of specific change in tube diameter and length on U has not changed. However, theoretical calculations have been carried out by Houston Research Institute under subcontract with Dow Chemical Company and are reported in OSW R&D Report Number 139. The most dramatic influence on tube diameter and length was seen in Effect XII (Table 4.6 and Figure 4.14) as expected. Changing to larger diameter tubes or shortening tubes (relative to 2 inches - 24-foot-long tubes) caused a marked lowering in surface area required (Figure 4.14). At larger tube sizes, the effect of tube length becomes less important. Figures 4.14 and 4.15 clearly show the minimum total surface area for Effect XII to be attained when the tube diameter is approximately 4 inches, whereas this minimum for Effect I occurs when the tube diameter is approximately 2 inches. This is due to the opposite effects on U of decreased tube size and driving force ΔT. Interfacial shear β is increased due to decreased tube size and the resulting increase in U is less than the decrease in ΔT due to pressure drop. The net result is decreased flux and, hence, the 2 inch tube appears to be the desirable diameter for high temperature/high pressure effects.

b. Tube Spacing — Since the overall coefficient U depends on both the brineside and steamside film coefficients, it is essential to improve the controlling coefficient when the other has reached its highest limit. For example, if the brineside coefficient is maximized by a suitable change in tube size and flow rate, then the steamside coefficient is the controlling coefficient (in the absence of non-condensibles and scale). Efforts must be made to increase the steamside coefficient by increasing the velocity of steam past the outside of the tubes. This can be done by decreasing the open cross-sectional area on the steam side. Tube spacing may be decreased to decrease this open area. Computer results (OSW R&D Report 139) indicated that the increased interfacial shear β (and hence U) was negligible.

c. Tube Material — Although high values of tube material thermal conductivity are desirable, the possible improvement in U is small. This is true because the tube wall resistance is small relative to individual film resistances on the brine and steam sides, so that a decrease in an already small resistance does not improve U. For example, consider the case of 90 : 10 copper-nickel alloy, which has thermal conductivity $k = 26$ Btu/hr ft ^{o}F, and aluminum brass, $k = 58$ Btu/hr ft ^{o}F. If these tubes are used in an effect

Table 4.6

Program for LTV Evaporator Heat Transfer Calculations

— Summary of Results

Tube Size	Effect Number I					Tube Size	Effect Number VIII					Tube Size	Effect Number XII				
O.D. 16 BWG	Run Number	Tube Length-ft	Tube Number	Total Area-ft^2	In.Hg.	O.D. 16 BWG	Run Number	Tube Length-ft	Tube Number	Total Area-ft^2	In.Hg.	O.D. 16 BWG	Run Number	Tube Length-ft	Tube Number	Total Area-ft^2	In.Hg.
3/4"	107	20	1776	6974	1.423	1"	802	20	1411	7388	0.7672	2"	1204	8	1495	6262	0.0108
3/4"	108	24	1571	7403	2.401	1"	801	24	1296	8143	1.018	2"	1203	12	1001	6289	0.0331
3/4"	109	30	1389	8182	2.979	1"	803	30	1203	9448	1.343	2"	1202	16	762	6384	0.0709
1"	101	20	1245	6519	0.5628	1 1/4"	817	8	2373	6213	0.0408	2"	1201	20	626	6555	0.1237
1"	102	24	1063	6679	0.8769	1 1/4"	818	12	1584	6220	0.1247	2"	1200	24	542	6811	0.1875
1"	103	30	892	7006	1.446	1 1/4"	819	16	1218	6377	0.2605	100%	991	24	578	7263	0.2473
2"	104	20	611	6398	0.0535	1 1/4"	820	20	1020	6676	0.4345	200%	992	24	582	7314	0.2769
2"	105	24	509	6396	0.0886	1 1/4"	821	24	901	7076	0.6304	300%	993	24	585	7351	0.2959
2"	106	30	408	6409	0.1630	1 1/4"	822	30	800	7854	0.9221	3"	1208	8	996	6258	0.0029
100%	111	24	498	6258	0.1220	2"	808	8	1493	6254	0.0081	3"	1207	12	663	6249	0.0090
200%	121	24	493	6195	0.1446	2"	807	12	989	6214	0.0254	3"	1206	16	499	6271	0.0198
300%	131	24	490	6158	0.1642	2"	806	16	741	6208	0.0564	3"	1205	20	401	6299	0.0365
3"	112	24	340	6409	0.0236	2"	804	20	595	6230	0.1036	4"	1212	8	747	6258	0.0149
3"	113	30	271	6385	0.0442	2"	888	24	500	6283	0.1680	4"	1211	12	498	6258	0.0080
3"	110	20	409	6425	0.0142	2"	805	30	411	6456	0.2917	4"	1210	16	373	6250	0.0036
4"	114	20	308	6451	0.0056	3"	812	8	998	6270	0.0022	4"	1209	20	299	6262	0.0012
4"	115	24	256	6434	0.0094	3"	811	12	664	6258	0.0067						
4"	116	30	204	6408	0.0176	3"	810	16	497	6245	0.0149						
						3"	809	20	396	6220	0.028						
						4"	816	8	749	6276	0.0009						
						4"	815	12	500	6283	0.0027						
						4"	814	16	374	6266	0.0060						
						4"	813	20	299	6262	0.0111						

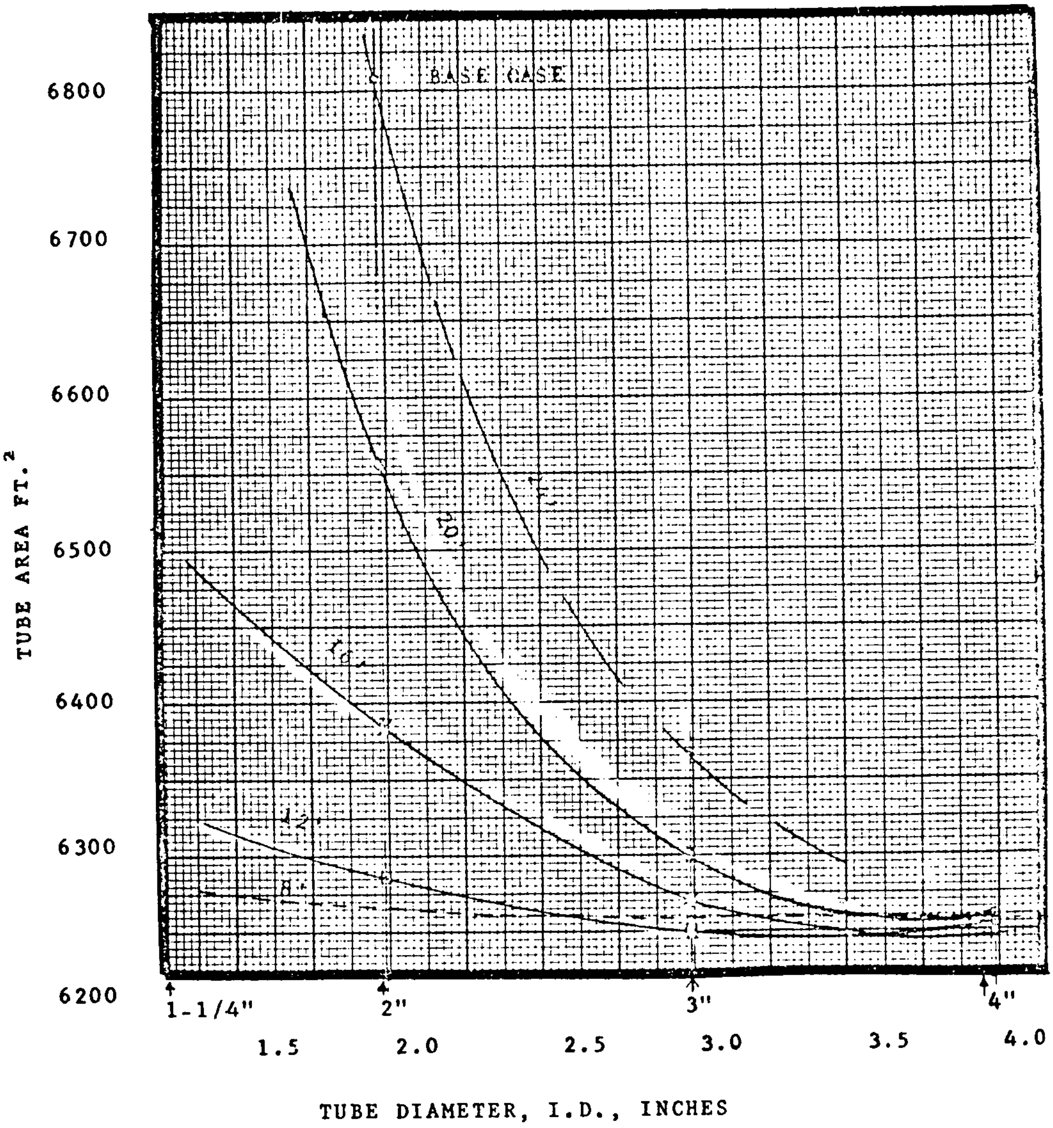


Figure 4.14. Total Area versus Tube Diameter - Effect Number XII

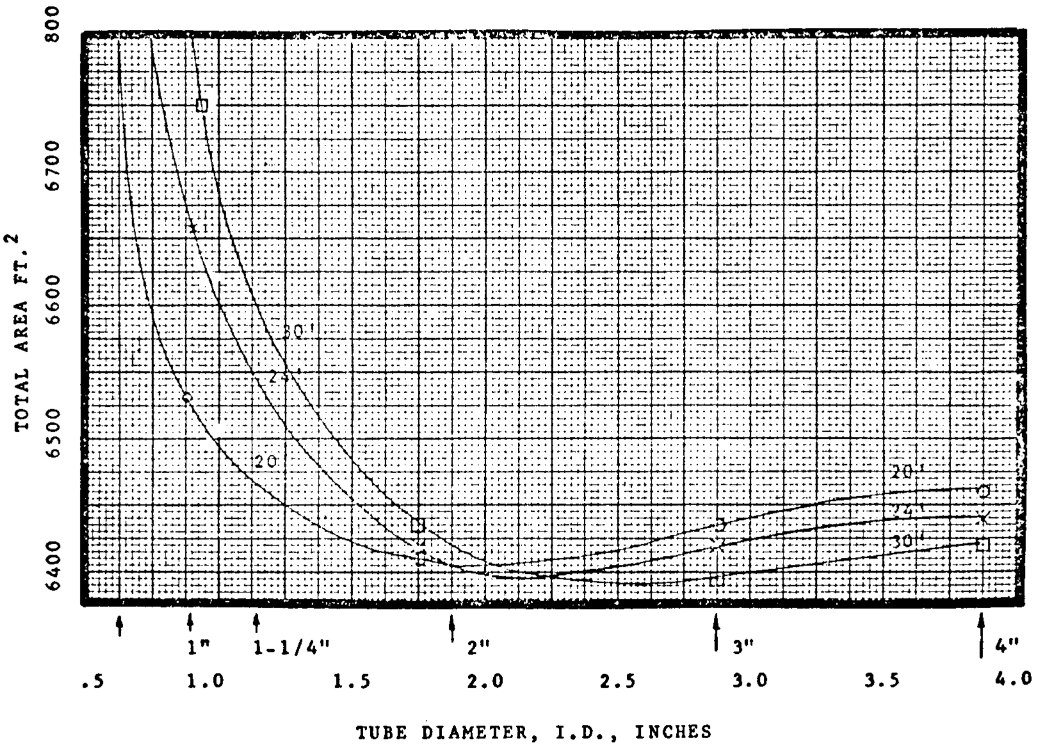


Figure 4.15. Total Area versus Tube Diameter - Effect Number I

where $U = 500$ Btu/hr x ft^2 $^\circ$F, then the overall resistance is $\frac{1}{U} = 0.002$. Therefore, the tube wall resistances would be:

$$\frac{(.065)(100)}{(12)(26)(.002)} = 10.41\% \text{ for copper-nickel alloy}$$

and

$$\frac{(.065)(100)}{(12)(58)(.002)} = 4.66\% \text{ for aluminum brass}$$

These percentages would be still smaller if individual coefficients ($h_b = 1250$ and $h_d = 1008$ approximately) were used. Even if the tube wall resistance were smaller, the tube material selection would have to be based on the material's corrosion and erosion resistance, structural strength, and cost. Initially, carbon-steel tubes were tried at Freeport, but their corrosion rate was high; so, the tubes were soon replaced with those of arsenical aluminum brass. Other tube materials tested were: 304 stainless steel, 306 and 316L stainless steel, titanium, C.D.A. alloy 194, and CuNi 90:10 alloy. More details on these are given in Chapters 2 and 6.

d. Steam and Brine Temperatures — The effect of absolute values of steam and brine temperatures on U is through their influence on the physical properties of steam and brine. The principal physical property affected by temperature is viscosity of liquid water (steam condensate) and brine films on the outside and inside surface of the tube. Increase in temperature significantly reduces the viscosity of water and brine. This results in thinner, viscous sublayers on the tube wall and, hence, an increase in heat transfer rates. Figures 4.2 and 4.6, where U has been plotted against brine temperature, illustrate this point. Since it is observed that increasing steam (or brine) temperature is conducive to an increase in U, a natural question arises regarding the limit to which a designer can raise this temperature. In a multiple effect plant only the first and last effect steam (and brine) temperatures are really under the control of the designer. In the intermediate effects, steam and brine temperatures are automatically fixed once the end values (first effect and last effect) are fixed. Now the limits on these end values are imposed by the scaling tendencies of seawater and the temperature of cooling water available. Corrosion rates are also accelerated at higher temperatures. The highest brine temperature (in the first effect) attainable without scaling has been found to be approximately 268°F for normal seawater.

2. Effect of Process Variables on U

The important process variables are:

a. Brine flow rates.

b. Vapor flow rates.

c. Recirculation of sump brine to top water box.

a. Brine Flow Rate versus U — The plant capacity determines the seawater feed rate to the first effect (amongst other things such as extraction ratio). Brine feed rates to other effects are determined by the amount of evaporation in the previous effects. The operator has no control over these flow rates. Brine feed rate per tube is then determined by the number of tubes in the evaporator (assuming adequate brine distributing devices such as spray nozzles or SVL weirs are provided on the tubes). The brine flow rate per tube determines its Reynolds Number in the tube as follows:

$$Re_L = \frac{4W_L}{\pi D \mu}$$

where

W_L = brine flow rate per tube

and
D = internal diameter of tube
μ = viscosity of brine

Increased flow rates per tube directly increase the brine Reynolds Number. The effect of Reynolds Number on the brineside coefficient h_d is given in Figures 7 through 9 of OSW R&D Report Number 74. These figures are reproduced here as Figure 4.16, 4.17, and 4.18. From these figures it is seen that up to a Reynolds Number of about 1000 there is actually a decrease in the brineside heat transfer coefficient. For $Re_L > 1000$ and for low values of interfacial shear β there is a slight increase in the coefficient, but at high values of β again there is a decrease in the coefficient with Reynolds Number. Thus, an increase or decrease in brine flow rates has only slight influence on U. This theory assumes that brine distribution is even, at low as well as at high brine rates.

Development Runs 14-6, 14-7, 14-8 and 14-9 were conducted to study the effect of reduced plant capacity on overall plant performance. Test capacities were set at 80, 60, 40 and 25 percent of design capacity. Blended brine feed rates were 412,000, 270,000, 204,000, and 120,000 lb/hr. Table 4.7 gives the overall coefficients along with effect steam temperatures for these reduced rate operations. Although at first sight it appears that the overall coefficient U is reduced at reduced brine feed rates, this is in fact not true. In Figure 4.19, these coefficients are plotted for 80 and 40 percent capacities against steam temperatures. It is seen from this figure that the coefficients for the two capacities fall in approximately the same region except for two points for Run 14-8. This means that the coefficient for a given effect is reduced due to the reduction in the steam temperature (or brine temperature) rather than due to the reduction in flow rates. As discussed above, theory predicts such a behavior. The reduction in coefficient can also arise partially due to reduced vapor (i.e., steam) rates, thus decreasing the steamside coefficient. The data of Runs 10-3 and 10-A5 (OSW R&D Report Number 253) where the latter run was at 60 percent of design rate can be viewed in the same light.

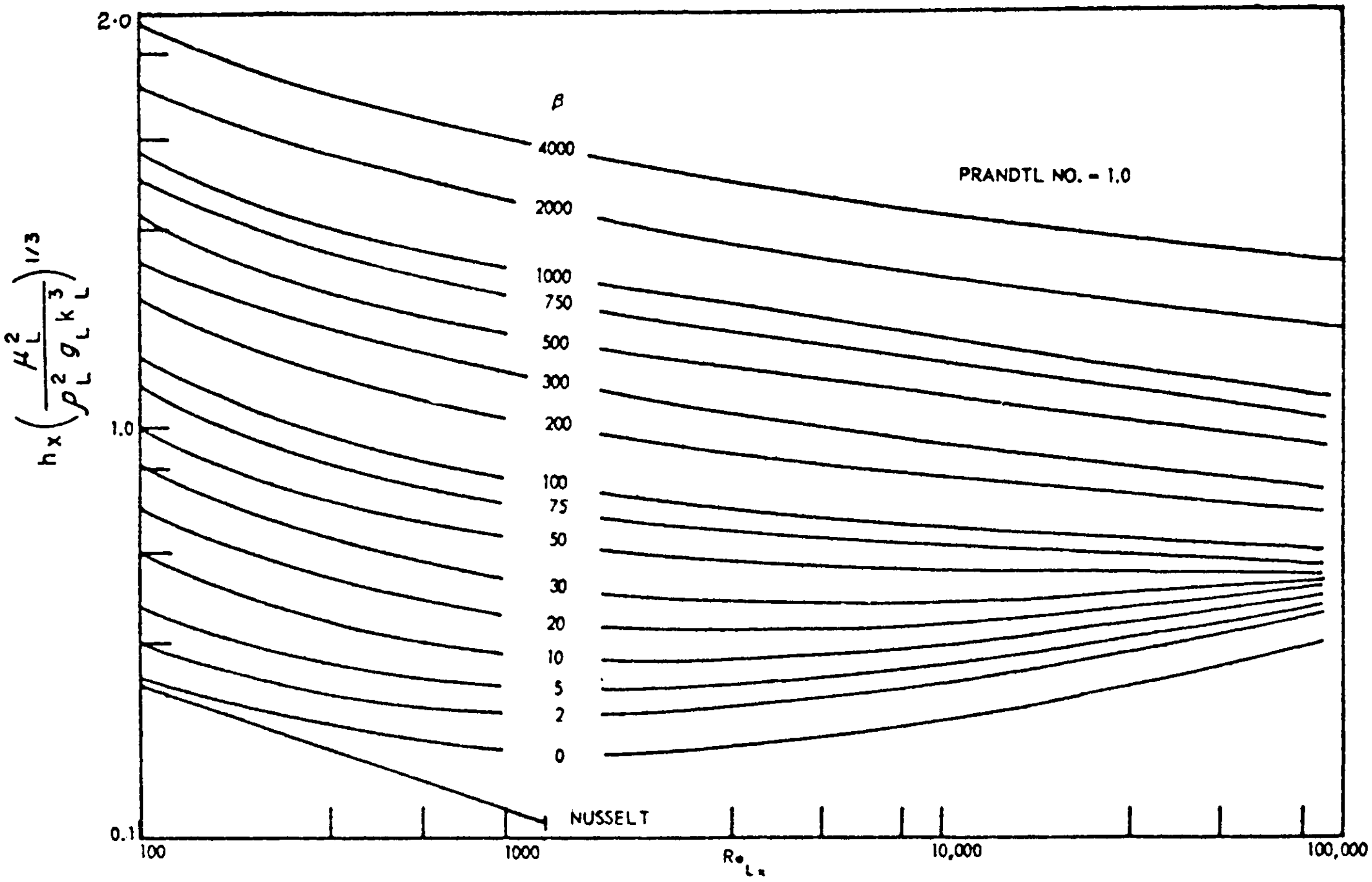


Figure 4.16. Local Coefficients: Prandtl Number = 1.0

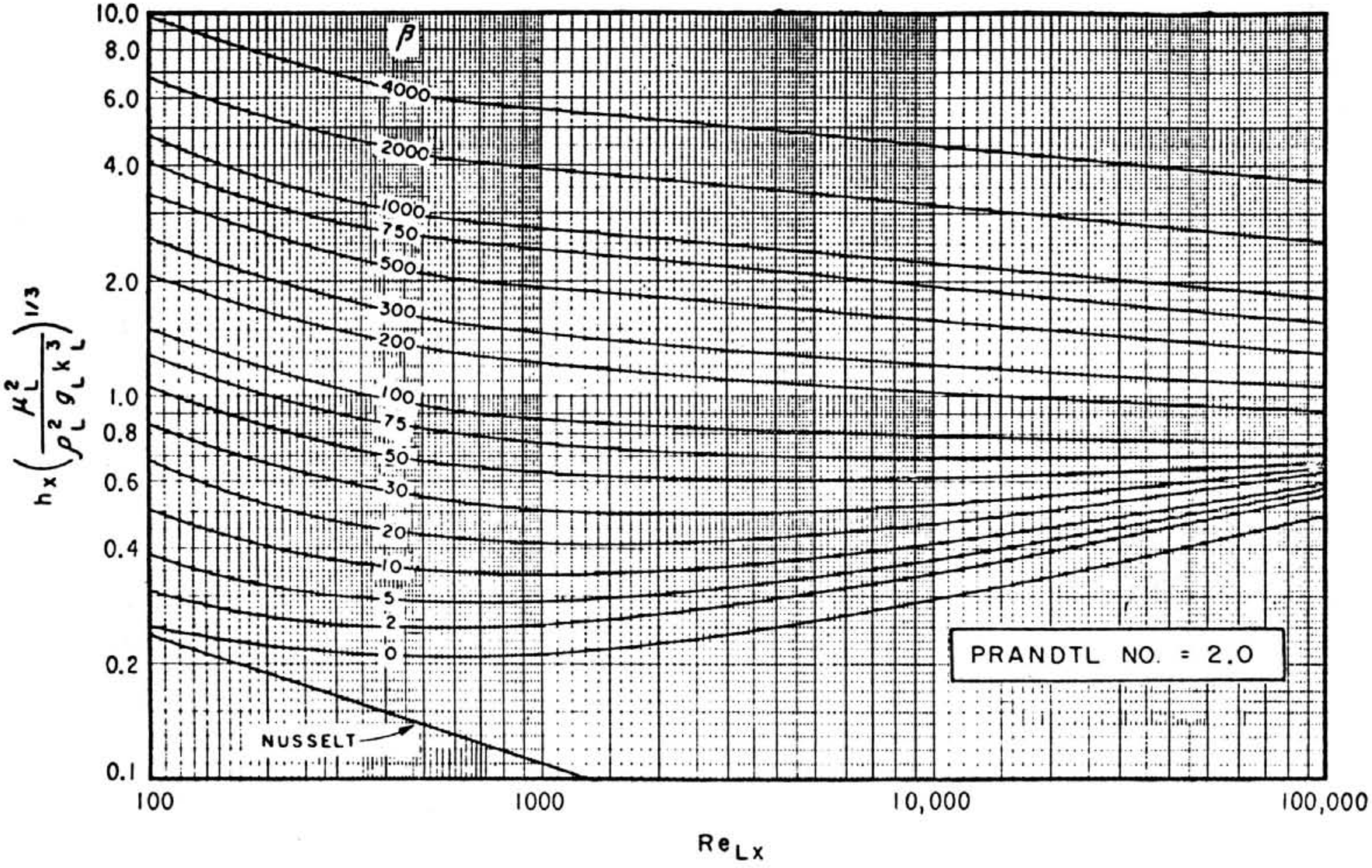


Figure 4.17. Local Coefficients: Prandtl Number = 2.0

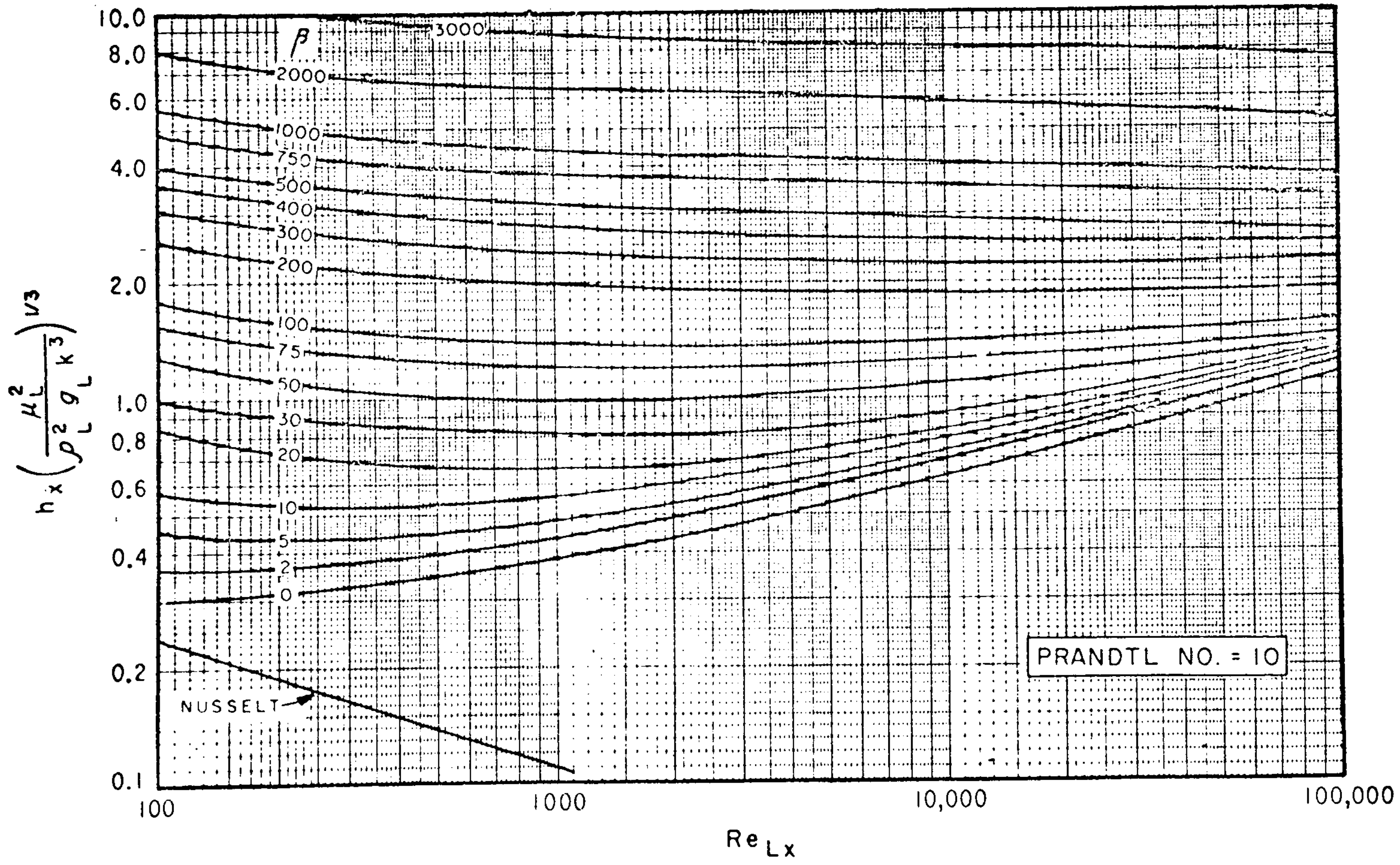


Figure 4.18. Local Coefficients: Prandtl Number = 10.0

Table 4.7

Overall Evaporative Coefficients at Low Production Rates

HEAT AND MATERIAL BALANCE NO.	14-6		14-7		14-8		14-9	
DATE	12/9/67		12/21/67		12/27/67		12/29/67	
NOMINAL RATE, PERCENT OF DESIGN PRODUCTION	80		60		40		25	
HEAT EXCHANGER	EVAPORATIVE COEFFICIENT BTU/HR/FT^2/ °F	STEAM TEMPERATURE INTO EFFECT °F	EVAPORATIVE COEFFICIENT BTU/HR/FT^2/ °F	STEAM TEMPERATURE INTO EFFECT °F	EVAPORATIVE COEFFICIENT BTU/HR/FT^2/ °F	STEAM TEMPERATURE INTO EFFECT °F	EVAPORATIVE COEFFICIENT BTU/HR/FT^2/ °F	STEAM TEMPERATURE INTO EFFECT °F
1	800	253.9	770	233.9	500	208.3	420	181.0
2	640	247.1	540	227.9	480	203.0	400	175.3
3	710	239.0	600	220.9	510	196.7	510	169.5
4	670	232.2	640	214.6	520	191.1	410	165.0
5	580	225.2	520	208.5	430	185.6	340	159.7
6	630	217.1	620	201.4	490	179.2	470	154.0
7	540	209.9	490	195.1	420	173.4	410	149.2
8	600	201.7	540	187.7	390	166.9	370	144.2
9	450	193.3	440	180.3	470	159.9	370	138.5
10	410	183.3	370	171.9	270	153.8	270	132.9
11	440	172.6	600	162.3	390	144.7	300	126.1
12	360	162.1	170	155.7	140	137.8	100	119.6
13	520	150.6	450	138.2	360	121.4	510	105.6
14	360	139.1	330	130.0	260	114.6	260	101.2
15	330	127.0	290	119.4	230	105.8	150	94.2
16	340	114.1	320	107.6	280	95.8	370	83.0
17	420	99.3	370	95.1	330	86.2	---	76.7
18	---	87.2	---	84.4	---	78.0	---	73.7
TOTAL ΔT °F		166.7		149.5		130.3		107.3

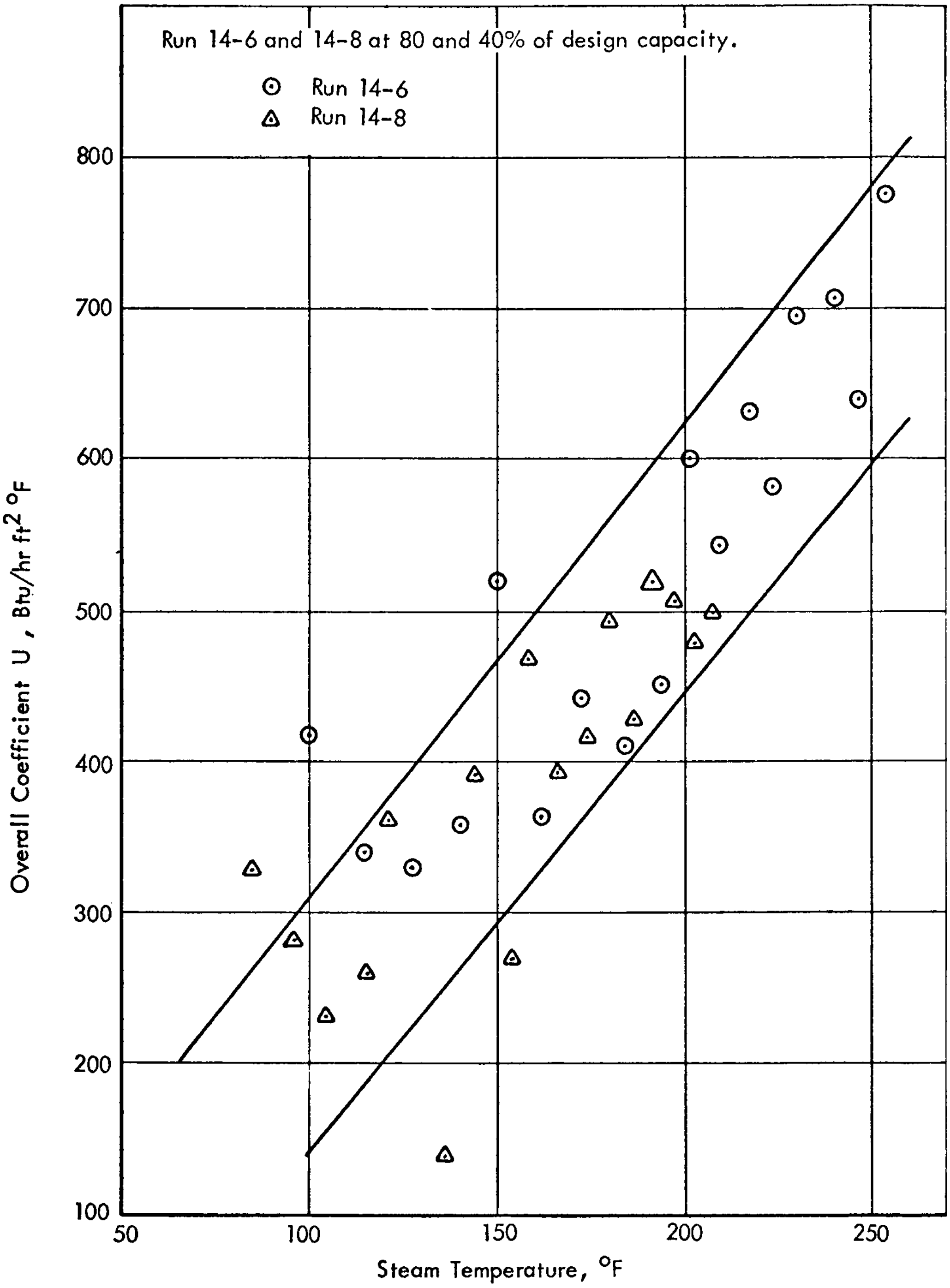


Figure 4.19. U at Reduced Capacities

b. Vapor Flow Rate versus U — The overall coefficient U depends on steamside coefficient h_b, as is evident from Equation 4.1. This coefficient depends on the interfacial shear β which in turn depends on the vapor velocity. Vapor flow cross-sectional area and volume of vapor flowing determine the vapor velocity. Since vapor seeks the path of least resistance, proper arrangement of tubes and vent locations is essential if high values of h_b (and hence of U) are to be secured.

Prengle, Dukler and Crump, Inc., showed (OSW R&D Report Number 74) that the steamside resistance is usually of the same order as the salineside resistance. Under conditions where, through high interfacial shear and brine recirculation, the saline side coefficient is increased significantly, the steamside coefficient becomes controlling. It then becomes necessary to look for means to raise this coefficient. In addition to methods which use grating type surfaces, some use should be made of interfacial shear β to improve this coefficient in order to move closer to the operational limit for this system. Tube spacings must be decreased and open area blanked with dummy tubes in order to increase vapor velocity on the steam side. As with vapor inside the tubes, the pressure drop resulting from high vapor velocity will decrease the driving force ΔT. However, operation with zero interfacial shear does not result in the optimum heat transfer fluxes ($^q/A = U\Delta T$).

Individual coefficients are hard to determine and this is especially true in plant size operation such as at Freeport. Effect of vapor velocity on the coefficient has to be indirectly inferred. For low production runs (see subsection D. 2. a above) the amount of vapor generated is less and so is its volume, so that its velocity through the same original tube and shell geometry is reduced. As mentioned earlier (Table 4. 7) the observed coefficients are lower partly due to this reason. Decline of performance (heat transfer) in Run 10A-5 at 60 percent design capacity is attributed (page 118, paragraph 6-119, Fifth Annual Report by Stearns-Roger) to the poor steam distribution in this run. Many tubes were removed from the tube bundle when it was noticed that there was excess surface in some effects. Steam distribution was also not satisfactory in the five-effect module.

c. Recirculation Rate of Sump Brine versus U — Brine recirculation from the sump of a given effect to the top water box of the same effect was initiated early in Freeport Plant life with the idea of securing better brine distribution to all the heating element tubes to reduce or eliminate scaling arising from insufficient feed. The following excerpt from the Second Annual Report (OSW R&D Report 100, page II-20) is pertinent:

> "It has been felt that the tendency of effect XII to become scaled is due to either one or a combination of the following: High concentration factor (in excess of 3.5); high vapor velocity through the tubes (this being a function of steaming rate and specific volume of the vapor); poor distribution of brine flow in the tubes; and <u>insufficient total brine flow for the number of tubes.</u> (Underscore added.)
>
> "The concensus was that the latter two could be eliminated by the installation of a brine recirculation system which would result in a flow of approximately 100 gallons per minute through the effect XII tubes. The distribution would be improved due to a positive hydraulic head on the distributor plate. In addition, it was felt that vapor velocities through the tubes would be decreased slightly if the flashing from effect XI brine would occur in the effect XII vapor body rather than in the water box where it normally occurs."

When recirculation was introduced in Effect XII there was still scale buildup (at C.F. = 3.5) and it "was not dramatically less than under similar conditions without recirculation." At a final concentration factor of 3.0, it has been reported that with the brine recirculation in operation it was possible to remove scale from Effect XII gradually. However, it is not certain whether the scale would not have dissolved without recirculation. It is better to improve liquid distribution by using efficient liquid distribution devices (see subsection C above) than to rely on recirculation. However, if it is known that maldistribution and scaling exist, then recirculation may be necessary to reduce the scaling although it increases pumping costs. Recirculation also increases the heat transfer area requirements (or, heat flux is reduced) as shown by computer calculations reported in OSW R&D Report Number 139, page 259. A comment relevant to this topic appears on pages 263-264 of this report:

> "Where an effect operates near saturation in the sump liquid, recirculation will help prevent local scaling (in the starved tubes) since it will tend to minimize local maldistributions.

> In addition, the larger flow rate at the same heat transfer rate results in a smaller concentration change across the tube, as shown above. However, this can be attained only at the price of increased area." (Underscore added.)

Dukler's Theory (OSW R&D Report Numbers 74 and 139) predicts only slight dependence of the heat transfer coefficient on the flow rate (through Reynolds Number) so that increase of brine flow rate per tube by recycling (or recirculation) will not significantly improve the coefficient except if it helps reduce maldistribution and thus scaling. The fact that recycling does not always reduce scaling or improve heat transfer rates is shown by the results obtained in Development Runs 6 and 7a, 7b, and 7c taken from Development Report Number 5 by Stearns-Roger (page 15, Table 4; and pages 23, 24, and 25, Tables 5, 6, and 7). This information is given in Table 4.8.

Table 4.8

First Effect Operation Conditions

High Temperature Tests - Runs 6 and 7

Operating Condition ↓ / Run No. →	6	7a	7b	7c
Average Flow Rate per Tube, Lb/HR	1294	1570	1558	1558
Average Steam Condensing Temp. °F	270.2	280	290	285
Average Seawater Feed Temp. °F	260	269.5	276.8	272.2
Overall Coefficient, U, BTU/HR $Ft^2 \cdot {}^oF$	757	Not reported	664	626

In spite of high steamside (and brineside) temperatures, the increased flow rate in Series 7 runs, the coefficient not only does not increase but actually decreases!

E. TYPES OF SURFACE USED AND PERFORMANCE

Three characteristics of the surface can affect the rates of heat transfer through the surface:

- thermal conductivity of the surface material;
- hydrodynamic roughness of the surface; and
- normal smooth surface versus enhanced surface such as double-fluted or spirally indented surface.

The effect of enhanced surfaces on U will be considered separately in the next subsection (F). The thermal conductivity of various metals used for heat transfer surface (tubes) in evaporators is listed in Table 4.9.

Table 4.9

Thermal Conductivity of Various Metals Used for Evaporator Tubes at Freeport

Metal	Thermal Conductivity Btu/hr ft °F
Stainless steel 316L	9.4
Titanium -5A	9.8
Low Carbon Steel	26
90/10 cupro-nickel	26
Ampco-grade 8	43
Aluminum brass	58
Admiralty	64
C.D.A. 194 alloy	151
Copper (phosphorized)	200

The hydrodynamic roughness of these surfaces affects the overall coefficient U to the extent that turbulence is induced in the film (condensate water or brine) by this roughness. Generally, an early (at low Reynolds Numbers) onset of turbulence is brought about by the tube surface roughness. However, roughness is undesirable because where mineral scaling is encountered it appears that scaling would be induced and retained more on a rough than on a smooth surface. In the Auxiliary Testing Unit at Freeport, two types of tubes were studied — aluminum brass tubes and 316L stainless steel. After approximately 500 hours run at 273°F aluminum brass tubes were found to be heavily fouled whereas the stainless steel tubes showed only light scale (Eighth Annual Report by Stearns-Roger, pages 37-38). What is more surprising is the fact that even with heavy fouling, heat transfer rates (U's) in ATU Run 15-2 were higher with fouled aluminum brass tubes than in slightly fouled stainless steel tubes (Development Report Number 9, page 120)! Lower thermal conductivity of stainless steel cannot account for all of the observed difference. The exact role of the tube alloy type and its hydrodynamic smoothness in the reduced scaling observed with the stainless steel tubes is not known. This phenomena deserves further investigation since Effects I and XIV, XV, XVI, which handle potentially scaling brines under optimum conditions, could be freed from scaling problems by installing such smooth (i.e., non-adherent scale) tubes in them.

F. VARIATION OF U WITH TUBE TYPE

For the major part of the developmental operation at Freeport (1961-1969) smooth surface tubes of the type normally utilized in process industries were incorporated in the evaporators and other heat exchangers. There are three main characteristics of tubes which can influence the rate of heat transfer:

1. Tube material (metals and alloys).
2. Tube dimensions (wall thickness, tube internal diameter and tube length).
3. Internal and external tube surface characteristics (smooth, double-fluted, spirally indented, etc.).

1. Tube Material

The influence of tube material on U is obvious — the higher the thermal conductivity of the material the higher U will be, in accordance with Equation 4.1. Thermal conductivities of some metals and alloys are summarized in Table 4.9. Since the tube wall resistance is given by x/k per unit of heat transfer area, the smaller the thickness of the wall, the lower will be the thermal resistance of the tube wall. Of course, the mechanical strength of the tube must be maintained and allowance made for corrosion and erosion of the tube surface. Table 4.2 contains information relating to tube materials, gage, outside diameter of tube, and tube length for the Development Run 16-3. From the tube resistance figures given there, it is seen that with low conducting materials, the resistance can be as high as 12.3% of the overall heat transfer resistance based on U.

2. Tube Dimensions

The effect of tube I.D. and length on U was discussed in subsection D.1. According to Dukler's Theory (OSW R&D Report Number 74) the tube diameter and length should be optimized for each effect; or, since this may be impractical, at least two or three sizes should be utilized. According to this theory, smaller tube sizes and longer lengths would give higher heat fluxes (${}^{q}/A = U \quad T$) in high-temperature/high-pressure effects; bigger tube diameters and shorter lengths would be used for low-pressure effects handling large volumes of vapor.

The principal objective of Development Runs Number 11 and 12 was to develop data on which a comparison of the performance of three-inch-diameter, 20-foot-long tubing could be made with two-inch-diameter, 24-foot-long tubing used in Development Run Number 10. The pressure drop, overall coefficient, etc., for the 2-inch and 3-inch-tube runs are shown in Table 4.10 (from OSW R&D Report Number 440, page 14).

It can be seen from Table 4.10 (last row for Effect XII) the pressure drop is lower in 3-inch tubes and overall coefficients are generally higher than for 2-inch tubes. Figure 4.20 shows that higher heat fluxes (which is the most appropriate quantity to compare) are obtained with 3-inch tubes only at low temperature differences ($<13^{o}$F) which is, in fact, the thermodynamically favored condition. Effects XIII, XVI, and XVII utilized tubes of 3 1/8, 2 1/2, and 3 inch diameter when the 12-effect Freeport Plant was modified to a 17-effect plant in July 1967.

Table 4.10

Comparative Heat Transfer Data

DATA SET NO.	10A-6	11-1	11A-1	11A-6	11A-9	11A-11	12-1	12-3	12A-2
DATE	3/23/66	5/25/66	7/1/66	7/14/66	7/22/65	12/16/66	11/3/65	11/23/66	1/13/67
PLANT SWF RATE M LB/HR	228.0	460.0	450.0	190.0	380.0	392.0	482.0	448.0	502.0
TOTAL PRODUCTION M LB/HR	186.3	340.3	347.7	141.2	280.4	285.6	347.0	338.4	342.0
PERCENT OF DESIGN (365,000)	---	93.2	95.3	38.7	76.8	78 2	95	87	93
HEAT REJECTION TEMP °F	105.4	118.8	115.6	100.3	105.0	109.9	111.1	110.8	105.7
EFFECT 1									
DUTY 10^6 BTU/HR	16.20	32.362	29.758	10.578	23.815	25.278	33.157	32.000	34.900
MTD/ΔTv	8.3/8.0	11.4/11.2	10.0/9.9	4.7/5.0	8.1/8.3	7.3/7.0	8.7/8.8	9.7/8.7	10.7/9.1
U_E BTU/HR-FT^2-°F	431	604	633	479	626	737	822	712	703
U_A BTU/HR-FT^2-°F	447	616	639	450	611	768	812	794	828
SURFACE AREA FT^2	4524	4700	4700	4700	4700	4700	4641	4641	4641
EVAP BRINE TEMP °F	212.9	258.2	256.2	188.9	235.9	244.6	263.4	261.0	261.0
EFFECT 10									
DUTY 10^6 BTU/HR	12.60	21.209	22.308			20.678	20.240	22.100	24.050
MTD/ΔTv	7.5/10.6	11.0/14.7	12.8/15.3			12.0/14.6	12.3/15.0	13.0/16.1	12.1/15.8
U_E BTU/HR-FT^2-°F	396	456	412			408	390	402	470
U_A BTU/HR-FT^2-°F	280	531	344			335	320	324	360
SURFACE AREA FT^2	4230	4700	4230			4230	4230	4230	4230
EVAP BRINE TEMP °F	135.3	156.2	150.0			144.0	154.6	167.3	154.6
EFFECT 11									
DUTY 10^6 BTU/HR	11.80	17.498	20.922.			20.176	19.783	17.800	17.800
MTD/ΔTv	9.95/13.0	10.98/11.1	12.7/14.4			13.4/15.0	15.8/19.2	11.9/16.2	14.9/18.1
U_E BTU/HR-FT^2-°F	243.2	323	334			305	254	474	243
U_A BTU/HR-FT^2-°F	188.0	422	294			272	209	347	200
SURFACE AREA FT^2	4935	4935	4935			4935	4935	4935	4935
EVAP BRINE TEMP °F	123.2	140.6	134.4			128.7	136.2	137.3	128.0
EFFECT 12									
DUTY 10^6 BTU/HR	9.60	37.580	20.180	8.291	17.473	19.583	19.905	9.250	21.700
MTD/ΔTv	8.1/13.4	11.65/15.8	12.9/16.3	7.4/9.9	12.7/16.6	12.8/16.1	18.9/21.9	18.4/22.6	14.3/19.4
U_E BTU/HR-FT^2-°F	287.2	347	364	260	320	356	246	304	351
U_A BTU/HR-FT^2-°F	173.5	256	288	194	245	283	213	248	259
SURFACE AREA FT^2	4113	4300	4302	4302	4302	4302	4302	4302	4302
EVAP BRINE TEMP °F	111.8	122.5	119.4	103.0	108.9	112.9	113.3	110.0	111.0
TUBE OD INCHES	2	3	3	3	3	3	3	3	3
MEAS ΔP IN TUBES PSI	0.130	.055	.079	045	445 ?	.045	.085	.090	040

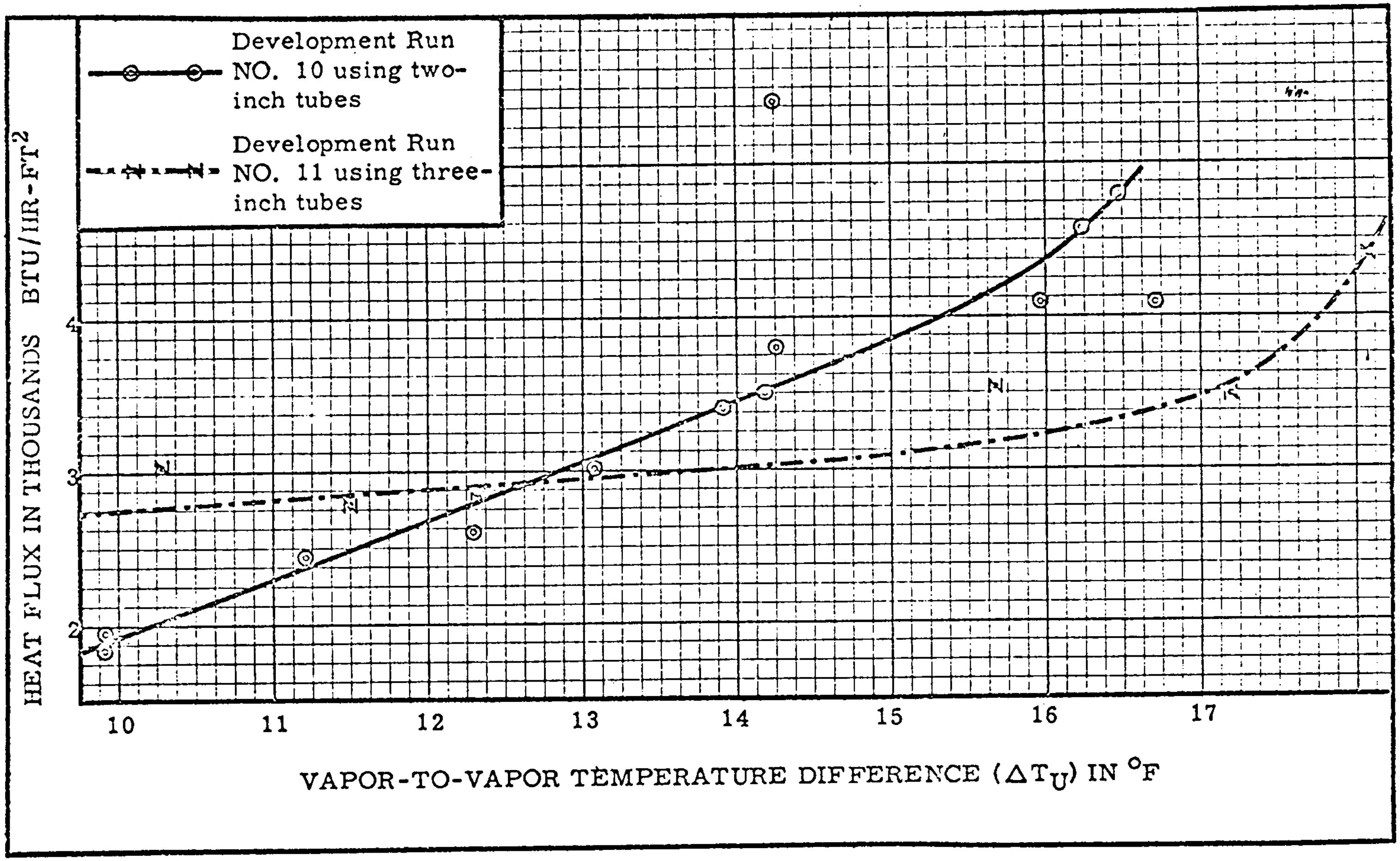


Figure 4.20. Effect XII Heat Flux versus Vapor-to-Vapor Temperature Difference, Development Runs Number 10 and 11

3. Double-fluted Tubes

Laboratory and pilot scale work at Oak Ridge National Laboratory had indicated very high heat transfer performance for double-fluted and spirally indented tubes. It was decided to test the long-term performance of these tubes on a large scale at the Freeport Plant. In February 1968 a test bundle of double-fluted tubes was installed in Effect XII of this plant. Partial view of a double-fluted tube is presented in Figure 4.21. The tubes are 11 feet 1 inch long, having 3 1/8 inch diameter smooth ends where the wall thickness is 0.065 inch. There are 84 flutes around the outside; the average flute depth is 0.037 inch on the outside and inside. An enhancement factor of 1.3 was applied to the fluted portion of the tubes.

Overall heat transfer coefficients and other related data taken during Development Run Number 15 are shown plotted in Figure 4.22. The coefficient is reported to average about 1400 Btu hr/ft^2 $^{\circ}$F (OSW R&D Report Number 479, page 104). Table 4.11 compares the heat transfer coefficients for Run 14 (smooth tubes) with those for Run 15. Values obtained with double-fluted tubes were about 250% greater than those values obtained with smooth tubes. On page 50 of Development Report Number 9 (by Stearns-Roger) it is reported that an increase of 300 lb/hr in the venting rate, or a considerable cut in the tube feed rate did not have a significant effect on the heat transfer coefficient.

During Run 16 the tube feed rate was 1100 lb/hr on recycle; the evaporator feed mode was sump-to-sump (from Effect XII) while venting full from the chest at about 500 lb/hr with the dome vent open. Overall coefficient data for this run are reported in Table 4.12 (Development Report Number 9, page 49). The reason(s) for the decline of U from 1400 to 1180 between Runs 15 and 16 were not discussed in the report.

Experimental studies by Dukler, et al, on single-fluted tubes of 2 inches and 3 inches diameter (10 feet long) showed that the high performance of double-fluted tubes was due to the fact that the steamside coefficient was three to five times higher than the brineside coefficient (OSW R&D Report Number 487). Further improvements in fluted tube heat transfer can come from research directed at reducing the resistance of the evaporating brine film.

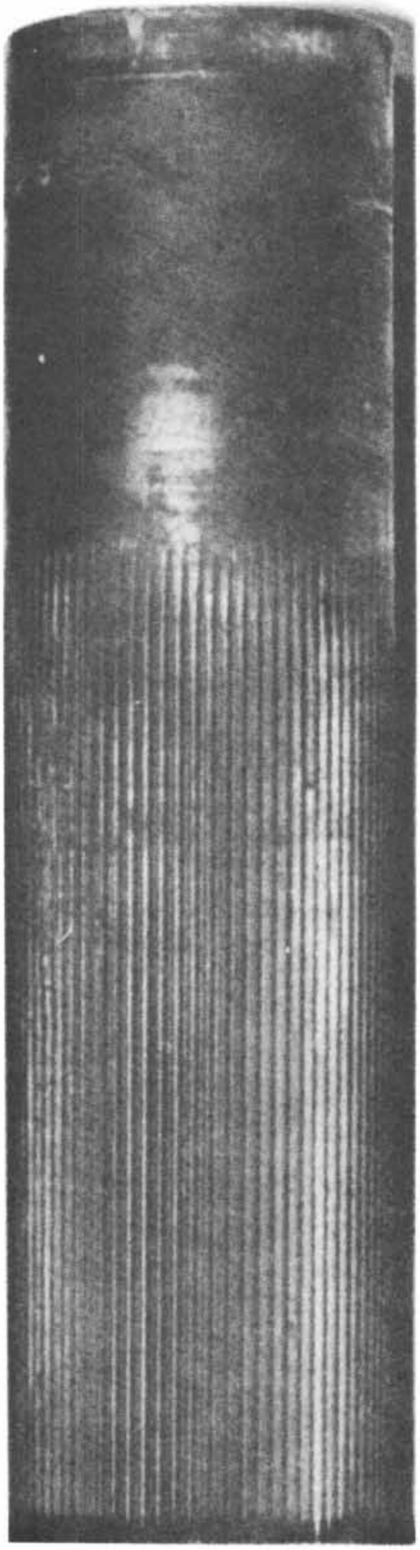


Figure 4.21. Three Inch Diameter Fluted Tube

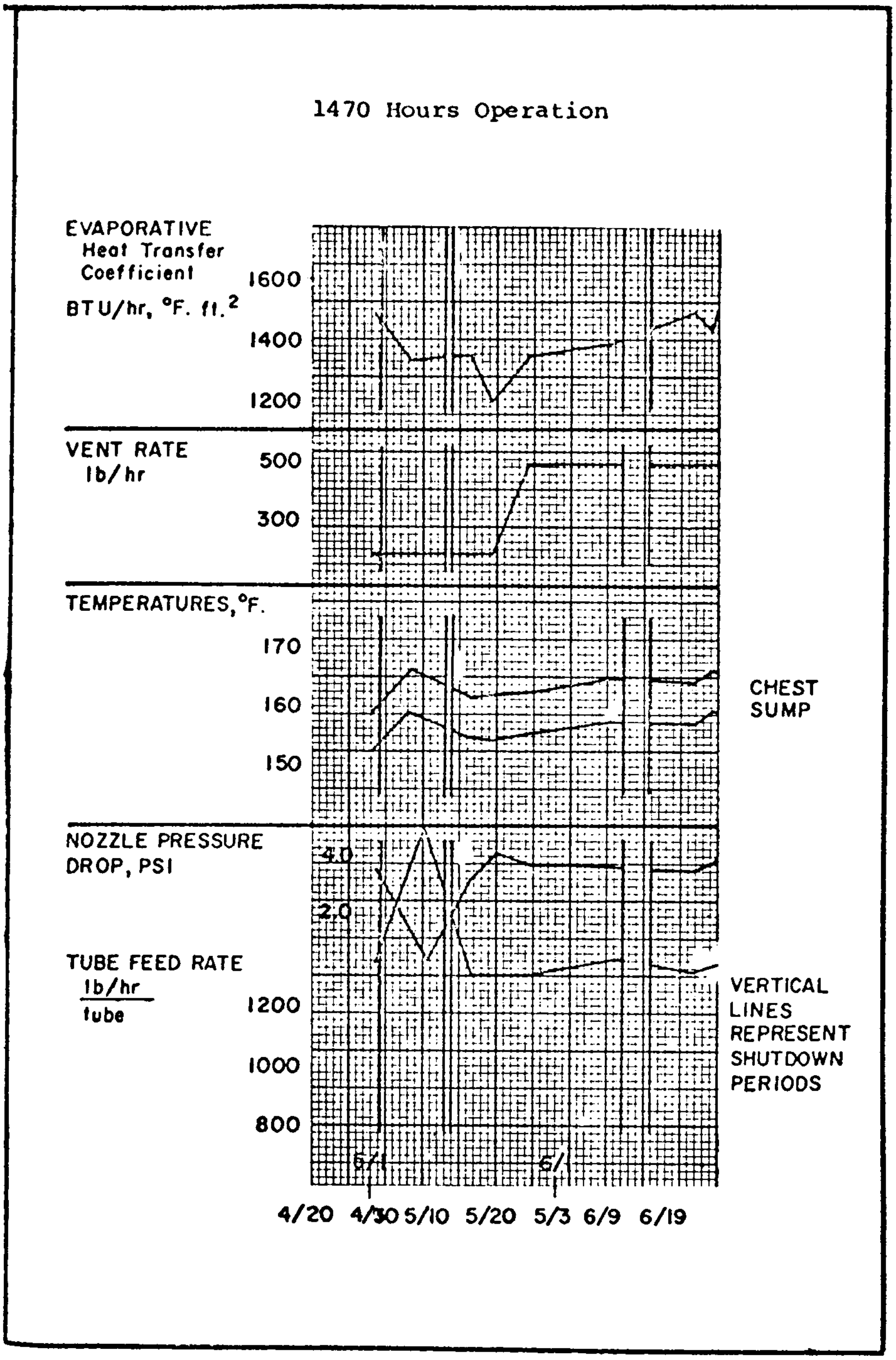


Figure 4.22. Double-fluted Tube Performance — Development Run Number 15

Table 4.11

Effect XIII Double-fluted Tube Performance

Date	Plant Feed M lb/hr	No. 13 Chest °F	Qt 10^6 Btu/hr	ΔT °F	Uo Btu/hr-ft^2-°F
Run 14 Average Smooth Tube:	485	158.5	16.2	7.8	500
Run 15:					
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
Run 15 Average:	470	165.0	16.4	6.9	1,380

Table 4.12
Effect XIII Double-fluted Tube Performance

Date	Plant Feed 10^3 lb/hr	No. 13 Chest °F	Qt 10^6 Btu/hr	ΔT °F	Uo Btu/hr-ft^2-°F
Run 16:					
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
Run 16 Average:	482	157.1	16.2	8.2	1,150

G. EFFECT OF NONCONDENSIBLES ACCUMULATION AND VENTING RATE ON U

Gases like air, carbon dioxide, and hydrogen (present in water vapor or steam used as a heating medium) do not condense at the normal temperature/pressure conditions prevailing in the evaporator steam chests. The gases tend to accumulate in the vicinity of the condensed water film, as depicted in Figure 4.23.

1. Effect of Noncondensibles on U from Theory

In the thermal circuit involved in transfer of heat from the steam to the evaporating brine side there are six thermal resistances acting in series:

- Bulk steam region — involves very small thermal resistance.
- The accumulated noncondensibles film region — involves not only thermal resistance but also mass transfer resistance since condensing steam has to diffuse through this film to reach the condensate film.
- Thin film resulting from condensation of steam — offers significant resistance to heat transfer.

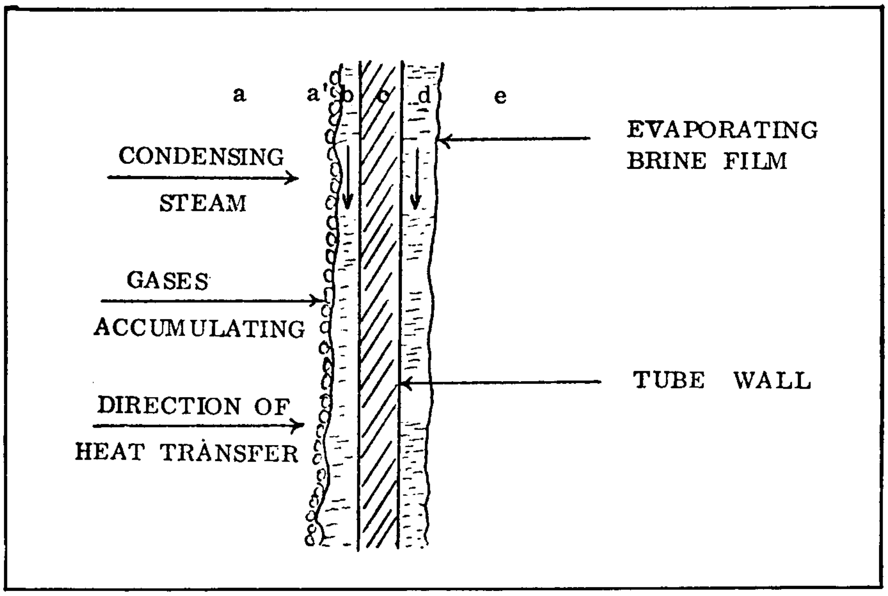


Figure 4.23. Noncondensibles Accumulation on Falling-Film Heat Transfer Surface

- The tube wall region — for low thermal conductivity alloys (or metals) the thermal resistance could be as high as 10 to 15% of the overall resistance.
- The evaporating brine film — according to Dukler's Theory (OSW R&D Report Number 74) this region offers about the same resistance as the steam condensate film, when using smooth tubes. With double-fluted tubes this is the controlling resistance.
- Bulk water vapor region — negligible resistance.

In the discussion above, the existence of a layer of scale on either the steam or the brine side has not been taken into consideration. Scaling problems will be discussed in a separate section.

Confining our attention to the steam side, we notice that the presence of a noncondensible film at the steam-liquid interface leads to the following effects:

- Thermal resistance increase due to the extremely low value of thermal conductivity of the gaseous film.
- Decrease in the rate of steam transfer to the steam-liquid (condensate water) interface.

- The condensing temperature of steam is reduced, thus reducing the temperature driving force.
- Sensible heat transfer between the bulk of steam and the condensate film affects the liquid-gas interface temperature.

Based on these considerations, Colburn and Hougen[1] proposed the following equation:

$$q' = U' (t_i - t_L) = h_G (t_b - t_i) + \lambda K_G (p_b - p_i)$$

where

q' = heat flux, Btu/hr ft^2
U' = overall coefficient based on $(t_i - t_L)$
h_G = local coefficient for transfer of sensible heat through gas film
K_G = mass transfer coefficient through gas film, lb of vapor condensed/hr ft^2 atm pressure difference
t_i = temperature of liquid-gas interface
t_b = bulk temperature of gas stream on shell side
t_L = bulk temperature of steam on tube side
p_b = bulk partial pressure of the shellside temperature, atm
p_i = vapor pressure of the liquid at temperature t_i, atm
λ = latent heat of condensation of steam, Btu/lb

In this equation the heat transferred away from the interface is equated to the sensible heat transferred from the vapor plus the latent heat given up at the interface by the diffusing vapor molecules. The interface temperature t_i is unknown and has to be obtained by trial and error. Since the gas concentration and K_G vary along the length of the tube, graphical or numerical technique must be used to apply this equation to the determination of the total heat transfer area. Agreement with experimental values was reported to be good (OSW R&D Report Number 287, page 47). A complete sample calculation is also given in Report Number 287. Table 4.13 compares the actual heat transfer fluxes. [in column marked $\left(\frac{dq}{dA}\right)_n$] with those that would exist if no air were present with

---

(1) Colburn, A. P., and O. A. Hougen, Ind. Eng. Chem., 26, 1178-1182 (1934)

Table 4.13

Experimental Data for Determining Effect of Noncondensables on Heat Transfer

RUN IDENTIFICATION NUMBER	CONDITIONS			HEAT FLUX BTU/(HR)(SQ FT)		HEAT TRANSFER COEFFICIENTS BTU/(HR)(SQ FT)(°F		EXP. EFF
	Tubeside Vacuum (in Hg)	Shellside Vacuum (in Hg)	Air Conc. (Mole %)	$(\frac{dq}{dA})_n$	$(\frac{dq}{dA})_s$	$h^{(1)}$	$h^{(2)}$	η %
IVA-155-21.8-600-1.2	25.0	21.73	1.0	4860	5770	1430	1700	84
IVA-162-23.2-600-4.0	25.0	23.77	3.7	1870	5040	645	1740	37
IVA-163-21.8-600-1.5	25.0	21.79	1.5	5040	6080	1400	1690	83
IVA-164-19.3-600-1.0	25.0	19.22	1.0	7800	7990	1620	1660	98
IVA-165-16.2-600-0.7	25.0	16.19	0.7	10700	11100	1570	1630	96
IVA-166-23.7-600-5.0	25.0	23.55	2.8	2520	3390	1400	1880	74
IVA-167-21.8-600-1.5	25.0	21.84	0.6	4440	5210	1480	1740	85
IVA-168-21.8-600-3.0	25.0	21.74	1.8	4460	6550	1140	1680	68
IVA-170-21.8-600-5.0	25.0	21.92	3.3	3960	7030	944	1670	56
IVA-171-21.8-600-10	25.0	21.80	6.6	3630	10150	586	1640	35
IVA-173-19.3-600-5.0	25.0	19.25	3.3	7060	10400	1100	1630	68
IVA-175-16.2-600-5.0	25.0	16.06	3.2	10500	13900	1220	1620	76
IVA-179-23.7-600-5.0	25.0	23.50	2.4	2230	3900	1010	1770	57
IVA-181-21.8-600-1.0	25.0	21.68	0.6	4960	5470	1550	1710	91
IVA-183-21.8-600-3.0	25.0	21.72	1.8	4680	6080	1300	1690	77
IVC-177-15.6-600-5.0	21.0	15.42	3.2	4970	8110	1020	1660	61
IVC-185-15.6-600-10	21.0	15.54	5.6	4220	10000	693	1630	42

---

(1) Apparent coefficient with air present as defined by Equation 3.3.

(2) Coefficient with pure saturated steam as defined by Equation 3.1 or 3.2.

steam $\left[\text{in column marked} \left(\frac{dq}{dA}\right)_s\right]$. The last column in this table gives the so-called experimental efficiency η which is the ratio $\left(\frac{dq}{dA}\right)_n \Big/ \left(\frac{dq}{dA}\right)_s$ times 100. A significant decrease in heat flux is seen with only a few percent concentration of air in steam. Figure 4.24 depicts graphically the effect of air concentration on heat flux efficiency. A trend for lower efficiencies with higher vacuums at a given air concentration was noticed.

These results suggest that for obtaining high heat fluxes in the presence of noncondensibles in the heating steam, proper arrangements (tube spacing, location of vents, etc.) must be made to remove the noncondensibles at a rate sufficient to reduce their impact on heat transfer.

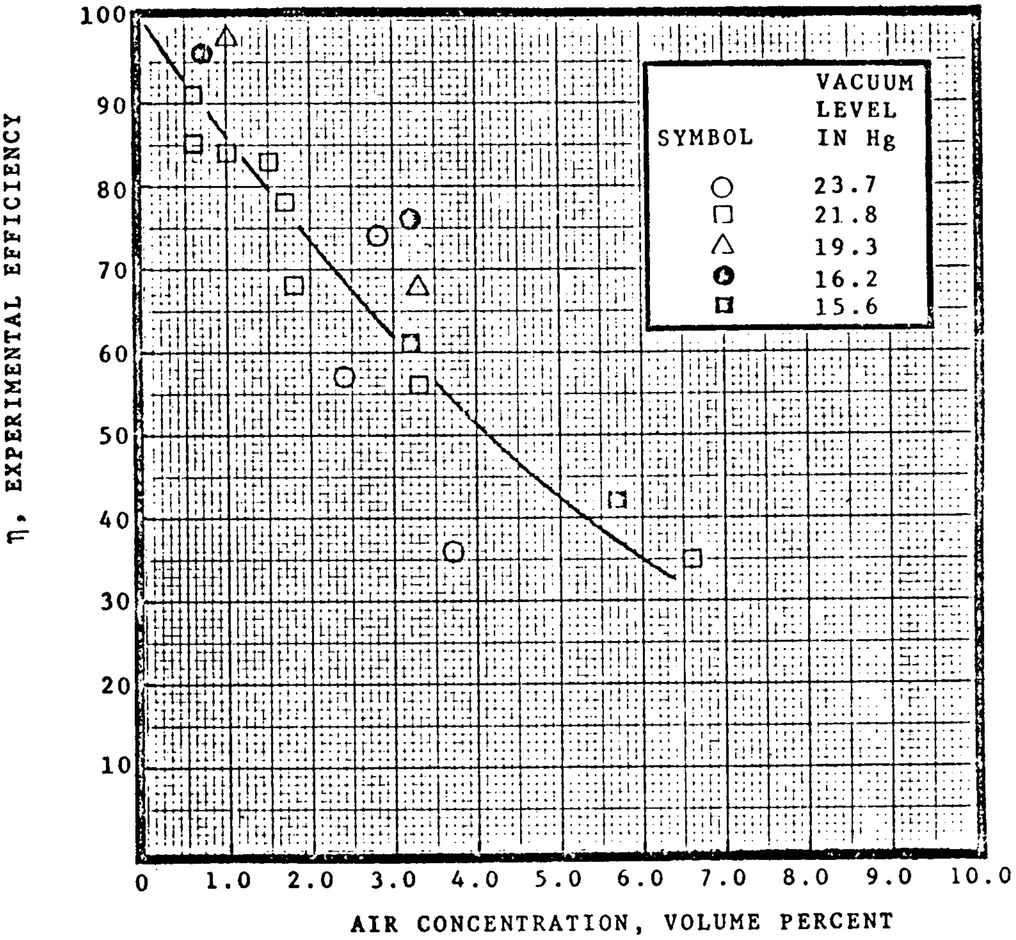


Figure 4.24. Effect of Air Concentration on Experimental Efficiency

2. Observations at Freeport Plant

Noncondensible gases in the evaporator steam chest can come from:

- Dissolved gases in feed water.
- Decomposition of bicarbonates, carbonates, etc., due to high temperature and/or presence of acid.
- Leakage of air into effects which operate at pressure below atmospheric.
- Small amounts of gases originating from electrode reactions (corrosion-type), e.g., hydrogen.

The best remedy in any situation is to eliminate the trouble at its source. Rather than trying to solve the venting problem, it would be best to prevent air or other noncondensibles from getting into the system. Dissolved gases in seawater feed are removed at Freeport by the deaerator-decarbonator, where sulfuric acid was added to chemically decompose the bicarbonates and carbonates. The dissolved oxygen level in the effluent seawater (from this unit) was reported (Eighth Annual Report, pages 29-30, by Stearns-Roger) as lying between 30 and 50 parts per billion, and the carbon dioxide content ranged from 0.3 ppm to as high as 9.9 ppm. Close control of DA-DC is, therefore, essential to reduce venting and corrosion (from O_2 and CO_2) problems.

Leakage of air into steam chests or at any point in the system is another problem that is best tackled at the source. All piping, vessels, joints, etc., must be airtight. Best quality gasket material must be used on all joints.

Corrosion reactions must be prevented both to reduce corrosion and to avoid generation of gases like hydrogen which can reduce the heat transfer rates. Carbon steel plugs were originally used at Freeport to plug the aluminum brass tubes in the various evaporator effects. The two dissimilar metals in contact, with seawater acting as an electrolyte, caused rapid corrosion of the plugs and generated significant amounts of hydrogen.

Vent locations in the 17-effect Freeport Plant are shown in Table 4.14.

A detailed study of the effect of vent location and vent rate on evaporator performance was conducted at Freeport in two phases during the fiscal year 1966. Phase I involved evaluation under the then standard conditions of

Table 4.14

Vent Handling

Development Run No: 15

Data Taken By:

Heat & Material Balance No: 4

Date:

EVAPORATOR	VALVE SET		FLOW	EXCHANGER	VALVE SET	FLOW
	TOP	BOT				
1	1/2	C	Cascade	302a	1/2	Cascade
2	3	C	Cascade	302b	1/2	Cascade
3	1	C	Cascade	303a	1/2	Cascade
4	Set By KSC	C	To Atmos.	303b	1/2	Cascade
5	2	C	Cascade	304	1/2	To Atmos.
6	1	C	Cascade	305	1/2	Cascade
7	1	C	Cascade	306a	1/2	Cascade
8	2	Wide Open	Cascade	306b	1/2	Cascade
9	2	C	Cascade	307	1	Cascade
10	Wide Open	Wide Open	Cascade	308	1	Cascade
11	Wide Open	Wide Open	Cascade	309	1	Cascade
12	Wide Open	C	Top Vent (New)	310	Wide Open	Cascade
MODULAR EFFECTS	VENT MODE		FLOW	311	Wide Open	Cascade
	TOP	VENT TUBE		312	New	Top Closed Bottom Open 3 Turns
13	Wide Open	-	Cascade	313	C	
14	Wide Open	-	Cascade	314	C	
15	Wide Open	-	Cascade	315	C	
16	Wide Open	-	Cascade	316	C	
17	Wide Open	-	Cascade	317	C	
				318	Wide Open	To B.C.

Effects II and III venting to the atmosphere via a common vent manifold. This evaluation led to revisions in the vent piping system and Phase II of the vent study. The revised vent piping used in Phase II is shown in Figure 4.25. The revised system enables the venting of Effects II, III, and IV to the atmosphere through individual lines such that any two of these three evaporators can be vented simultaneously. Each of Effects II, III, and IV were individually vented to the atmosphere, while all other effects were vented in series to their respective downstream effect. The effect of vent location and venting rate on the overall coefficient is shown in Figure 4.26 where the detrimental effect of vapor binding (noncondensibles accumulation) on U in Effect III is clearly seen (while venting Effect IV to atmosphere). The following observations were made in this study:

- The method of venting can significantly alter plant operation and production rate. Table 4.15 illustrates this.
- The operating temperature profile can be altered by changes in vent flow rate; however, the changes are small, as long as positive venting is maintained.
- The operating steam economy ratio can be altered by changes in vent rate or vent location, vent location changes being more significant than vent rate changes. The lower the temperature level of venting, the greater the steam economy.
- Insufficient venting reduces the operating heat transfer coefficient in an evaporator and can throttle plant production rate if the operating flashing range is predetermined and cannot be increased.

H. PREHEATER PERFORMANCE

The preheaters play an important role in the economic production of potable water from saline water in any evaporative distillation process such as the one utilized at the Freeport, Texas, desalination plant. Seawater temperatures in the Gulf of Texas average around 60 to 88°F. The first effect evaporating temperature is in the range 250 - 275°F; therefore, if the feed water were not preheated, there would be only sensible heating of this feed in the first effect heating element, and no product vapor! How this preheating is accomplished is also important since it has direct influence on the cost of water produced. W. L. Badger and Associates, the architect-engineer for the original 12-effect Freeport Plant, proposed using the condensate from each effect steam chest and part of the vapor

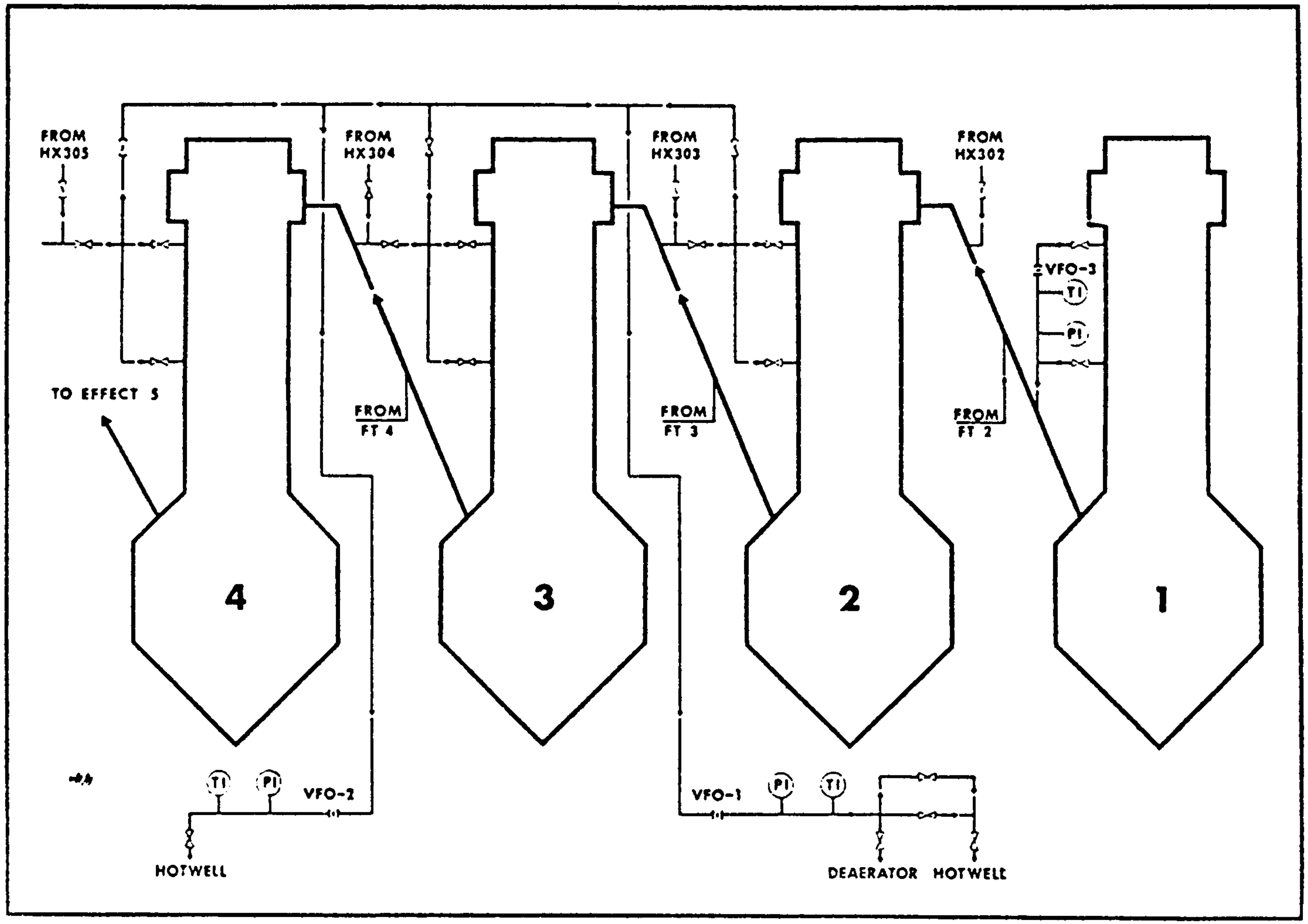


Figure 4.25. Effects I through IV Vent Piping

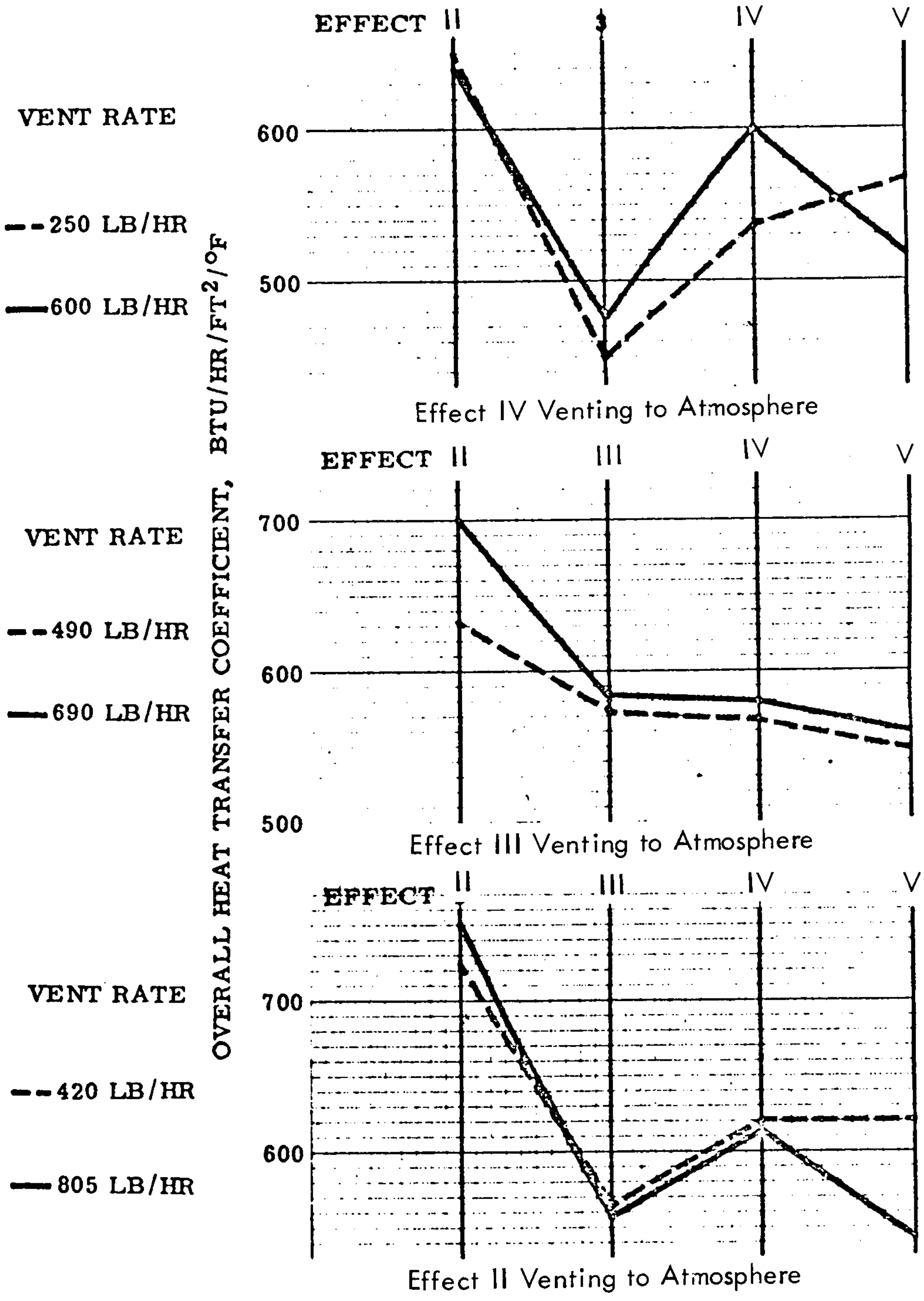


Figure 4.26. Evaporator Heat Transfer Coefficients

Table 4.15
Average Daily Production versus Various Venting Modes

ATM VENT LOCATION EFFECT	VENT RATE LB/HR	AVERAGE GROSS DAILY PRODUCTION GALLONS	REMARKS
2	805	1,043,700	
2	620	1,046,000	
2	420	1,051,000	
3	690	1,041,000	
3	575	1,043,100	
3	490	1,040,200	
4	600	1,026,400	Effect 3 vapor bound
4	250	988,300	Effect 3 vapor bound

from the vapor head of each effect as a heating medium in two types of heat exchangers called Series 200 (condensate coolers) and Series 300 (bleed vapor condensers) respectively. The original arrangement of these exchangers is shown in Figure 4.27. There were 14 units of the 200 Series and 12 of the 300 Series for preheating the feed. Figure 4.28 contains sketches of both Series 200 and 300 heat exchangers. Since heat transfer coefficients in liquid-to-liquid heat exchangers are very low (about 75 to 200 Btu/hr ft^2 ^{o}F) the heat transfer surface in such exchangers is not efficiently utilized. The operation and maintenance contractors for the Freeport Plant, Stearns-Roger Corporation, suggested to OSW that exchangers 202 through 207 be bypassed and instead flash-tanks 1 through 7 be utilized to recover heat from condensate stream. This was accepted, and no appreciable reduction in steam economy was noted (page IV-2, First Annual Report). This arrangement eliminated six condensate pumps and produced reductions in power consumption and maintenance costs.

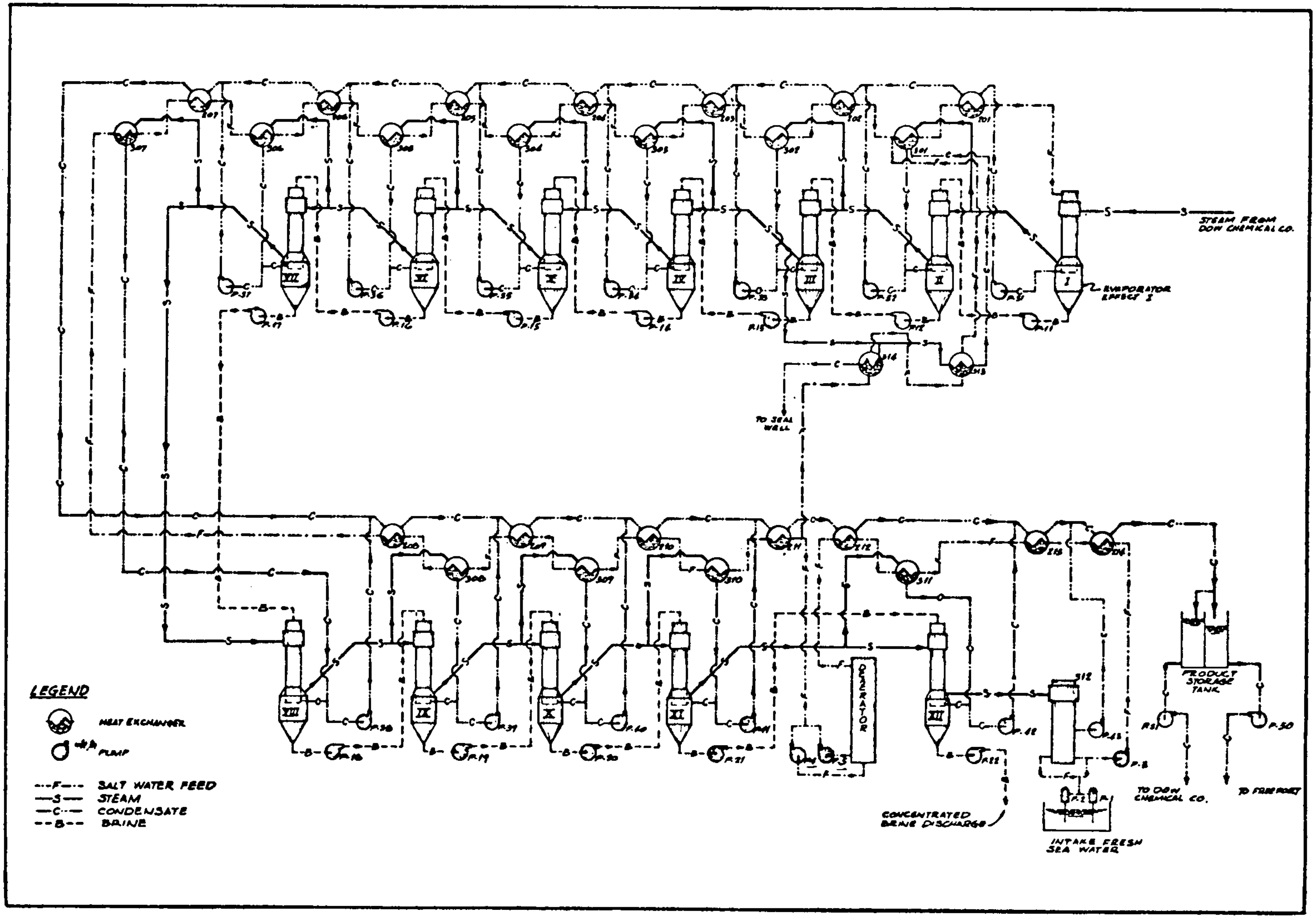


Figure 4.27. Freeport Saline Water Demonstration Plant Flow Diagram

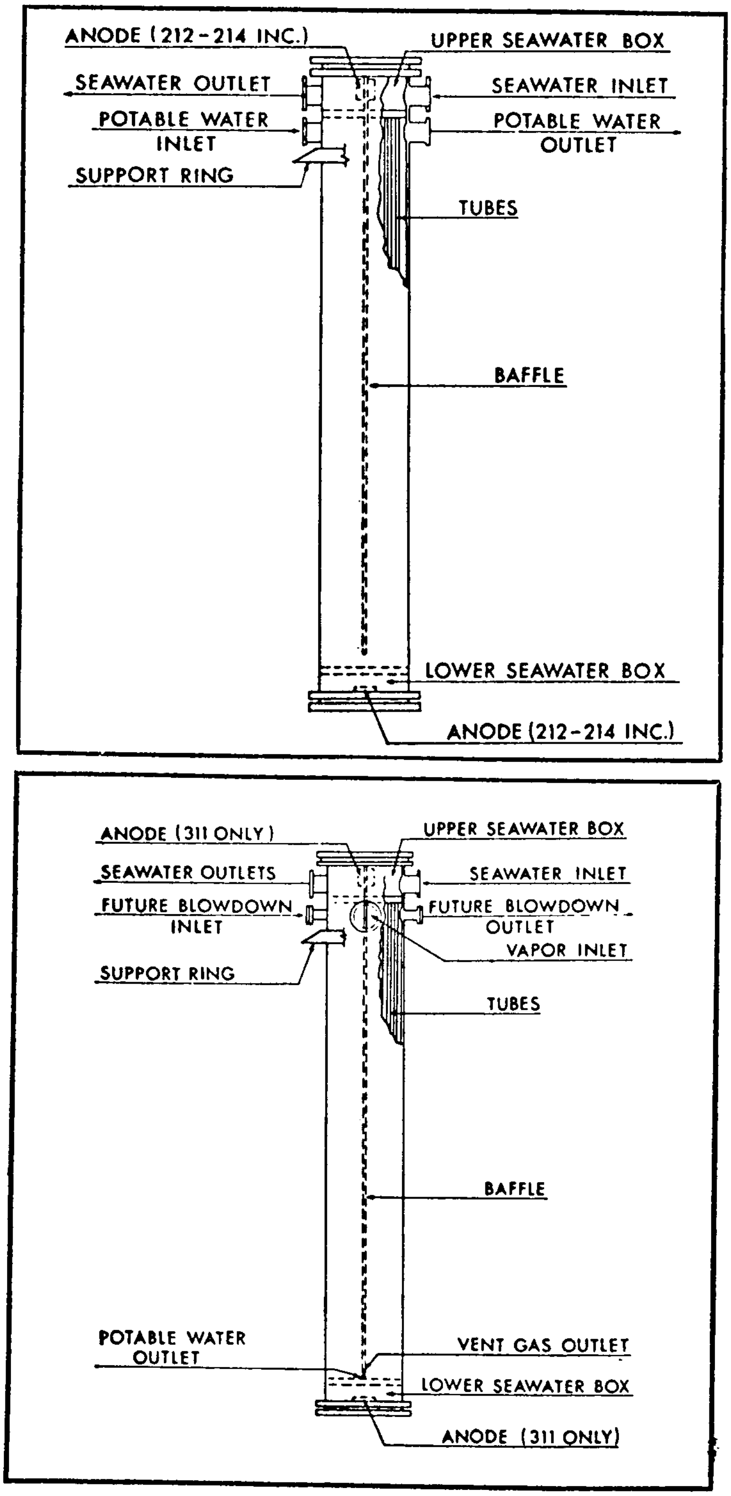


Figure 4.28. Series 200 (Upper) and 300 (Lower) Preheat Exchangers

This revised arrangement is shown in Figure 4.29 (from Annual Report Number 3, page 147). The temperatures of various streams, amounts of heat transferred, and heat transfer areas of each exchanger are shown for Development Run Number 7c in Figure 4.30 (Figure 5-8, Fourth Annual Report). The temperature profile for these exchangers is shown in Figure 4.31. Since small temperature driving forces lead to more efficient operation (energywise) as dictated by the Second Law of Thermodynamics[1], it was suggested (OSW R&D Report Number 171, page 108) that installation of additional effects in the low temperature end would reduce the temperature driving forces and hence improve the preheat circuit performance.

Overall heat transfer coefficients for all the preheat exchangers are given in Table 4.16. If the heat transferred in all the exchangers (except 312, which is a heat rejection condenser) is added together we get 91,306,800 Btu/hr, which is more than 270% of the heat supplied in the first evaporator by prime steam. Of this total, 56.9885 million Btu's are transferred in Series 300 exchangers having 17,287 ft^2 total heat transfer area giving a heat flux of 3296.6 Btu/hr ft^2, whereas Series 200 give a heat flux of only 2131.0 Btu/hr ft^2. This low flux and the low heat transfer coefficient (in Table 4.16) indicate that all of the 200 Series exchangers could be replaced with flash tank heat recovery mode without much loss in thermal economy. The resultant duty increase for exchangers 308 through 311 would actually be better from the performance standpoint as was noticed for units 301-307 when cross exchangers 202-207 were removed (OSW R&D Report Number 171, page 126).

1. All-flash Product Heat Recovery

It was decided to check the feasibility of all-flash product heat recovery by excluding preheaters 208 through 211 during the Development Run Number 8 operations. In this all-flash operation the saturated condensate of Effects II through XI was successively flashed to Pump P-44 which delivered the condensate to the product header at the inlet to HX-212 (heat rejection condenser).

The seawater feed approach to condensing-steam temperature in the preheaters (HX-300's) used to be low during normal flash operation (condensate flashing of Effects II through VII). With all-flash operation, closer approach was secured which was attributable to the increased duty imposed upon the preheater

---

(1) Kays, D. D., "Desalting Sea Water in the ME-LTV Evaporator", First International Symposium on Water Desalination, October 1965.

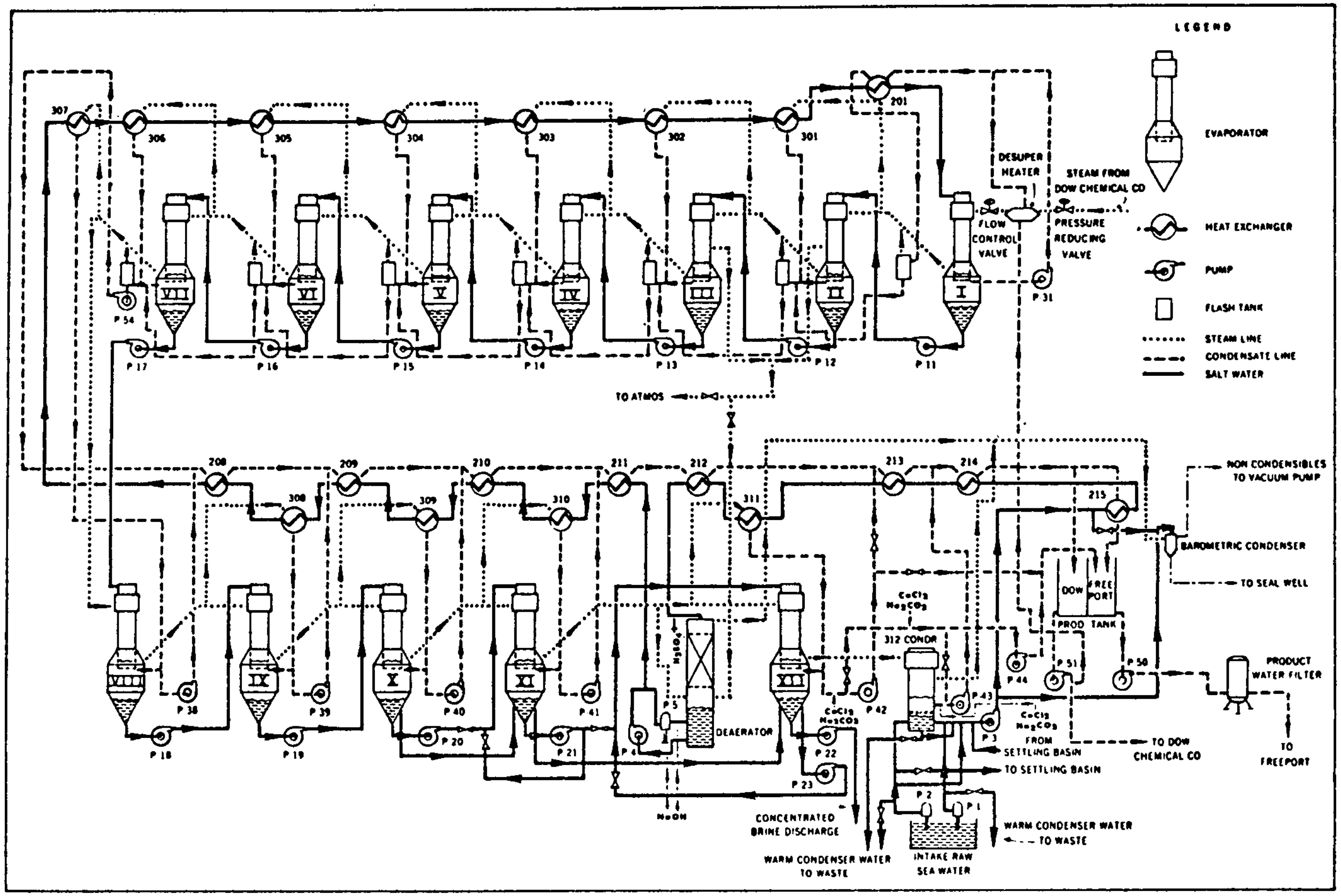


Figure 4.29. Process Flow Diagram

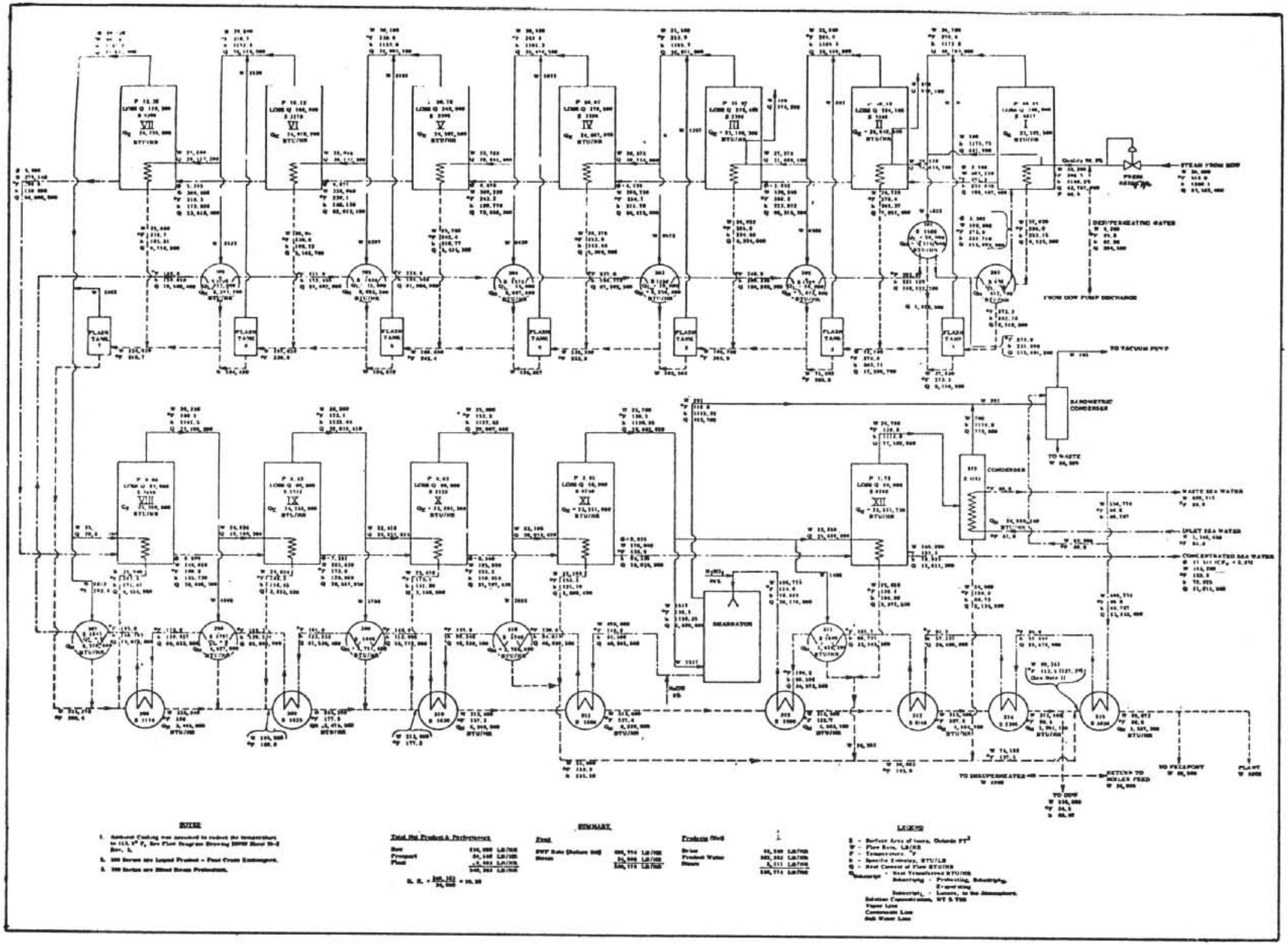


Figure 4.30. Heat and Material Balance Diagram, Development Run 7c

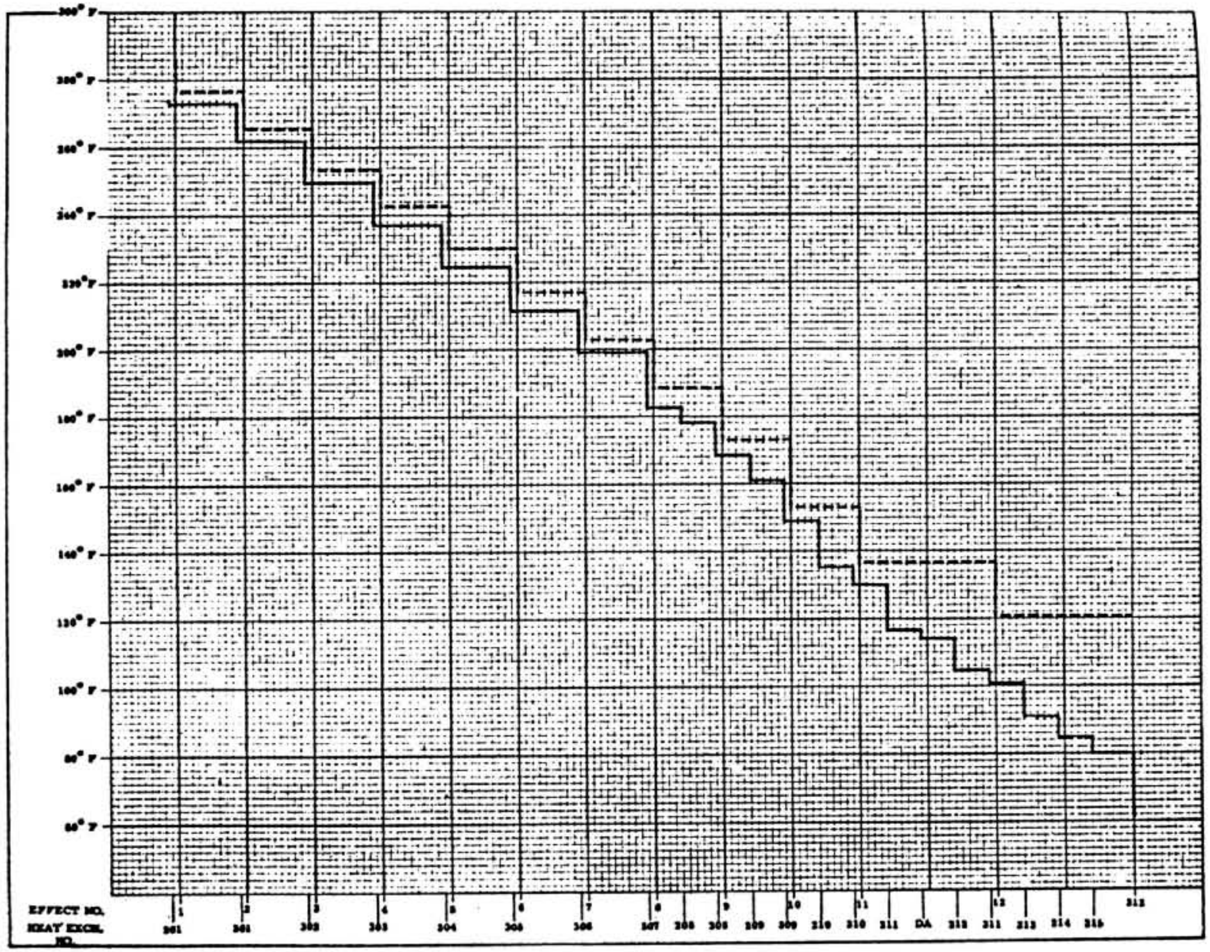


Figure 4.31. Temperature Profiles, Development Run 7c

Table 4.16

Individual Performance of Preheating Exchangers for Development Run 7c

Hx	Surface Area	· TEMP B_i	C_o	B_o	C_i	ΔT_i	ΔT_o	LMTD	Q_H	"U"	Δt_{swf}
301	1509.1	262.0	276.4	273.0	276.4	14.4	3.4	7.64	5,379,000	466.8	11.0
302	1388.6	249.9	265.9	262.0	265.9	16.0	3.9	8.58	5,872,900	492.9	12.1
303	1435.8	237.0	253.9	249.9	253.9	16.9	4.0	8.96	6,256,800	486.3	12.9
304	1372.0	224.4	242.4	237.0	242.4	18.0	5.4	10.48	6,087,600	369.5	12.6
305	1435.8	211.9	230.0	224.4	230.0	18.1	5.6	10.68	6,072,300	392.7	12.5
306	1708.2	198.8	216.7	211.9	216.7	17.9	4.8	9.96	6,293,200	369.9	11.1
307	1645.4	183.0	203.4	198.8	203.4	20.4	4.6	10.62	6,510,600	372.6	15.8
308	1797.3	168.3	188.5	178.0	188.5	20.2	10.5	14.83	4,627,900	173.6	9.7
309	1446.2	148.9	173.1	161.0	173.1	24.2	12.1	17.46	5,731,400	227.8	12.1
310	1739.7	130.0	153.3	135.8	153.3	23.3	17.5	20.35	2,708,600	76.5	5.8
311	1608.7	101.1	136.1	104.2	136.1	35.0	31.9	33.48	1,428,200	26.5	3.1
312	4192.0	61.0	170.8	80.0	120.8	59.8	40.8	49.74	24,668,240	118.3	19.0
201	336.0	273.0	273.3	273.9	284.0	0.3	10.1	2.79	412,700	276.1	0.9
208	1173.8	178.0	190.0	183.0	203.4	12.0	20.4	15.85	3,445,000	185.2	5.0
209	1524.8	161.0	177.6	168.3	190.0	16.6	21.7	19.47	3,475,500	117.1	7.3
210	1629.6	135.8	137.3	148.9	177.6	21.5	28.7	24.91	6,249,900	154.0	13.1
211	1886.4	116.8	137.4	130.0	157.3	20.6	27.3	23.84	6,229,000	138.5	13.2
212	1860.2	104.2	122.7	114.0	137.4	18.5	23.4	20.85	4,603,100	118.7	9.8
213	2148.4	91.2	107.8	101.1	122.7	16.6	21.6	19.08	4,664,700	113.3	9.9
214	2305.6	85.0	98.5	91.2	107.8	13.5	16.6	14.98	2,901,100	84.0	6.2
215	3039.2	80.0	(89.0) 90.0	85.0	113.5*	10.0	28.5	17.62	2,337,300	43.6	5.0

Total 91,306,800 excluding HX-312

Explanation of Headings

Hx = Heat Exchanger No.
Surface Area = FT^2 (Outside)
B_i = Brine in
C_o = Condensate Out
B_o = Brine Out
C_i = Condensate in
ΔT_i = $C_o - B_i$ in °F
Δt_{swf} = $B_o - B_i$

ΔT_o = $C_i - B_o$ in °F
LMTD = $\dfrac{(\text{Larger } \Delta T - \text{Smaller } \Delta T) \text{ in } ^\circ F}{(\text{Natural log } \dfrac{\text{Larger } \Delta T}{\text{Smaller } \Delta T})}$
Q_H = exchanger duty in BTU/HR
"U" = over-all heat transfer coefficient in BTU/SQ-FT/HR

*Assumed losses to atmosphere reduced inlet temperature of condensate

$Q_H = 56,988,500$ excluding 312
$Q_H = 34,318,300$

(HX-300's) when the corresponding cross-exchanger (HX-200's) was removed. This can be noticed from Figures 4.32 and 4.33 taken from Development Report Number 6, pages 12-13.

Removal of cross-exchangers from the ME-LTV process no doubt reduces to some extent the steam economy (see Table 4.17) but the advantages were stated (Development Report Number 6) to be reduced capital and maintenance costs, lower seawater feed pumping costs, and improved process simplicity.

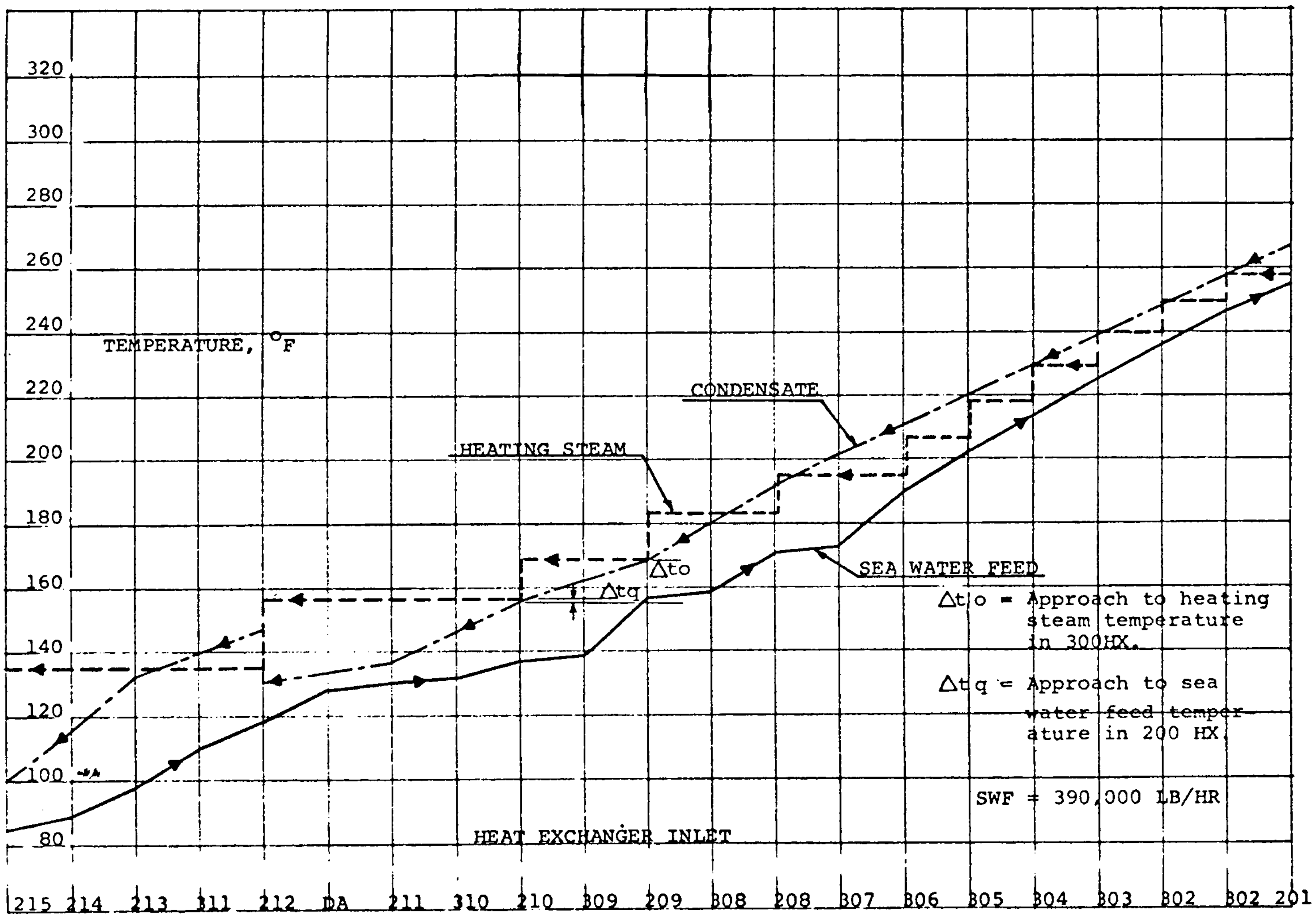


Figure 4.32. Temperature Profiles HMB 8-5 All-Flash

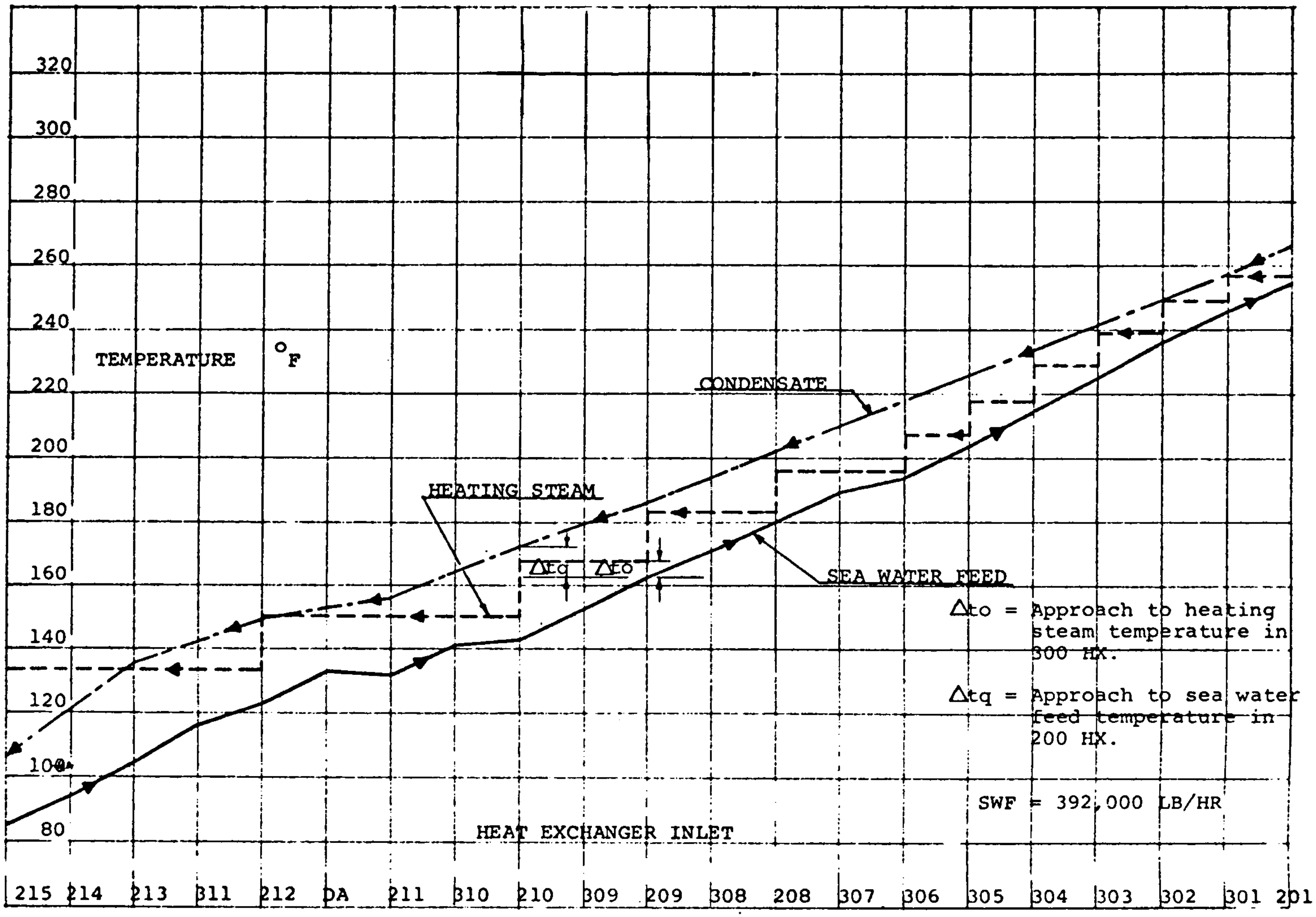


Figure 4.33. Temperature Profiles HMB 8-6 Normal Flash

Table 4.17

Plant Steam Economy

Date	Gross Production (gallons)	*Purchased Dow Steam (pounds)	Economy Gross**	Economy Net	Remarks
10-21	1,054,980	840,000	10.46	9.46	6 Effects on flash (Effects 6-7)
10-25	1,024,460	871,000	9.80	8.80	9 Effects on flash (Effects 2-10)
10-26	1,036,610	857,000	10.08	9.09	8 Effects on flash (Effects 2-4,6-10)
10-27	1,053,120	848,000	10.34	9.34	9 Effects on flash (Effects 3-11)
10-28	1,060,910	774,000	11.42	10.42	3 Effects on flash (Effects 1, 10, 11)

*Purchased Dow Steam based on billing totals from Dow Chemical Company.
**Gross production based on plant product meters.

Actual plant operations during this run disclosed some equipment limitations: inadequate condensate piping sizes and similar preheater condensate drain piping limitations, resulting in high condensate levels throughout the process. However, these problems were removed in the next shutdown period.

Development Runs 11-1 and 11A-18 (OSW R&D Number 440) compared the "normal operation" (using 200 and 300 Series HX's) with an "all-flash" operation (only HX-300's). It was found that with the elimination of 9% (6200 ft^2) of the preheating surface there was only 3% reduction in the steam economy (from 10.3 to 9.9). There are, in addition, savings of condensate pumping costs, maintenance costs, and capital costs of 200 Series exchangers.

2. Preheaters in the 17-Effect Freeport Plant

Improved steam economy and reduced water cost expectations from a 17-effect plant led to its adoption in fiscal year 1967. With this major modification to the original 12-effect plant the preheat train was also modified. Series 200 exchangers were eliminated (except the product coolers HX-212, 213) while the five-effect module had an additional five exchangers of U-tube design (HX-312 through 317) installed inside the module itself, thus saving shell and piping costs. The new heat exchanger configuration is given in Table 4.18.

Table 4.18

Heat Exchanger Heat Transfer Configuration - Development Runs Number 13, 14, and 15

Exchanger Number	Outside Heat Transfer Area A_o^*	Open Tubes	Tube Plugs	Tube Material	OD (inches)	BWG	Length (feet)	Tube Sheet Thickness	
								Top	Bottom
PREHEAT EXCHANGERS	(ft^2)								
200	519	248	2	Admiralty	1	18	8	1-15/16	1-15/16
302a	1,796	343	0	Al-Brass	1	18	20	2	2
302b	1,555	297	44	Cu-Ni 90-10	1	18	20	2	2
303a	1,377	263	80	Admiralty	1	18	20	2	2
303b	1,435	274	69	Al-Brass	1	18	20	2	2
304	2,293	438	22	Al-Brass	1	18	20	2	2
305	3,011	575	5	Al-Brass	1	18	20	2-15/32	2-15/32
306a	1,696	324	1	Al-Brass	1	18	20	2-27/32	2-27/32
306b	1,791	342	1	Al-Brass	1	18	20	2	2
307	1,791	340	1	Cu-Fe	1	18	20	2	2
---	---	2		Al-Brass	1	18	20	--	--
308	1,785	312	2	Al-Brass	1	18	20	2	2
---	---	29	--	Cu-Ni 90-10	1	18	20	--	--
309	1,796	343	0	Al-Brass	1	18	20	2	2
310	1,765	274	6	Admiralty	1	18	20	2	2
---	---	63	--	Cu-Ni 90-10	1	18	20	--	--
311	1,738	332	11	Al-Brass	1	18	20	2	2
312	1,979	504	0	Cu-Ni 90-10	3/4	20	20	1-1/2	1-1/2
313	1,455	180	0	Cu-Ni 90-10	3/4	20	20	1-3/4	--
314	1,455	180	0	Cu-Ni 90-10	3/4	18	U	1-3/4	--
315	1,765	218	0	Cu-Ni 90-10	3/4	18	U	2	--
316	2,420	370	0	Cu-Ni 90-10	3/4	18	U	2-1/4	--
317	2,420	370	0	Cu-Ni 90-10	3/4	18	U	2-3/4	--
TOTAL	35,842	---	--	---	---	--	--	--	--
PRODUCT COOLERS									
212	1,845	352	108	Cu-Ni	1	18	20	2-15/32	2-15/32
213	2,140	409	51	Admiralty	1	18	20	2-15/32	2-15/32
TOTAL	3,985	---	--	---	---	--	--	--	--
HEAT REJECTION CONDENSER									
318	2,220	---	--	Cu-Ni 90-10	3/4	18	--	--	--

$^*A_o = \frac{D_o}{12} L_T N$ Where: D_o = Tube Outside Diameter, inches

N = Number of Open Tubes

L_T = Total Tube Length (Including Tube Sheet), FT

3. Heat Transfer Rates in the Preheaters

As discussed earlier in subsection H.1 (and Table 4.16) the heat flux in Series 200 cross-exchangers is about 2100 Btu/hr ft^2 versus 330 Btu/hr ft^2 flux in the Series 300 bleed vapor condensers. This is not unexpected, though, since liquid-to-liquid coefficient is predicted by heat transfer theory to be lower than the condensing vapor-to-liquid coefficient. It must be pointed out here that the liquid is not flowing as a film (as in the evaporator effects) but the whole tube cross-section is full of liquid (i.e., seawater feed) in both Series 200 and 300 exchangers. Any decrease in plant capacity, therefore, adversely affects the liquid-to-liquid exchangers (Series 200) more than the vapor-to-liquid preheaters (Series 300). This is shown in Tables 4.19, 4.20, and 4.21. It is to be noted that HX-215 coefficient is higher with lower seawater feed rate in Run Number 10A-5. This is explained by the fact that during this run <u>all</u> of the condensate product was being passed through the shell side of this exchanger, instead of only Freeport product (less than 50% of the total) as in Run 10-3. Greater flows produce higher velocities and more intense turbulence so that the individual coefficient on the greater flow side is augmented.

The history of the overall coefficient of heat transfer in both the Series 200 and 300 preheaters is tabulated in Tables 4.22 and 4.23. The U data was available for only the particular runs included in the table. For quick appraisal of the heat transfer performance of a given exchanger over the plant life these data are graphically displayed in Figures 4.34, 4.35, and 4.36. For greater clarity the Series 300 exchanger data was split into two parts (HX-301 through 306, and HX-307 through 312) and plotted separately in Figures 4.35 and 4.36.

4. Preheaters in the 17-Effect Plant

When the Freeport Plant was modified from 12 to 17 effects, the preheat circuit was considerably changed. Additional surface requirements in the high temperature end were met by using old exchangers in parallel without tubing them out to the fullest. The old and new heat exchanger numbers along with the design overall coefficient U values are given in Table 4.24 taken from the Stearns-Roger document, "Feasibility Studv - A Study to Increase the Efficiency of the Freeport Demonstration Plant," submitted to OSW in September 1965. Development Runs 13, 14, 15 and 16 were carried out at Freeport after this modification to 17-effect operation. The preheater configuration for these runs is given in Table 4.18. The average overall coefficients for the Development Runs 14, 15 and 16 are presented in Table 4.25 (taken from Development Report Number 9 by Stearns-Roger, page 42).

Table 4.19

Heat Exchanger Performance Summary, Heat and Material Balance Data 10-3

HX	TEMPERATURE °F				TEMP DIFF Δt °F				SURFACE AREA FT^2	DUTY BTU/HR	"U"
	BRINE		CONDENSATE		$C_o - B_i$	$B_o - B_i$	$C_i - B_o$	LMTD			
	IN (B_i)	OUT (B_o)	IN (C_i)	OUT (C_o)	(ΔT_i)	(Δt_{swf})	(ΔT_o)				
301	249.6	260.6	264.0	264.0	14.4	11.0	3.4	7.62	1,795.9	5,227,850	382.0
302	238.8	249.6	253.6	253.6	14.8	10.8	4.0	8.25	1,387.5	5,129,160	448.1
303	223.0	238.8	242.3	242.3	19.3	15.8	3.5	9.26	1,434.7	7,528,000	566.6
304	213.6	223.0	231.0	231.0	16.4	9.4	8.0	13.11	1,560.3	4,473,770	218.7
305	202.9	213.6	219.5	219.5	16.6	10.7	5.9	10.37	1,795.9	5,107,180	265.8
306	192.8	202.9	208.9	208.9	16.1	10.1	6.0	10.23	1,670.3	4,832,080	282.8
307	180.7	192.8	197.1	197.1	16.4	12.1	4.3	9.04	1,633.6	5,761,760	390.2
308	170.5	177.1	184.5	184.5	14.0	6.6	7.4	10.34	1,701.9	3,122,150	177.4
309	152.9	162.7	169.8	169.8	16.9	9.8	7.1	11.32	1,445.1	4,668,080	285.4
310	133.3	143.1	151.3	151.3	18.0	9.8	8.2	12.47	1,738.4	4,689,600	216.3
311	99.5	104.2	135.7	135.7	36.2	4.7	31.5	34.06	1,701.7	2,236,690	38.6
312	71.3	110.0	119.5	103.0	31.7	38.7	9.5	18.41	4,192.0	22,279,930	288.7
201	260.6	261.5	274.7	262.8	2.2	0.9	13.2	6.14	436.0	401,020	156.0
208	177.1	180.7	197.1	190.0	12.9	3.6	16.4	14.58	1,174.0	1,695,250	101.6
209	162.7	170.5	188.7	174.7	12.0	7.8	18.2	14.89	1,525.0	3,727,620	180.4
210	143.1	152.9	174.2	158.2	15.1	9.8	21.3	18.10	1,630.0	4,686,120	174.6
211	119.8	133.3	158.2	136.2	16.4	13.5	24.9	20.29	1,886.0	6,452,040	194.9
212	104.2	115.9	136.2	117.2	13.0	11.7	20.3	16.37	1,860.0	5,572,580	217.9
213	87.5	99.5	117.2	99.2	11.7	12.0	17.7	14.53	2,148.0	5,278,760	213.3
214	78.4	87.5	99.5	85.5	7.1	9.1	12.0	9.33	2,306.0	4,403,220	272.9
215	71.3	78.4	115.5	75.8	4.5	7.1	37.1	15.45	3,039.0	3,459,380	91.0
TOTAL						186.3*			37,935.3		
						3.9 DA Heating					

ACTIVE PLANT SURFACE AREA:	FT^2
EVAPORATING	48,684
PREHEATING	17,739
HEAT RECOVERY	16,004
HEAT REJECTION	4,192
TOTAL ACTIVE SURFACE	86,619

SURFACE EFFECTIVENESS

4.15 lb/hr-ft^2 of total surface

7.38 lb/hr-ft^2 of evaporating surface

*Total SWF Preheat

SWF Rate, lb/hr 495,000

Vapor Rate, lb/hr 327,660

"U" = overall heat transfer coefficient in BTU/SQ-FT/HR

LMTD = $\dfrac{(\text{Larger } \Delta T - \text{Smaller } \Delta T)}{\left\{\text{Natural log } \dfrac{\text{Larger } \Delta T}{\text{Smaller } \Delta T}\right\}}$ in °F

Table 4.20

Heat Exchanger Performance Summary, Heat and Material Balance Data 10A-5

HX	TEMPERATURE °F				TEMP DIFF Δt °F				SURFACE AREA FT²	DUTY BTU/HR	"U"
	BRINE		CONDENSATE		$C_o - B_i$	$B_o - B_i$	$C_i - B_o$	LMTD			
	IN (B_i)	OUT (B_o)	IN (C_i)	OUT (C_o)	(ΔT_i)	(Δt_{swf})	(ΔT_o)				
301	214.3	220.3	222.7	222.7	8.4	6.0	2.4	4.79	1,797.3	1,295,280	162.3
302	206.2	214.3	215.9	215.9	9.7	8.1	1.6	4.79	1,388.6	2,097,020	315.3
303	197.6	206.2	207.0	207.0	9.4	8.6	0.8	3.49	1,435.8	2,235,920	444.9
304	188.9	197.6	198.5	198.5	9.6	8.7	0.9	3.52	1,561.5	1,986,720	372.0
305	178.7	188.9	190.1	190.1	11.4	10.2	1.2	4.53	1,797.3	2,636,830	323.9
306	168.4	178.7	180.0	180.0	11.6	10.3	1.3	4.71	1,671.6	2,683,290	341.5
307	157.9	168.4	169.9	169.9	12.0	10.5	1.5	5.05	1,634.9	2,720,140	329.5
308	145.6	154.5	160.0	160.0	14.4	8.9	5.5	9.25	1,703.0	2,305,340	146.2
309	133.5	140.8	148.6	148.6	15.1	7.3	7.8	11.05	1,446.2	1,908,120	119.5
310	119.7	127.5	136.8	136.8	17.1	7.8	9.3	12.80	1,739.7	2,031,590	91.2
311	96.9	102.8	123.4	123.4	26.5	5.9	20.6	23.51	1,703.0	1,537,920	38.4
312	67.4	75.0	109.2	70.0	2.6	7.6	34.2	12.26	4,190.0	13,331,320	338.9
201	220.3	220.9	231.0	224.0	3.7	0.6	10.1	6.37	436.8	151,950	55.2
208	154.5	157.9	169.9	162.6	8.1	3.4	12.0	9.92	1,173.8	887,970	79.9
209	140.8	145.6	162.6	152.5	11.7	4.8	17.0	14.21	1,524.8	1,233,400	59.6
210	127.5	133.5	152.5	139.7	12.2	6.0	19.0	15.35	1,629.6	1,566,530	85.1
211	112.9	119.7	139.7	125.2	12.3	6.8	20.0	15.88	1,886.4	1,775,650	62.4
212	102.8	110.2	124.7	114.2	11.4	7.4	14.5	12.97	1,860.2	1,921,590	81.3
213	83.8	96.9	114.8	97.3	13.5	13.1	17.9	15.60	2,148.4	3,435,480	125.9
214	75.5	83.8	95.5	85.0	9.5	8.3	11.7	10.56	2,305.6	2,188,920	104.6
215	67.4	75.5	85.0	74.7	7.3	8.1	9.5	8.35	3,039.2	2,147,070	109.4
TOTAL						150.8*			38,073.7		
						2.7	DA Heating				

ACTIVE PLANT SURFACE AREA:	FT²
EVAPORATING	47,862
PREHEATING	17,878.9
HEAT RECOVERY	16,004.8
HEAT REJECTION	4,190
TOTAL ACTIVE SURFACE	85,935.7

SURFACE EFFECTIVENESS

2.37 lb/hr-ft² of total surface

4.26 lb/hr-ft² of evaporating surface

*Total SWF Preheat

SWF Rate, lb/hr 270,000

Vapor Rate, lb/hr 187,740

"U" = overall heat transfer coefficient in BTU/SQ-FT/HR

$$\text{LMTD} = \frac{(\text{Larger } \Delta T - \text{Smaller } \Delta T)}{\left(\text{Natural log } \dfrac{\text{Larger } \Delta T}{\text{Smaller } \Delta T}\right)} \text{ in } {}^{\circ}\text{F}$$

Table 4.21

Effect of Production Rate on 200-Series Heat Exchanger Performance

Data NO.	10-3	10A-5
SWF Rate, LB/HR	495,000	270,000
Extracted Vapor Rate	327,660	187,740
HEAT EXCHANGER NO.	OVERALL HEAT TRANSFER CO-EFFICIENT, BTU/HR/FT^2/^{o}F	
201	156.0	55.2
208	101.4	79.9
209	180.4	59.6
210	174.6	65.1
211	194.9	62.4
212	217.9	81.3
213	213.3	125.9
214	272.9	104.6
215	91.0	109.4

Table 4.22

History of Heat Transfer Rates in Preheat Exchangers, Series 200

Exchanger No.	Development Run No.					
	6-11	7b-2	7c-4	9-2	10-3	10A-5
201	63	152	276	116	156	55
208	446	166	185	181	102	80
209	166	128	117	122	180	60
210	149	169	154	139	175	65
211	115	110	138	126	195	62
212	202	104	119	187	218	81
213	153	140	113	110	213	126
214	172	75	84	61	273	105
215	82	--	44	44	91	109

NOTE: The table contains the overall coefficient U values rounded to whole numbers and having units of Btu/hr ft^2 ^{o}F.

Table 4.23

History of Heat Transfer Rates in Preheat Exchangers, Series 300

Exchanger No.	Development Run No.					
	6-11	7b-2	7c-4	9-2	10-3	10A-5
301	564	525	447	492	382	162
302	546	152	493	468	448	315
303	466	515	486	586	567	445
304	401	440	369	413	219	372
305	556	536	393	413	266	324
306	294	409	370	335	283	341
307	377	297	373	320	390	330
308	203	68	174	342	177	146
309	174	148	228	295	285	119
310	179	112	76	217	216	91
311	123	63	26	51	39	38
312	345	133	118	250	289	339

NOTE: The table contains the overall coefficient U values rounded to whole numbers and having units of Btu/hr ft^2 ^{o}F.

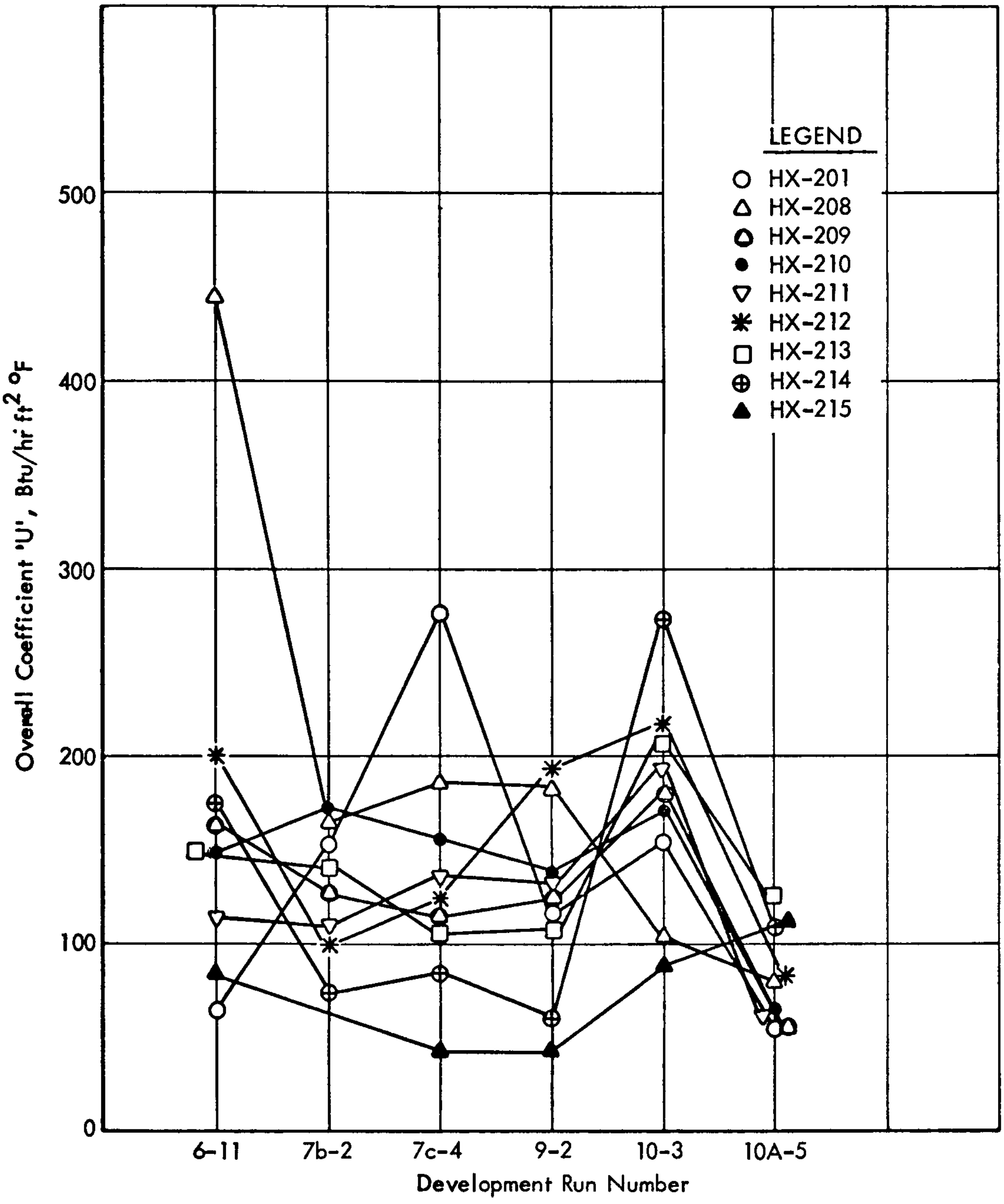


Figure 4.34. History of U in Series 200 Preheaters

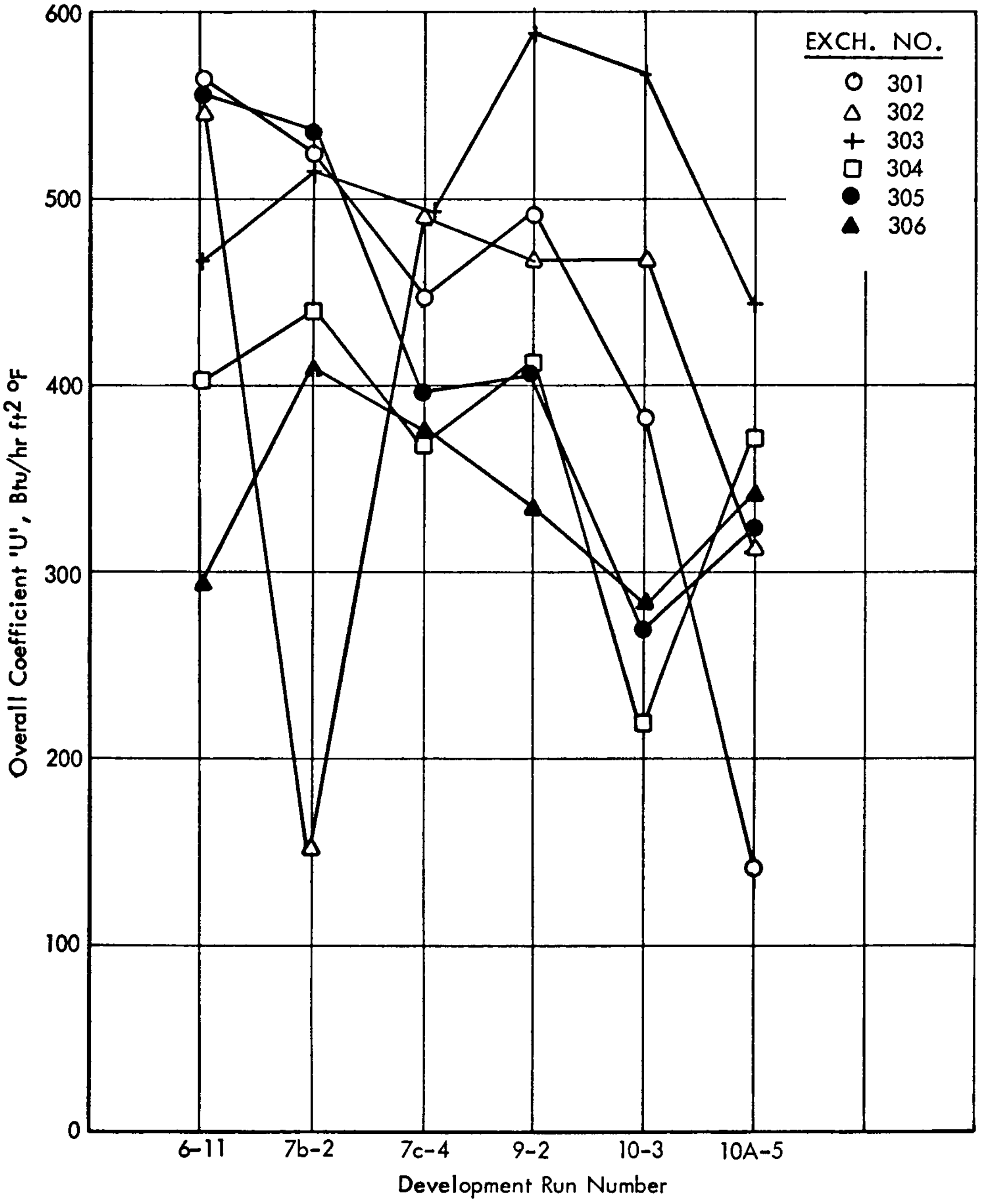


Figure 4.35. History of U in Series 300 Preheaters

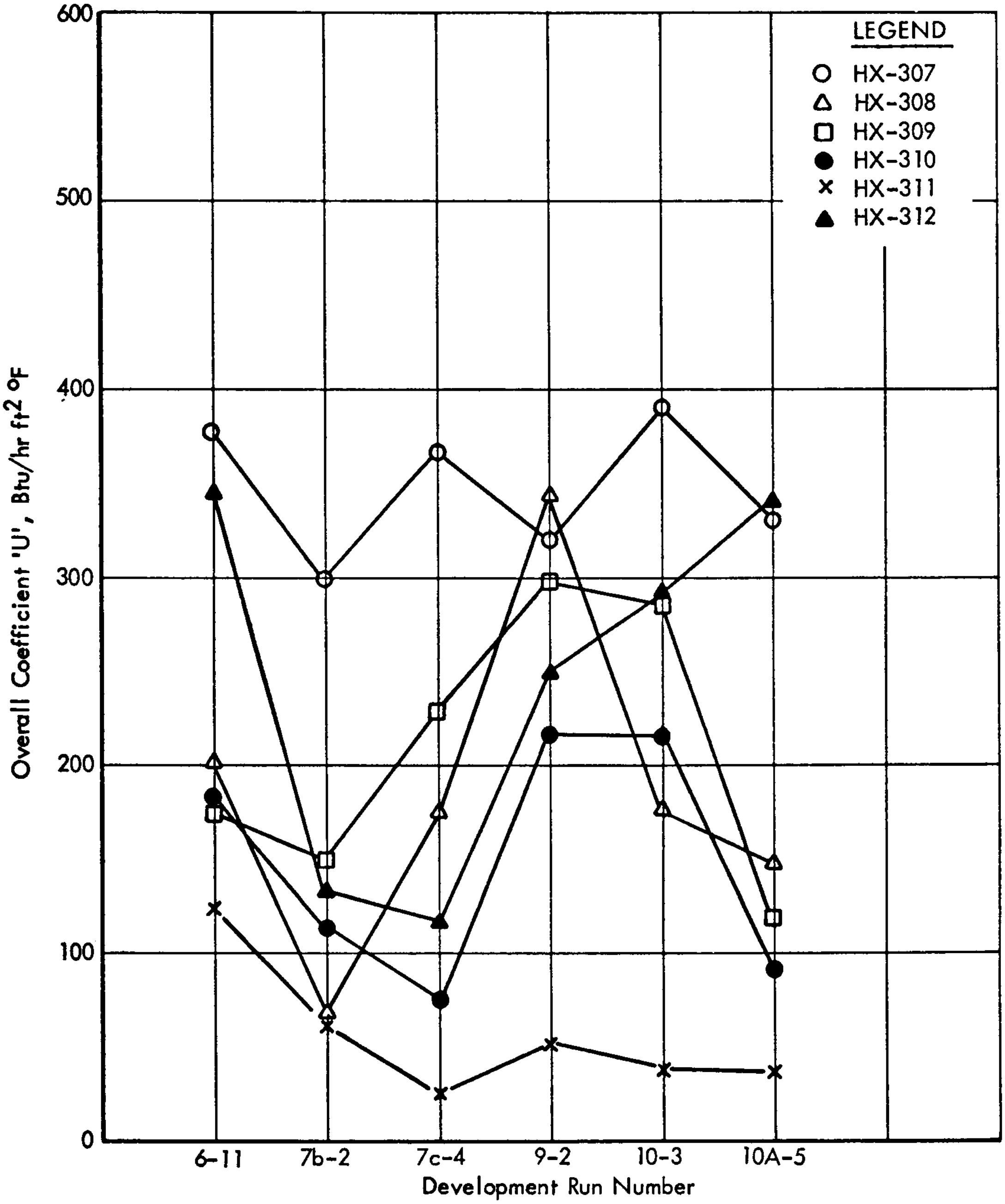


Figure 4.36. History of U in Series 300 Preheaters

Table 4.24

Preheating Circuit Process Summary — 17 Effect Plant

Sea Water Feed Rate 510,000 Lb/Hr   SFR = 20.4

HX NO.		TEMPERATURE ^oF					DUTY (Q_h)	Design "U"	Calculated
NEW	OLD	B_i	C_o	B_o	S_i	LMTD	Btu/Hour	Btu/Hr Ft^2 ^oF	O.S. Surface Req'd, Ft^2
200	201	264	267	264.5	275	5.58	243,530	300	146
302a	301	255	266	264	266	5.26	4,383,450	325	2,563
302b	304								
303a	302	246	257	255	257	5.26	4,383,450	325	2,563
303b	303								
304	214	237	248	246	248	5.26	4,383,450	325	2,563
305	215	227	239	237	239	5.59	4,870,500	325	2,682
306a	208	216	230	227	230	7.14	5,357,550	325	2,368
306b	305								
307	306	206	220	216	220	8.00	4,870,500	325	1,874
308	307	196	210	206	210	8.00	4,870,500	325	1,874
309	308	186	200	196	200	8.00	4,870,500	325	1,874
310	309	176	190	186	190	8.00	4,870,500	325	1,874
311	310	166	180	176	180	8.00	4,870,500	325	1,874
312	311	156	170	166	170	8.00	4,870,500	325	1,874
313	New	146	160	156	160	8.00	4,870,500	350	1,740
314	New	135	150	146	150	8.00	5,357,550	350	1,914
315	New	124	139	135	139	8.33	5,357,550	350	1,838
316	New	111	127	122	127	9.47	5,357,550	350	1,617
317	New	101	114	111	114	6.83	4,860,950	350	2,034
318	New	70	100	80	100	24.69	19,921,600	350	2,305
212	212	80	85	101	113.2	8.09	10,208,000	325	3,883
213	213								
								TOTAL	39,460

C_o Condensate temperature out.
S_i Steam or condensate temperature in.

Table 4.25

Preheater Exchangers - Heat Transfer Coefficients (U_o), Btu/hr, °F, ft^2

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)	N	360*	420*	2.7	400*	290*	390*
HX-302(b)	560*	430	420	3.1	530*	480*	470*
HX-303(a)	9.00*	490*	310*	3.6	860*	540*	570*
HX-303(b)	510	350	480	3.4	490	630*	450
HX-304	370*	150	130	2.1	310*	370*	220*
HX-305	290*	250	260*	1.6	250	340*	270*
HX-306(a)	480*	460*	430*	2.9	475*	370*	460*
HX-306(x)	940	780	930	<8.6? 4.3?	400	330	880$^{(1)}$
HX-307	330	320	270	2.7	420	400	310
HX-308	320	270	300	2.7	390	410	300
HX-309	360	330	280	2.7	355	410	320
HX-310	360	270	300	2.8	330	280	310
HX-311	80	790	260	2.8	380	290	380
HX-312	520*	540*	540*	6.5	430	410*	540
HX-313	380	380	360	9.9 } ?	725	460	380
HX-314	390	310	280	9.9 } ?	445	380	330
HX-315	420	390	350	8.3	440	350	390
HX-316	440*	240	230	4.8	350	330	300
HX-317	320	320	320	4.8	435*	340	320

* = LTD Less than 2°F.
N = LTD Negative
(1) Changed to vapor sheer preheater.

It is seen from this table that the performance of the preheat exchangers in Development Runs 14, 15 and 16 was close to the design expectation. Where the U values were abnormally low the Annual Reports (Numbers 7 and 8) by Stearns-Roger explain this as being due to vapor binding (noncondensibles accumulation), low steam rate due to low production rate, low brine velocities in the old plant, improper or insufficient venting, fouling of new tubes, inleakage of air due to impairment of vacuum seals, or errors in temperature measurement.

In these tests it was shown that the all-flash method of product heat recovery is feasible both technically and economically. When the 200 Series exchangers were removed from the preheat circuit, there was actually an increase in performance of the 300 Series exchangers. This can be noticed from Figure 4.36 where HX-308 through 312 have U values in the range 100 to 250 Btu/hr ft^2 ^{o}F when the 200 Series exchangers were present as against U values of over 300 Btu/hr ft^2 ^{o}F for Runs 14, 15 and 16 after elimination of the Series 200 units (Table 4.18). The U tube preheaters used in the 5-effect module yielded overall heat transfer coefficient close to or greater than the design values.

5. Experimental Horizontal Vapor Shear Preheater

During the shutdown between Runs 15 and 16, a new experimental preheat exchanger was installed at the Freeport Plant. It was designed and furnished to the plant by Oak Ridge National Laboratory, Oak Ridge, Tennessee. This unit was called HX-306x and replaced the unit HX-306b. A comparison of the two preheaters appears in Table 4.26.

This unit was designed to allow two modes of operation on the steam side — full condensing and flow-through. The experimental data gathered over an operational period of 3,100 hours gave overall coefficients in the range 790 to 1100 Btu/hr ft^2 ^{o}F as against 370 Btu/hr ft^2 for the normal HX-306b. This improvement has been attributed to the following (Development Report Number 9, page 58):

- The higher velocity on the brine side (due to smaller number of tubes).
- The spirally indented tube surface.
- The horizontal tube configuration.
- The steamside baffling.

Table 4.26

Comparison of Preheater Data

	Conventional	Experimental
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 inch	1 inch
Tube Wall	0.049 inch	0.035 inch
Tube Length	20 feet	8 feet
Tube Material	Al-Brass	Cu-Nl, 90-10
Vessel Orientation	Vertical	Horizontal
Passes Brine Side	2	1
Passes Steam Side	N/A	3

I. HEAT REJECTION CONDENSER PERFORMANCE

This heat transfer unit has the special duty of condensing the low temperature and hence low pressure vapor from the last effect in a multiple-effect evaporator train, using the raw seawater feed as the coolant. The critical importance of this unit arises from the fact that it controls one of the two temperatures on which depends the overall plant production rate. Any malfunction in this unit which reduces the heat transfer coefficient will raise the condensing temperature in the last effect (as well as the preceding effects) thus reducing the overall plant temperature driving force. Since the production rate is directly related to the overall temperature driving force ($W_L = UA\Delta T/\lambda$), any increase in condensing temperature leads to reduced production.

For higher thermodynamic efficiency, according to the Second Law of Thermodynamics, this unit should operate at the lowest possible temperature, and therefore at the lowest temperature difference, since seawater temperature is fixed (not under operator control).

In the original 12-effect plant, HX-312 was the heat rejection condenser. The condenser can be operated either as 4-pass (tube side) or 2-pass unit. This is shown in Figure 4.37. Although the pump work is increased in the two-pass mode this results in improved operation and hence improved overall steam economy (Third Annual Report by Stearns-Roger, page 135).

The overall heat transfer coefficients for these exchangers are given in Tables 4.23 and 4.25. It has been reported (Fourth Annual Report by Stearns-Roger, page 126, paragraph 5.125) that the overall coefficient should be in excess of 400 Btu/hr ft^2 $^{\circ}$F. Therefore, the actual performance of the condenser is below expectations. The subcooling of the condensate by as much as 20°F was also indicative of the poor performance. Probable reasons for this were stated to be:

- inadequate venting,
- accumulation of condensate in the steam chest,
- improper arrangement of tubes, and
- baffling of the steam side.

It was suggested in the Fourth Annual Report that the condenser be replaced with one of good design capable of operating at 2 in. Hg pressure. The surface requirement could also be reduced from 4192 ft^2 (the then active surface) to about 2500 ft^2.

With the major modification to 17-effects the HX-312 was modified and used as preheat exchanger, while a new exchanger HX-318 was installed to condense vapors from Effect XVII. Table 4.27 compares the performance of this exchanger during Run 15 with the approximate design values. The low U values indicate that the performance was not as expected. Changing of the coolant water inlet point (normally at the top) to the bottom of the exchanger did not appreciably change the coefficient. The steam approach temperature remained approximately 25°F, even when the inlet seawater temperature was changed by 10°F.

During the final Development Run 16 the overall coefficient ranged from 160 to 450 Btu/hr ft^2 $^{\circ}$F, which is below the design value. Performance of this condenser fell off significantly during cold weather operations, according to the Eighth Annual Report by Stearns-Roger (page 15). No reasons have been given in the report for this below-design performance. Since this type of surface condenser is widely used in the power generation industry, its design and operation should present no problems.

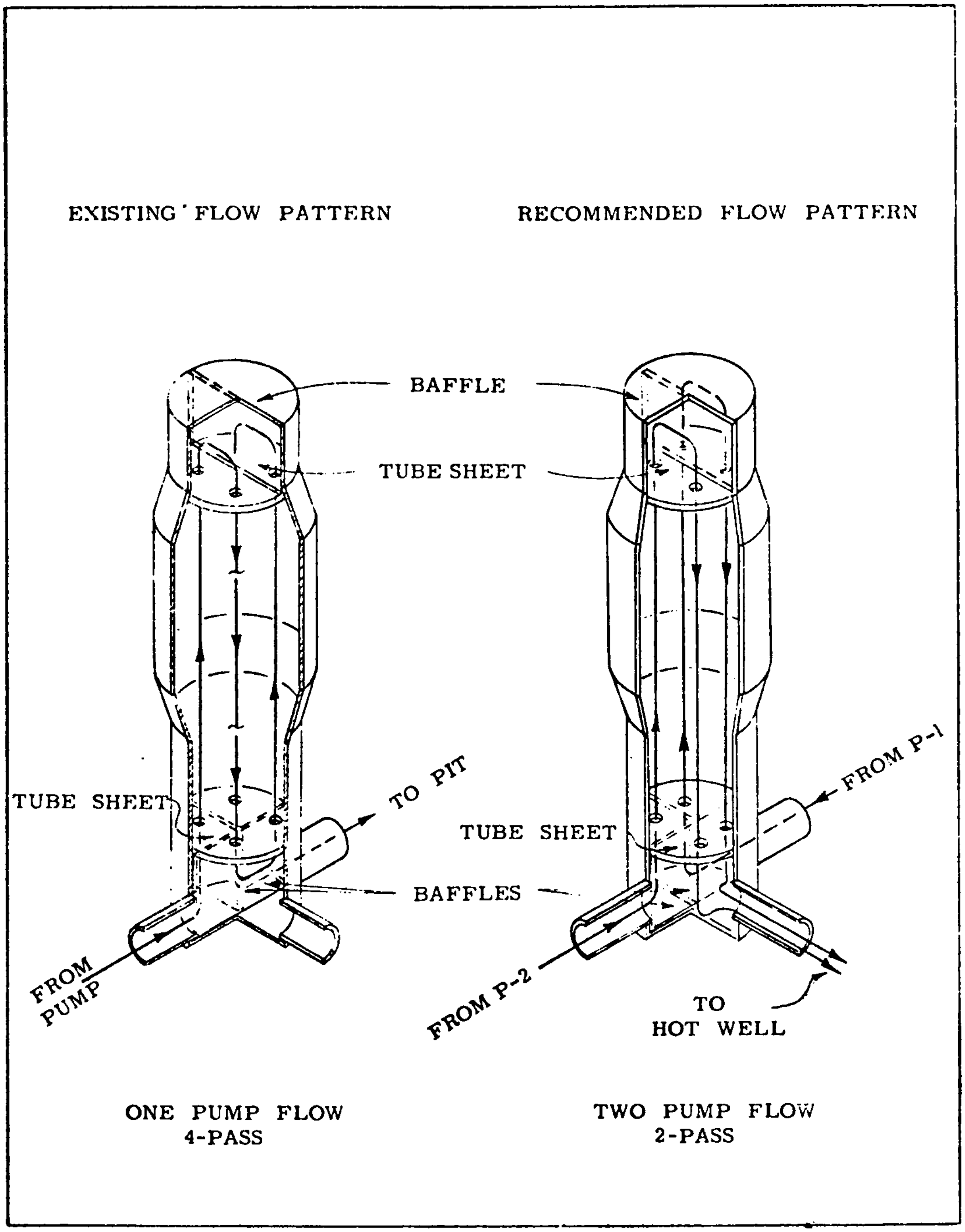


Figure 4.37. 312 Condenser Flow Patterns

Table 4.27

Heat Rejection Condenser Performance — HX-318

DEVELOPMENT RUN NO. 15				
HEAT AND MATERIAL BALANCE NUMBER	1	3	4	APPROX
DATE	5/2/68	5/28/68	6/11/68	DESIGN
I. DATA				
A. Cooling Brine to HX-318				
1. Brine Outlet Temperature, $^{\circ}F^{(1)}$	84.1	91.6	93.8	90
2. Brine Inlet Temperature, $^{\circ}F^{(1)}$	73.6	79.9	81.4	75
3. Cooling Brine Enters Exchanger From:	Top	Bottom	Bottom	Top
4. Cooling Brine Rate (Calculated From Heat Balance), MM LBS/HR	1.75	1.55	1.40	1.37
B. Vapor to HX-314				
1. Vapor into HX-318 Temperature, $^{\circ}F^{(2)}$	97.6	103.8	107.8	100
2. Condensate out of HX-318 Temperature, $^{\circ}F^{(3)}$	94.8	103.0	105.2	---
3. Condensate Rate (Measured with Flow Meter at P-42 Discharge), K LBS/HR	16.8	16.6	16.0	20.0
C. Barometric Condenser Brine Outlet Temperature, $^{\circ}F^{(1)}$	85.1	93.5	95.2	---
II. CALCULATED VALUES				
A. Steam Approach to Cooling Brine, $^{\circ}F$	24.0	23.9	26.4	25
B. HX-318 LMTD, $^{\circ}F$	18.3	17.4	19.6	16.4
C. Overall Heat Transfer Coefficient, BTU/HR/FT^2/$^{\circ}F$	430	450	380	530

1. Uncalibrated 30-124°F thermometer with 0.2°F smallest scale division, probable accuracy of ± 0.2°F.
2. Certified thermometer with 0.2°F smallest scale division.
3. Uncalibrated 30-120°F thermometer with 1.0°F smallest scale division.
4. Effective heat transfer = 2220 FT^2.

Chapter 5

SCALE CONTROL HISTORY

The production of pure water from seawater by the LTV falling-film distillation process involves preheating seawater to the saturation temperature corresponding to the first effect and then separation by evaporation of pure water from the seawater. Both of these steps can lead to supersaturation with respect to some or all of the dissolved salts in the seawater and finally to their deposition on the heat transfer surfaces, piping, valves, etc. This mineral deposit or scale has low thermal conductivity so that it acts almost like insulation; hence, heat transfer rates are significantly reduced. Since the plant production rate is directly dependent on the rates of heat transfer, it is obvious that scaling is undesirable. Scaling is also undesirable in piping, valves, etc., because it reduces flow cross-sections and increases pumping costs. From the operational point of view, scale on indicating and controlling instrument sensors can lead to complete process upset or breakdown. A more important aspect of scaling, from the design point of view, is the fact that scaling thresholds determine the maximum temperature driving force ΔT that can be used. This maximum ΔT limits the number of evaporator effects that can be used (and therefore, the thermal economy) and ultimately the plant capacity. Unit water cost depends directly on these two factors. Higher thermal economies and capacities lead to reduced water costs. Evaluation of scale control methods and the maximum temperatures attainable without scaling were two of the prime objectives of the Freeport Demonstration Plant operation. The discussion in this section is divided into the following subsections:

A. Theory of Scale Formation.

B. Scale Prevention Techniques.

C. Effectiveness of Scale Prevention Methods Tried.

D. Variation in U with Scale Control Methods.

E. Scale Deposition Rate.

F. Scale Removal Techniques.

G. Pretreatment Methods.

A. THEORY OF SCALE FORMATION

Scale is defined[1] as a deposit formed by precipitation from solution of a substance which has a solubility that decreases with an increase in temperature. An important characteristic of these scales is that generally they are very adherent to the surface so that strong physical methods (e.g., drilling) have to be adopted to remove them. If scales were non-adherent, the scaling problem would indeed be simplified to a great extent.

Scaling problems have been known for a long time and extensive research has been carried out to understand various aspects of scaling and scale prevention. Some 900 abstract cards were prepared by W. L. Badger and Associates under contract with the Office of Saline Water in 1959. The Saline Water Conversion Report by OSW each year lists a number of projects wherein scaling research is being carried out. Despite this effort, Badger has aptly commented (OSW R&D Report Number 25) that "the causes of its formation and means of prevention have not always been (and are not now) perfectly understood." Scaling and scale-prevention have, therefore, been called both an art and a science.

1. Mechanisms of Scale Formation

It was generally assumed, before 1925, that the following two steps were involved in scale formation:

- Precipitation in the bulk of the liquid of the scaling substance.
- Impingement of the precipitated particles on the surrounding surface, and compaction leading to scale formation.

Some instances of scale formation by this mechanism may occur, for instance, when suspended fine solids settle out at points of low velocity (turbulence) or when fluid circulation is poor or non-existent. However, subsequent research

---

(1) OSW R&D Progress Report Number 25, page 1, (July 1959).

has disproved this theory for the most common occurrences of scale formation. Partridge and White[1] mention a 42-day chemical balance test on a boiler where 89 percent of the entering $CaSO_4$ was deposited as anhydrite scale while only 3 percent was found in the sludge from the boiler blowdown. These same authors showed experimentally that scale formation initiates at the heated surface rather than in the bulk of the solution.

The currently accepted mechanism involves the following steps:

- A solution of salt or salts having inverted solubility characteristics is heated and/or evaporated. Supersaturation results. This is highest near the hot walls.
- Cluster or nuclei formation takes place — i.e., a few hundred molecules of scaling substance come together to form a fragment of a new phase.
- Since this cluster has high surface energy per unit volume, and is under constant bombardment by other molecules to break it up, it attaches itself to the walls or the enclosing surface.
- Supersaturation being highest near the heated surface, this cluster induces further shower of nucleii and crystal growth takes place. Scale becomes thicker as crystallization proceeds.

The above-mentioned steps in scale formation can be summarized as: supersaturation from heating or evaporation, nucleation, attachment of nucleii to walls or enclosing surface, nucleii growth and finally hard, adherent scale formation. Any factor that affects any of these steps affects scale formation. The following is a brief list of such factors:

- Chelating, sequestering or complexation of the scale forming ions, e.g., with EDTA. This essentially means preventing supersaturation by keeping the ions in solution.
- Initial formation of clusters (tiny crystals) is extremely difficult. Carefully filtered solutions can be taken to concentrations

---

(1) Partridge, E.P. and A.H. White, Ind. Eng. Chem. 21, Number 9, 834 (1929)

much higher than the equilibirium value without crystal formation.[1] Presence of solid particles, dust, seed crystals, or the metal walls helps nucleation by reducing the free energy barrier.[2] This is another reason for the adherence of the crystals to the heating surface.

- Attachment of the scale nucleii to the wall can be affected by a number of factors such as the presence of starch particles that swell on hydration, or modification of the wall surface by coating it with epoxy resin. Fabuss and Lu of Monsanto Research Corporation, who submitted R&D Report Number 258 to OSW in 1967, report (on page 86) that calcium sulfate precipitation was considerably less on an epoxy resin coated surface than on a stainless steel surface.

- Nucleation and nuclei growth involve an induction period and a certain detention time. At low supersaturations this time could be a few hours. At the San Diego Plant of OSW, brine residence time being only about 100 seconds, scale formation could be avoided during many experiments (OSW R&D Report Number 133, page 93). Dispersants like polyphosphates, tannin, and ligno-sulfonates are assumed to act by triggering nuclei growth termination reactions.

The earlier statement that scaling is both an art and a science can perhaps be better appreciated in the light of the few factors mentioned above. It is the difficulty of predicting the net effect of so many factors which affect scaling that makes for its complexity.

2. Scale Forming Reactions

Seawater varies in composition around the world but it has been found that the ratio of the various components is generally constant. For comparative purposes a standard is prepared (and supplied) by the Hydrographic Laboratories of Copenhagen, Denmark. The ionic composition of this "normal water" is given in Table 5.1.

---

(1) LaMer, V.K., Ind. Eng. Chem. 44, 1270 (1952)
(2) OSW R&D Report Number 500.

Table 5.1

Composition of "Normal Water"

Component	Parts per million	Grams per liter at 20°C
Na	10,767.8	11.0348
Mg	1,297.5	1.3297
Ca	408.1	0.4182
K	387.6	0.3972
Sr	13.6	0.0139
Cl	19,360.5	19.8406
SO_4	2,701.7	2.7687
HCO_3	142.5	0.1460
Br	65.9	0.0675
F	1.3	0.0014
H_3BO_3	26.5	0.0272
I	0.05	0.00005
Si	0.02 to 4.0	0.00002 to 0.004
Others	1.3	
Total Solids	35,174.5	36.0468
Water	964,825.5	988.7532
Total	1,000,000.0	1,024.8

Total alkalinity — 119.8 mg/l.

This normal water has a chlorinity of 19.381 corresponding to a salinity of 35.01, and a density of 1.0248 at 20°C.

If the elements in the above table are combined in the form in which they usually deposit, then the seawater with a chlorinity of 19.0 (slightly diluted compared to normal seawater) would have the composition given in Table 5.2.

Table 5.2

Composition of Seawater in Terms of Compounds

Compound	Parts by weight per million
NaCl	26,780
$MgCl_2$	3,214
$MgSO_4$	2,233
$CaSO_4$ $2H_2O$	1,548
KCl	725
$CaCO_3$	109
NaBr	83
Water, etc.	965,308

The analysis assumes that bicarbonate decomposes to carbonate before precipitation occurs, and that strontium is analyzed with calcium (OSW R&D Report Number 25, page 33).

There are mainly two types of scale: alkaline scale and acid scale. Alkaline scale arises from decomposition of the bicarbonate (HCO_3) ion. On heating seawater the following reaction occurs below approximately 180°F:

$$2HCO_3^- \rightleftharpoons CO_2 + CO_3^{--} + H_2O \tag{5.1}$$

$$Ca^{++} + CO_3^{--} \rightleftharpoons CaCO_3 \tag{5.2}$$

Solubility product of $CaCO_3$ in pure water at 25°C is given approximately as (OSW R&D Report Number 133, page 13):

$$\left[Ca^{++}\right]\left[CO_3^{--}\right] = K_{sp} = 10^{-8} \tag{5.3}$$

so that $CaCO_3$ has extremely low solubility, supersaturation is reached very quickly, and it precipitates out at very low concentrations.

At higher temperatures (>180°F approximately) the carbonate ion produced in reaction Equation 5.1 undergoes further reaction with water to produce CO_2 (gas) and OH^- ions:

$$CO_3^{--} + H_2O \rightleftharpoons CO_2 + 2OH^- \qquad (5.4)$$

$$Mg^{++} + 2(OH) \rightleftharpoons Mg(OH)_2 \qquad (5.5)$$

The solubility product for $Mg(OH)_2$ is even smaller than that for $CaCO_3$:

$$\left[Mg^{++}\right]\left[OH^-\right]^2 = K_{sp} = 10^{-11} \qquad (5.6)$$

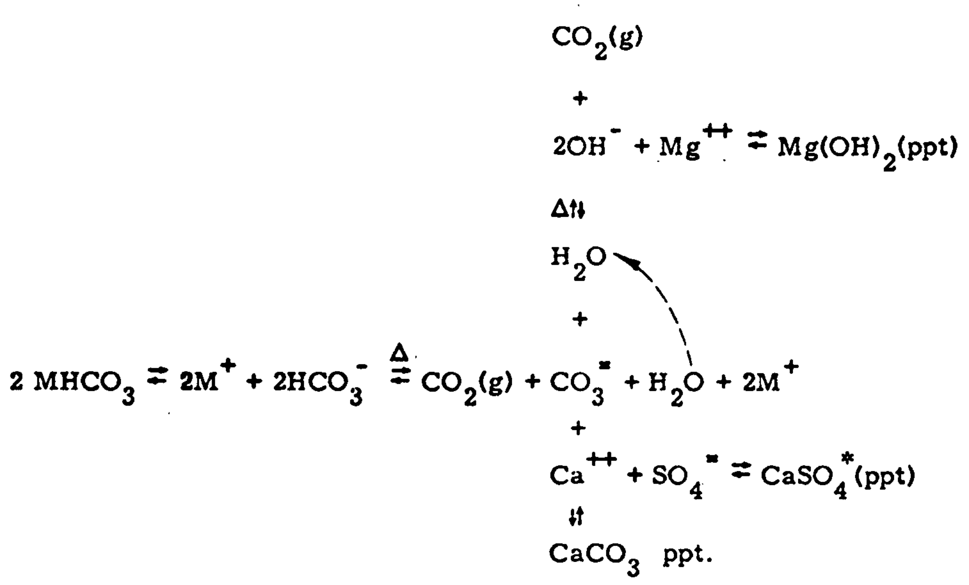
$Mg(OH)_2$, therefore, precipitates around 180°F and above. The interrelation among these reactions (Equations 5.1 through 5.6) is shown in Figure 5.1 (from OSW R&D Report Number 133, page 14) which shows that the various scale forming reactions are actually competing with one another.

The second type of scale, called the acid scale, is due to three forms of calcium sulfate — $CaSO_4$ or anhydrite, $CaSO_4 . 1/2\,H_2O$ or calcium sulfate hemihydrate, and $CaSO_4 . 2H_2O$ or gypsum. Hemihydrate is commercially known as plaster of Paris while gypsum is referred to as the dihydrate. The solubility product of calcium sulfate is given as:

$$\left[Ca^{++}\right]\left[SO_4^{--}\right] \rightleftharpoons K_{sp} = 10^{-5} \qquad (5.7)$$

which is seen to be much higher than that of either $CaCO_3$ or $Mg(OH)_2$. For practical engineering purposes solubility equilibria are most conveniently presented as diagrams here depicted as Figure 5.2, 5.3, and 5.4. The Langelier Index is commonly used to quickly estimate the scaling tendencies of saline waters. This is defined as the difference ($pH-pH_s$) between the actual measured pH of the water and the calculated pH_s at saturation with calcium carbonate or magnesium hydroxide. A positive index indicates tendency for scale deposition while negative values indicate unsaturation and hence tendency to dissolve away scale (corrosive tendencies).

While the precipitation of $CaCO_3$ and $Mg(OH)_2$ is mainly affected by CO_3^{--} concentration, pH, and temperature, the solubility of calcium sulfates is in addition affected by the concentration of other ions present. Figure 5.5 shows that the solubility of gypsum is 26% higher in magnesium chloride as compared to that in sodium chloride at an ionic strength of 2.0.

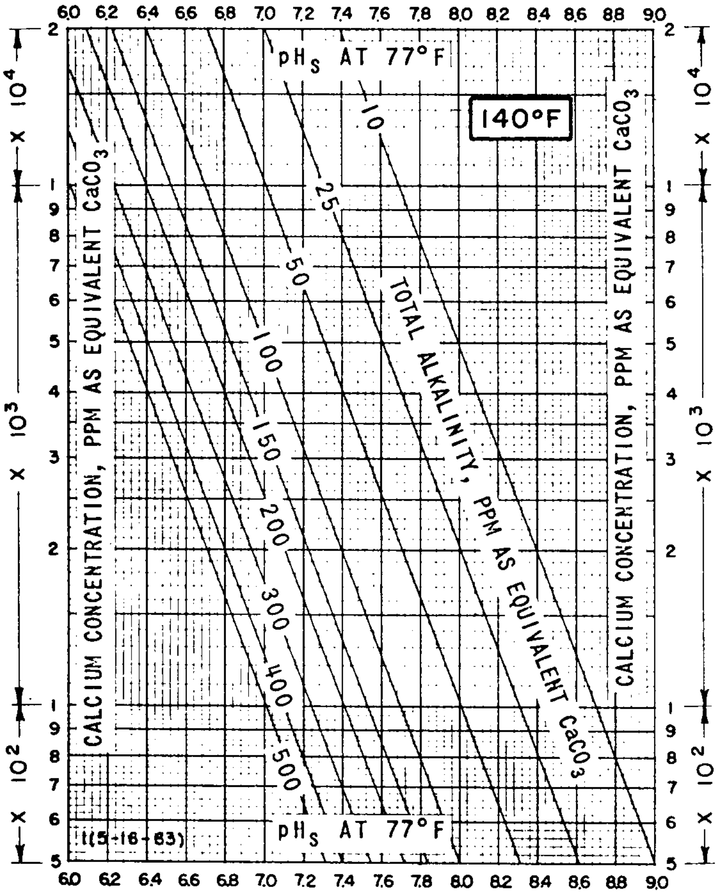


Note: "M" stands for any univalent positive ion which forms a soluble bicarbonate.

Note: In sea water at ambient temperature under equilibrium conditions, all dissolved CO_2 is primarily in the form of HCO_3^-. Molecular CO_2 and $CO_3^=$ exist only in trace quantities.

*Note: This reaction is independent of HCO_3^- breakdown and forms only as a result of sea water concentration.

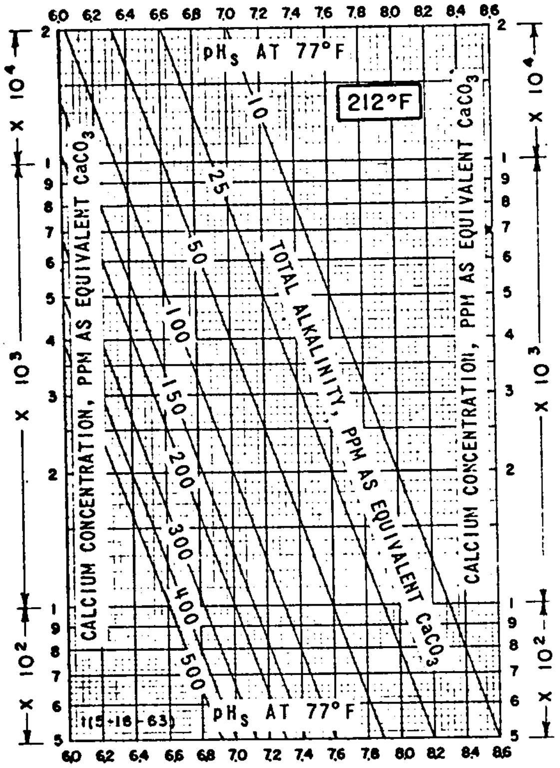
Figure 5.1. Composite Reaction Equations for Precipitation of Scale-forming Compounds from Seawater



NOTES:

1. SEE SHEET OSW 10.150 FOR A DISCUSSION ON THE USE OF THIS CHART.
2. THIS CHART IS DESIGNED FOR CONCENTRATED SEA WATER HAVING A TOTAL MINERAL CONTENT BETWEEN 25,000 AND 100,000 MG PER LITER. CHART MAY ALSO BE USED FOR SIMILAR SALINE WATERS (HARBOR WATERS, ETC.) PROVIDING THE TOTAL ALKALINITY IS DUE ONLY TO BICARBONATE.

Figure 5.2. Calcium Carbonate in Concentrated Seawater — Stability Diagram (140°F)



ILLUSTRATION

ASSUME: Ca = 2,000, TOTAL ALKALINITY = 100 (BOTH EXPRESSED AS ppm EQUIVALENT $CaCO_3$); pH (AT 77°F) = 8.6. FIND THE SATURATION pH AND SATURATION INDEX AT 212°F

SOLUTION: REFERRING TO THE 212°F CHART, LOCATE INTERSECTION OF Ca = 2,000 AND TOTAL ALKALINITY = 100. READ pH$_S$ = 7.0. SATURATION INDEX = 8.6 − 7.0 = 1.6.

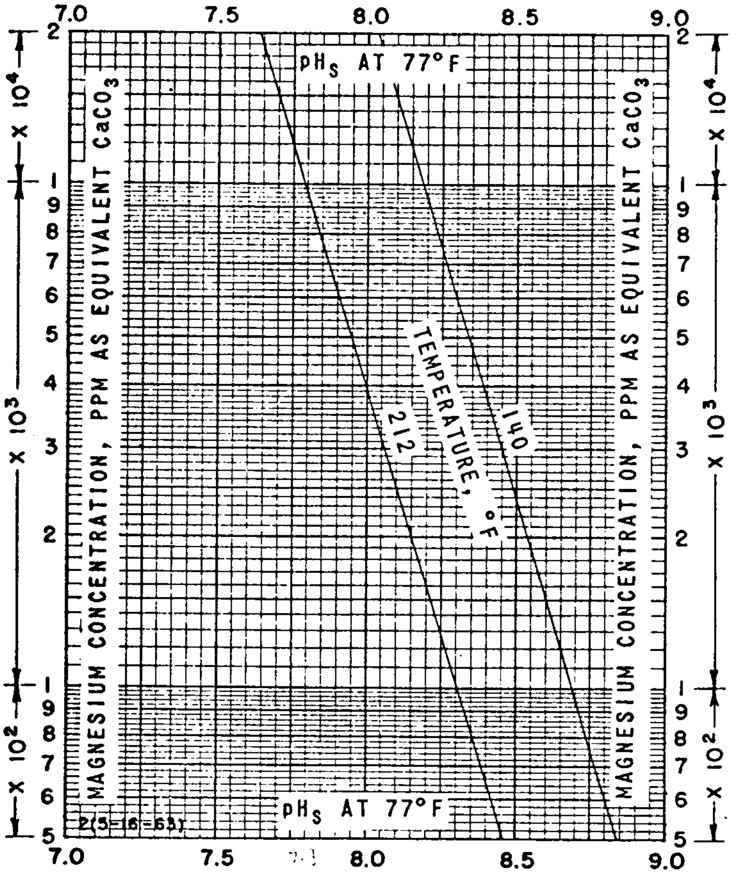
THIS CHART PREPARED BY
THE M.W. KELLOGG CO.
FOR THE OFFICE OF SALINE WATER
U.S. DEPT. OF THE INTERIOR
UNDER CONTRACT NO. 14-01-0001-291

REFERENCE:

Langelier, W.F., Caldwell, D.H., and Lawrence, W.B., *Causes and Preventives of Scale and Corrosion in Thermocompression Equipment when Employed for the Distillation of Sea Water*, Inst. Engg. Research Report, Univ. of California, Berkeley, Calif. (August 15, 1950).

TECHNICAL DATA BOOK
OFFICE OF SALINE WATER
CURVE No. OSW 11 151
APPROVED W.C.S. DATE 5-63

Figure 5.2 (cont'd). Calcium Carbonate in Concentrated Seawater — Stability Diagram (212°F)



NOTES:

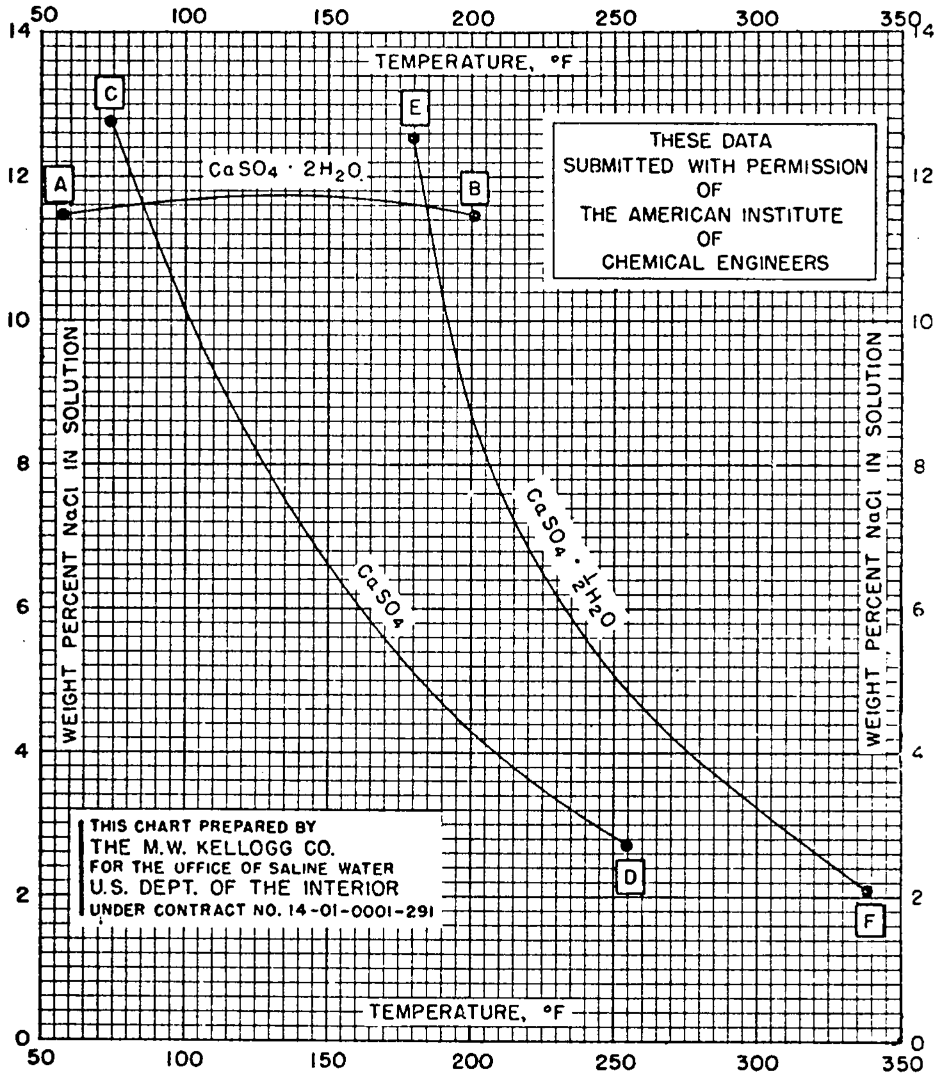
1. SEE SHEET OSW 10.150 FOR A DISCUSSION ON THE USE OF THIS CHART.
2. THIS CHART IS DESIGNED FOR CONCENTRATED SEA WATER HAVING A TOTAL MINERAL CONTENT BETWEEN 25,000 AND 100,000 MG PER LITER. CHART MAY ALSO BE USED FOR SIMILAR SALINE WATERS (HARBOR WATERS, ETC.) PROVIDING THE TOTAL ALKALINITY IS DUE ONLY TO BICARBONATE.

ILLUSTRATION

ASSUME: Mg = 10,000 PPM AS EQUIVALENT $CaCO_3$, AND pH (AT 77°F) = 8.6. FIND THE SATURATION pH AND SATURATION INDEX WITH RESPECT TO MAGNESIUM HYDROXIDE AT 212°F.

SOLUTION: LOCATE INTERSECTION OF Mg = 10,000 AND THE 212°F CURVE. READ pH_S = 7.8. SATURATION INDEX = 8.6 - 7.8 = 0.8.

Figure 5.3. Magnesium Hydroxide in Concentrated Seawater — Stability Diagram



NOTES:

1. SOLUBILITY LIMITS SHOWN ARE <u>APPROXIMATE</u>.
2. THE AREA ABOVE CURVE *AB* REPRESENTS SUPERSATURATION WITH RESPECT TO $CaSO_4 \cdot 2H_2O$.
3. THE AREA ABOVE CURVE *CD* REPRESENTS SUPERSATURATION WITH RESPECT TO $CaSO_4$.
4. THE AREA ABOVE CURVE *EF* REPRESENTS SUPERSATURATION WITH RESPECT TO $CaSO_4 \cdot \frac{1}{2}H_2O$.

Figure 5.4. Calcium Sulfate — Approximate Solubility Limits in Seawater Concentrates

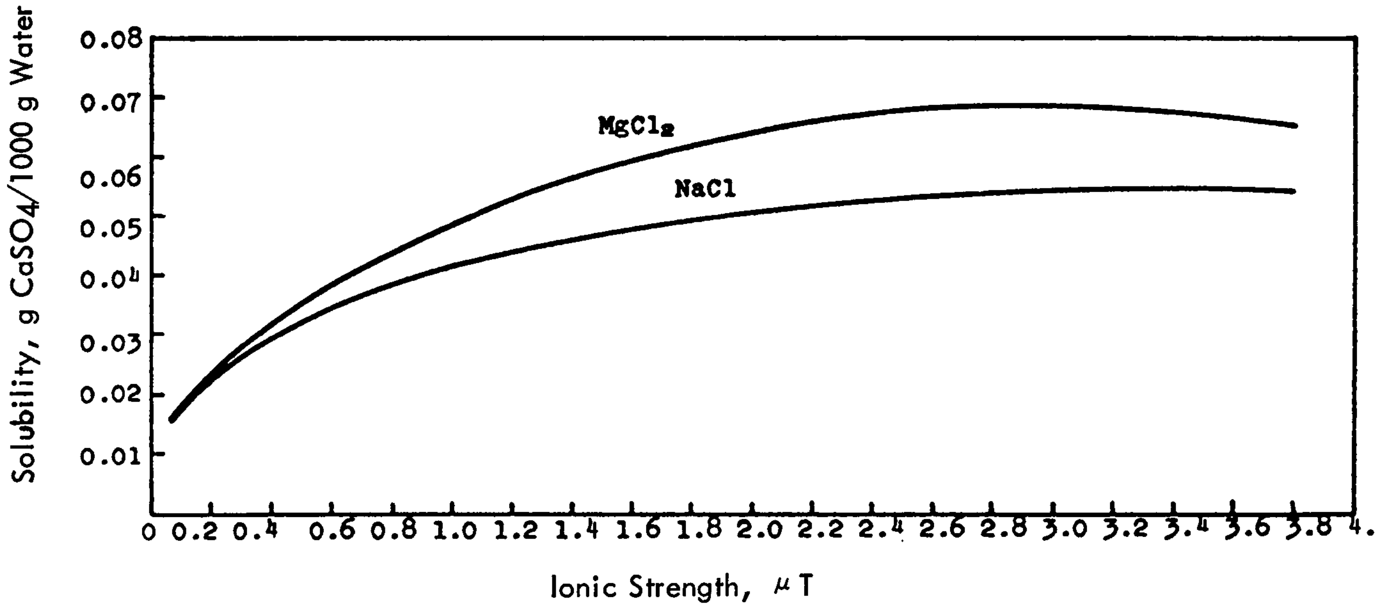


Figure 5.5. Effect of NaCl and $MgCl_2$ on the Solubility of Gypsum at 25°C

B. SCALE PREVENTION TECHNIQUES

Since scales are adherent and hard to remove it is best to plan and design a plant for scale prevention rather than for scale removal after formation. The scale prevention techniques fall into two broad categories corresponding to the two main types of scale:

- alkaline scale prevention techniques, and
- acid scale prevention techniques.

As seen in the previous section, the alkaline scale is due to the precipitation of $CaCO_3$ and $Mg(OH)_2$ which result mainly from the decomposition of the bicarbonate ion HCO_3. In distillation processes, such as were in operation at Freeport, seawater has to be heated to a temperature of 250°F to 280°F so that alkaline scale formation is quite possible. The following scale prevention techniques are available in this case:

- Hot lime-soda process.
- pH control using organic or inorganic acids, or salts of strong acids (e.g., ferric chloride).
- Seeding with fine crystals to cause precipitation away from the surface.
- Sequestrants, chelatants, complexing agents, or dispersants. Hagevap-LP and polyphosphates come under this category.
- High temperature heating to decompose bicarbonates and release CO_2 as a gas.
- Ion-exchange or reverse osmosis to either selectively remove bicarbonate ions or to eliminate all anions and cations.

Out of these various possibilities only three methods have been widely used on large scale due either to economic or technical reasons: the pH control, seeding techniques, and the polyphosphate (Hagevap-LP is a trade name for a polyphosphate-based formula) method.

1. The Polyphosphate Technique

During the early 1940's when wartime conditions caused the number and importance of seawater distillation plants to increase rapidly, the U.S. Navy introduced the "Navy Starch and Boiler Compound" treatment, which reduced or eliminated the earlier thermal shock treatment and the physical cleaning frequency. The exact behavior of this material was — and remains — something of a mystery. It consisted of equal parts of finely divided cornstarch and commercially available phosphate-based boiler treatment compound. The minerial deposits (scale) are rendered soft and spongy rather than hard and adherent so that they could be easily brushed off. After World War II a preparation called Hagevap-LP was developed and marketed, which appears to have been derived from the Navy Boiler Compound. This preparation contains sodium polyphosphates, lignin sulfonic acid derivatives, and various esters of polyalkylene glycols. A number of formulations are described by Liddell[1]. It is not exactly known how these compounds act but from the nature of the components involved it is postulated that the polyphosphates act as sequestrants for calcium and magnesium ions; lignin sulfonic acid, starch, tannin, etc., act as dispersants (i.e., coating surfaces such that scale adherence and crystal growth are inhibited), while the polyalkylene glycols are surface active agents which tend to retard foaming of the seawater. According to Liddell, these formulations act synergistically so that the total beneficial effect is greater than the additive effect of each acting alone. The striking fact about these polyphosphates is that generally 2 to 4 parts per million (ppm) are sufficient for their action. Chemical reactions with all the scaling ions are therefore obviously to be ruled out.

One limitation to the use of polyphosphates is the fact that at temperatures above about 190°F they undergo chemical change to orthophosphates, which are not effective as anti-scaling agents. It has also been noted that a few ppm of iron ions can "poison" these polyphosphates.

2. The pH Control Technique

Since alkaline scales are due to the presence of the bicarbonates of calcium and magnesium, they can be decomposed with any organic or inorganic acid, or even with salts like ferric chloride which form highly acidic aqueous solutions. For sulfuric acid, the overall chemical reaction is written as:

$$H_2SO_4 + 2HCO_3^- \rightleftharpoons 2H_2O + 2CO_2 + SO_4^{--}$$

so that when using 95% pure sulfuric acid on seawater containing 140 ppm HCO_3^-, 118.3 ppm of acid is required. When the CO_2 produced in this reaction is

---

(1) "Method of Treating Seawater" Hagan Corp., Pittsburgh, Pa. (by R. W. Liddell), U.S. Patent 2,782,162, February 19, 1957.

removed (along with deaeration) the reaction becomes irreversible and almost complete decomposition of HCO_3 can be secured. Sulfuric acid is objected to on two counts — first, because it produces about 10% more sulfate ions; and second because it is dangerous to handle and corrosion problems can occur. Experience has shown that both of these objections are not as serious as they appear at first sight. This is discussed in later sections.

An advantage of an acid such as sulfuric acid is that it not only prevents formation of scale but also can dissolve and remove alkaline scale already deposited on heat transfer or other surfaces.

3. The Seeding Technique

This technique has also been referred to as the "sludge recycle" method since theoretically the seeds can be recycled indefinitely. The basis for this technique is the fact that the presence of a large number of small particles in suspension throughout the whole body of solution enables the nucleation and crystal growth to occur on these particles rather than on the heat transfer or other surface where it is undesirable.

The seed particles are more effective if they are very fine (about 10 microns), obtained by precipitation in situ, and of the same or similar crystalline structure as the scale-forming substance. Even seeds of a foreign substance can be effective for scale deposition away from the equipment surface as in the "contact stabilization" method utilizing a bed of sand particles developed by Langelier and coworkers[1]. Once coated with the scaling substance ($CaCO_3$, $Mg(OH)_2$, etc.) from initial nucleation the seeds act like the crystals of the scaling component and facilitate further precipitation. Substances, added for internal water treatment, which form precipitates are effective as seeds. The principal application of foreign seeds has been in steam boilers and sugar evaporators. In the mining of sulfur where large quantitites of high temperature seawater are required, the use of alluvial mud in suspension has been used for scale prevention[2]. Impalpable silica has been used in British seawater evaporators[3].

---

(1) Langelier, W.F., Caldwell, D.H., and Lawrence, W.B., Ind. Eng. Chem. 42, 126 (1950). Also, J. Am Water Works Assoc. 46, 461 (1954).

(2) Axelrad, B.A., U.S. Patent Number 2,817,504, December 24, 1957.

(3) Jenkins, D.S., Howe, H.D., and Murphy, G.W., U.S. Department of Interior Saline Water Conversion Program, Inspection of Research and Development in Europe and N.W. Africa, Sept. 1954, of O.E.C. Working Party Number 8

Seeding technique has the advantages of practically no raw material cost and very little corrosion, if at all[1]. The limitations of this technique are that the seeds must be small in size and the recovery of fine seeds is difficult; the seeds must be kept in suspension or else they will settle out at points of low turbulence and block flow passages or if deposited on heat transfer surface, a baked-on coating may be produced; seeds of calcium sulfate hemihydrate and other similar substances may plaster the surface[1]. Experiences with alkaline scale prevention techniques at Freeport are given later in this section.

4. Acid Scale Prevention Techniques

Acid scales from seawater are due mainly to the precipitation of anydrite, hemihydrate, and dihydrate of calcium sulfate. There are four different approaches to preventing acid scale formation:

- Scale modification methods where the coherence and adherence of the scale is reduced so that hard deposit is not formed on the heat transfer surface. Starch, tannins, and polyphosphates come under this category.
- Chemical decomposition — with soda ash to produce calcium carbonate and sodium sulfate; precipitation with monammonium phosphate and sodium hydroxide or with triammonium phosphate to produce, ultimately, high analysis fertilizer (OSW R&D Report Numbers 66 and 57).
- Sludge recycle or seeding technique.
- Ion exchange, electrodialysis, Donnan softening, reverse osmosis, etc. Some of these techniques are discussed in the Pretreatment Methods Section (Chapter 5, G).

Only one of these various techniques for preventing acid scales were tried at the Freeport Plant. This was the polyphosphate technique. The main emphasis at Freeport appeared to be on designing the process such that the solubility limits of neither of the three calcium sulfates were exceeded. This is depicted

---

(1) Standiford, F.C., and Sinek, J.R., Chem. Eng. Progress, 57, Number 1, 58 (1961).

in Figure 5. 6. taken from a paper by Jebens and Dykstra[1]. The dotted line which represents the temperature concentration factor points in the 12-effect evaporator, is seen to lie to the left of, and below, the three solubility curves.

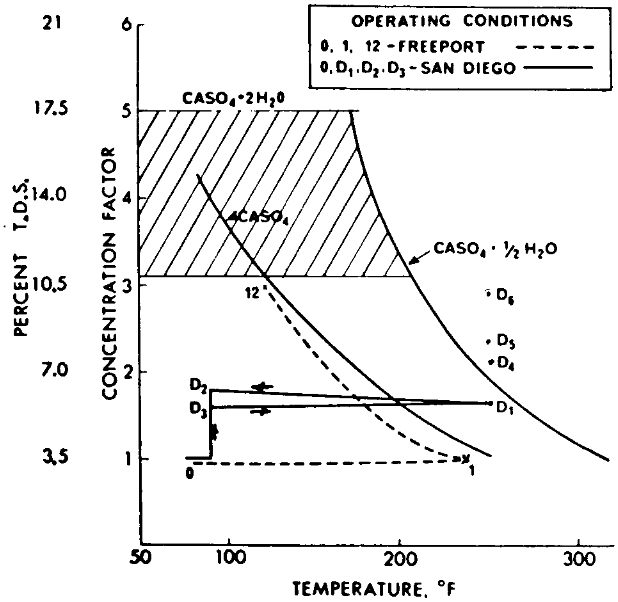


Figure 5.6. Approximate Solubility Limits of Calcium Sulfate

C. EFFECTIVENESS OF THE SCALE PREVENTION METHODS TRIED AT FREEPORT

To help better understand the later discussion on the effectiveness of the scale prevention methods tested at the Freeport Plant, it is deemed necessary to briefly discuss the background of the process and the preliminary work that led to the adoption of the first anti-scale technique.

The first work done by W. L. Badger Associates for the Office of Saline Water was a paper study to develop evaporation systems that would produce

---

(1) Jebens, R.H. and D.I. Dykstra, Chem. Eng. Prog. 61, Number 8, 68, (1965).

potable water from seawater on a large scale at the lowest possible cost[1]. The cost estimates were based on the assumption of using the simplest and cheapest type of evaporator that could be kept free of scale at little or no cost. A falling-film long-tube-vertical (LTV) evaporator was chosen because of its high heat transfer coefficients and certain other advantages. Seawater passes only once through the tubes of this evaporator. Twelve effects were suggested for this evaporator to operate over the range of 250°F to 120°F. Pilot plant studies were later conducted at Wrightsville Beach, N. C., to obtain heat transfer and scaling data with seawater. Two most promising methods selected for scale prevention tests were acid injection (pH control) and seeding. The acid used was sulfuric acid without inhibitors. The seeding tests used $CaCO_3$, $Mg(OH)_2$, and various crystal forms of $CaSO_4$ as seeds.

Not many runs were carried out with acid because the efficacy of acid had been proven in tests with thermocompression stills. Two runs during this period showed that even when equilibrium pH for $CaCO_3$ was exceeded by 0.3 to 0.4 units (positive Langelier Index) no scale formed at 192°F with an overall driving force of 8-10°F. However, scale did form when the pH was exceeded by 1.5 units. Somewhere between pH=8.3 to 7.5 lies the maximum permissible pH.

Initial series of runs with $CaCO_3$ seeds (obtained initially by adding soda ash to seawater) were quite successful. No scales formed over a week-long period. When temperature and seawater concentrations were raised to simulate backward feed evaporator $CaSO_4$ scale resulted that could not be prevented by the use of either $CaCO_3$ seeds or in combination with either of the two forms of $CaSO_4$. When $CaSO_4$ scale proved to be troublesome, the whole flowsheet of the commercial version of the process was changed to a forward feed arrangement, so that Effect I handled raw seawater (normal or low concentration) at the highest temperature while the XII Effect would have the most concentrated seawater but at the lowest temperature. A number of successful (scale-free) runs were made with $Mg(OH)_2$ slurry which seemed to prevent even $CaCO_3$ precipitation. A final summarizing run was made to test this scale prevention technique under the most severe conditions to be expected in a commercial version. Temperature for this run was 250°F and an LTV discharge concentration factor of 1.6. In the first 593 hours of operation, the concentration factor frequently went up to 1.8 and hemihydrate scale formed, although very slowly. This scale was removed by rinsing with warm seawater

---

(1) Badger and Standiford, Nat. Acad. Sci-Nat. Research Council Publication 568, 103 (1958)

and the run was continued another 1001 hours under such conditions that the concentration factor of 1.8 was reached only occasionally. No scale was found at the end of this run. Details on this pilot plant investigation are given in OSW R&D Progress Report Number 26. A 1% slurry of $Mg(OH)_2$ was therefore recommended for the Freeport Plant.

1. Seeding Technique at Freeport

The Freeport Plant was constructed by Chicago Bridge and Iron Company with only slight modifications to the original design submitted by W. L. Badger and Associates, Inc. The process flowsheet provided for a deaerator (to remove dissolved gases from seawater) the bottom section of which contained a slurry agitator to mix uniformly the recovered $Mg(OH)_2$ crystals. A clarifier and classifier were available to recover these crystals from the concentrated brine discharge (final effect blowdown). A 1% concentration of $Mg(OH)_2$ crystals was used in the seawater feed. Within two days of operation it was found that the feed control valve and the first effect were heavily scaled-up. Reason for the failure of the seed scale control technique was stated to be "contamination from dredging operations"[1]. This reference further comments that "subsequent pilot plant work showed that the cause of this scaling was a lower than expected solubility of magnesium hydroxide at the temperature of seed addition, thereby raising the pH enough to precipitate $CaCO_3$. The remedy was simple — add the seeds after preheating the seawater to a higher temperature. However, acidification proved to be so successful that it has been used ever since, and seeding has not been tried again." Since $Mg(OH)_2$ solubility decreases with an increase in temperature, it is hard to see how higher temperature of seed addition would have helped in scale prevention. However, since the seeding technique is successfully utilized in salt, sugar, and sulfite waste liquor recovery industries, it appears that this method may have been prematurely abandoned at Freeport.

2. Effectiveness of pH Scale Prevention Technique

As mentioned earlier, pH scale control technique involved addition of stoichiometric quantities of sulfuric acid to seawater feed before it was sprayed into the deaerator tower. Stripping steam was introduced at the bottom

---

(1) W. L. Badger & Associates, Inc., "Heat Transfer and Scale Prevention for Seawater in the LTV Evaporator" Report 487 to OSW (February 1964)

of the tower at a rate of approximately 1 pound per 1,000 pounds seawater feed and was taken from the XI Effect vapor head. This steam was essential for stripping away from the seawater feed the dissolved gases (O_2, N_2, CO_2, etc.) and the CO_2 generated in the decomposition of HCO_3 ions by the acid added. The residual alkalinity in the seawater feed after deaeration and decarbonation ranged from 7 ppm to 10 ppm. As long as the acid feed rate was correct and the acid mixing with the seawater feed was efficient, there were no problems with alkaline scale. However, acid scale, due to calcium sulfate, was still a continuing problem.

In June 1961, when Stearns-Roger's Operating and Maintenance crew took over the plant, an inspection was carried out of Effects I, II, III, IV, X and XII which revealed that seventeen tubes were plugged with scale in XIIth Effect and three in the Xth Effect. This indicated that the final concentration factor of 3.0 would have to be used instead of the design value of 4.0. Standiford and Bjork (op. cit.), of W. L. Badger Associates, explained that this situation occurred because of the use of $CaSO_4$ solubility data obtained in 1849 by Usiglio. Subsequent data by Tanaka, et al[1], indicated, according to Standiford and Bjork, that saturation with respect to gypsum is reached at lower concentration factors. The change to lower concentration factor at Freeport meant that more seawater had to be fed relative to the prime steam rate used, more feed had to be acidified, more heating surface was required in the preheaters, and the load on the brine transfer pumps would be greater. A slight advantage of this change was the reduced boiling point elevation losses from the overall plant temperature driving force.

In the First Annual Report by Stearns-Roger to OSW it was stated (page II-3 and II-13) that heavy scaling continued in Effects I and XII with deposits becoming progressively less towards the center effects. Unsuccessful attempts were made to remove scale by low pH operation. Several tubes had to be drilled out in March 1962. It was argued that this scale was due to high seawater pH relative to the equilibrium pH of 7.5. Operation with pH in the range 6.5 to 7.5 was recommended for future runs.

In April 1962, sodium hexametaphosphate was injected at the suction of the evaporator feed pumps P-4 and P-5 at a rate of one pound per hour. The anticipated effect of allowing the brine to stay supersaturated without scaling did not materialize, probably due to destruction of the phosphate activity at the high first effect temperature. Two additional factors affecting scale formation

---

(1) Tanaka, U., K. Nakamura, and R. Hara, J. Chem. Soc. Japan, Ind. Chem. Sect., 34, 779 (1931).

were detected during this period: the first effect steam temperature fluctuation and the poor distribution of seawater feed to the heating element tubes in the evaporator effects. Since the heat transfer rate is not significantly affected by the liquor rate, a decrease in this rate due to poor distribution leads to overconcentration and ultimately to scaling. As discussed earlier in Chapter 4, Heat Transfer History, a significant amount of development effort was devoted at Freeport to find solutions to this problem of liquid distribution. Both the spray nozzles and the SVL weir were found to be superior to the original distribution plate.

Once the alkaline scale problem was solved, efforts were directed to find solutions to acid scale (due to $CaSO_4$ crystal modifications), and to establish maximum temperature possible in the first effect without scaling. Studies were continued up to the last Development Run 16 to optimize the operation of the deaerator-decarbonator at the Freeport Plant. The various annual reports (1 through 8) and other documents submitted by Stearns-Roger to OSW contained the following conclusions:

- Acidification of seawater to a pH = 4.0 at inlet to deaerator was adequate for effective prevention of alkaline scales.
- Seawater feed entering the deaerator should flash down by 0.5^{o} to $1^{o}F$ for optimum CO_2 stripping.
- Stripping steam rate should be about 1 pound per 1000 pounds of normal seawater feed.
- Although the deaerator effluent alkalinity ranged from 7 to 10 ppm $CaCO_3$, it can go up to 30 ppm without scaling in the first effect.
- When polyphosphate (Nalco 918 at Freeport) is used for $CaSO_4$ scale prevention it is effective if the maximum temperature at and subsequent to the point of injection is limited to $170^{o}F$. About 3 to 5 ppm concentration in seawater is adequate for scale prevention.
- Incipient scaling point in Effect I appears to lie between 268 and $273^{o}F$ with blended seawater at Freeport (concentration factor = 1.0).

When scale does form on equipment surfaces, the effect of the rate of formation on the heat transfer coefficient U and method of removal must be considered.

D. VARIATION IN U WITH SCALE CONTROL METHODS

Since water desalination economics is closely tied in with the rates of heat transfer, any variation in U, (overall heat transfer coefficient) due either to scaling or other reasons, is bound to have repercussions. The direction of this influence is known — any scale formation will tend to reduce the rate of heat transfer, or increase the temperature driving force across the heat transfer surface for the same initial rate. If conditions are not changed, scale will continue to build up until the tube is plugged. Of course, the rate of scale deposition generally decreases as the deposit increases in thickness. No specific data were taken at Freeport for the variation of U with the scale control methods (seeding, pH, and polyphosphate) applied. However, the following brief theoretical treatment explains what variations in U can be expected with a particular scale control method.

1. Theoretical Effect of Scale Control Methods on U

The three methods of scale control applied at Freeport were:

- Seeding or sludge recycle method.
- Acidification or pH method.
- Polyphosphate method.

The sludge recycle method was abandoned after only two days of operation at Freeport, so it will not be considered further. The acification of seawater feed with sulfuric acid method is meant for the control of alkaline scale, and depends for its efficacy on the addition of the correct amount of acid, thorough mixing with the seawater feed, and on the proper design and functioning of the deaerator to desorb CO_2. If any of these steps are inefficient, alkaline scale ($CaCO_3$, $Mg(OH)_2$) will form on equipment and heat transfer surfaces. The acid scale results from overconcentration. The extent of reduction in U will depend on the thickness and extent (area blanketed) of scaling, porous or dense nature of deposit, the change in temperature and temperature driving force ΔT, and any change in flow pattern or velocity due to the nature of the deposit (smooth or rough) and the reduced flow cross-sectional area. It is interesting to consider the chain of events triggered by scale formation in any given effect in a multiple effect evaporator. For example, consider scale formation in the heating element tubes of Effect VI. Since scale is a poor heat conductor, the heat transfer through the tubes will drop. Less steam, therefore, is condensed in the steam chest of this effect while Effect V continues to discharge vapors at its initial rate. Pressure builds in the steam chest of Effect VI. The condensing temperature of this steam goes up. On the tube side, less vapor is produced so the evaporation temperature is decreased. Overall ΔT across this effect, therefore, goes up. On the steam side (outside of tubes) the individual film coefficient may tend to increase due to increased steam condensation temperature, while the brineside coefficient may decrease; thus the overall coefficient is subject to opposing influences. Disregarding all these other side effects, the direct reduction in U from scale formation can be calculated from the following equation:

$$U = \frac{1}{\frac{1}{h_b} + \frac{1}{h_c} + \frac{1}{h_d} + \frac{x_s}{k_s}} \qquad (5.8)$$

where h_b, h_c, h_d are the heat transfer coefficients for the steam side, tube wall, and brine film respectively (see Figure 4.1); the thickness of the scale is given as x_s and its heat conductivity by k_s. Since x_s is a function of time, U also varies with time. Equation 5.8 is for thin depsoits only, otherwise the area ratios will have to be included in this equation. Two cases of scale formation will be considered to calculate roughly the effect of scale formation on U: a uniform deposit of 1/16 inch thickness of $Mg(OH)_2$ and of $CaSO_4$ (anhydrite) on the inside surface of an aluminum brass tube 1.87 inch I.D. Equation 5.8 can be rewritten as:

$$\frac{1}{U} = \frac{1}{h_b} + \frac{1}{h_c} + \frac{1}{h_d} + \frac{x_s}{k_s} \qquad (5.9)$$

Now $\frac{1}{U}$ is thermal resistance per unit area under scaled conditions and if U_o represents the overall coefficient under initial clean condition, then Equation 5.9 can be rewritten as:

$$\frac{1}{U} = \frac{1}{U_o} + \frac{x_s}{k_s} \qquad (5.10)$$

so that from this equation it is possible to calculate the effect of scaling on U.

Assume that under clean initial condition U_o = 500 Btu/hr ft^2 ^oF. If the brine-side temperature is taken as 200^{o}F, then the thermal conductivities of $Mg(OH)_2$ and $CaSO_4$ (anhydrite) are 0.32 and 2.06 in Btu/hr ft ^oF units [1,2]. Substituting these values in Equation 5.10, it is seen that a 1/16 inch thick scale of $Mg(OH)_2$ reduces the overall coefficient by 89.1% (from 500 to 54.7) while a similar scale of anhydrite causes the coefficient to drop by 55.8% (from 500 to 220.8 Btu/hr ft^2 ^oF). It is to be noted that a given thickness of scale will cause a greater percentage drop of the coefficient in the low temperature effects (where U is smaller) than in the high temperature effects. Also, as the scale thickness grows the overall coefficient will keep continually decreasing until the tube is completely plugged when brine flow will stop and also the heat flux.

E. SCALE DEPOSITION RATE

Scale deposition on heat transfer surfaces may be uniform all over the surface or it may be severe only on certain parts of the surface. Both types of scaling patterns were observed at the Freeport Plant. The importance of the knowledge of scaling rate arises from the fact that it helps in planning for optimum plant operation period before shutdown for cleaning. Semi-theoretical treatment of scaling rates is possible only for uniform pattern of scaling on heat transfer surfaces.

All conditions essential to scaling are obviously essential to scale deposition rates. For example, there must be supersaturation with respect to the scaling component. In addition, since continual scale deposition involves crystal growth, the rate of scale deposition is subject to the laws of crystallization and mass transfer. If scale inhibitors like polyphosphates are added, the scaling rates will be affected. The supersaturation arises mainly from evaporation of water from the brine. Therefore, the quantity of heat transferred across the tube surface determines the degree of supersaturation of brine. Based on this last assumption, the following equation has been suggested to correlate the rate of heat transfer with the rate of scale deposition:

$$\frac{1}{U^2} = \frac{1}{U_o^2} + C\Theta \qquad (5.11)$$

---

(1) Perry, et al, Chemical Engineers' Handbook, Fourth Edition, pages 3-211 (1963)

(2) Partridge, E.P. and A.H. White, Ind. Eng. Chem. 21, 839 (1929)

where U and U_0 represent the overall coefficient under scaled and clean conditions respectively, while Θ is the time elapsed in days (starting from initially clean surface) and C is a constant. This equation states that when the change in U is due to scale deposition alone, then a plot of $1/U^2$ against time Θ should be a straight line. Heat transfer data were continuously monitored for two Effects, XI and XII, to obtain apparent U as a function of time. Figures 5.7 and 5.8 depict the observed and calculated data[1]. From these plots it is easily seen that, within reasonable limits of error, the linearity of $1/U^2$ values against Θ is borne out in both the cases (Effects XI and XII). The discontinuities in the curves in these figures were explained[1] as being due to a shutdown necessitated by a pump failure. During the startup operation the effects were exposed to normal seawater so that there was a slight dissolution of the previous scale and the U values went up. The second part of the curve, however, again shows linearity of $1/U^2$ versus Θ. The rate of scaling is indicated by the slope of the $1/U^2$ versus Θ curve. The higher the slope, the greater is the rate of scaling. Theoretically only two points would suffice to establishe the slope of this line and hence the scaling rate. It is, however, better to use four or five points to obtain the scaling rate with greater confidence and accuracy.

As mentioned earlier, this scaling rate correlation is also useful to determine the optimum time of plant operation for maximum capacity between periods of scale removal.

A scheme for quantitative physical measurement of scale was devised and used to measure the amount of scale during each outage. Dividing this amount by the number of days during which scaling occurred gave the average rate of scaling. Figure 5.9 shows the type of gage used to measure the extent of scaling. There were 4 probes of diameters ranging from 1 inch to 1 3/4 inch in 1/4 inch steps. Only alternate rows of tubes were gaged and the results doubled to obtain the total scale deposited. The method is not very accurate, therefore, but is a compromise to reduce labor and downtime to a reasonable level while serving to affix a quantitative value to scaling. For tubes equipped with V-notch weirs, the first gage dropped is 1 1/2 inch in diameter since the I.D. of the weir is less than 1 3/4 inch.

A tube not allowing the 1 inch gage to pass was considered plugged below that point. This was verified by visual inspection[2]. All scale was assumed to be gypsum and its density was taken as 0.0838 lb/in^3. The average thickness

---

(1) Third Annual Report by Stearns-Roger Corp. to OSW, page 154 (1964).

(2) Development Report Number 1, Stearns-Roger Corp., page 11 (1964). This report contains considerable data on scaling pattern and tube gaging results.

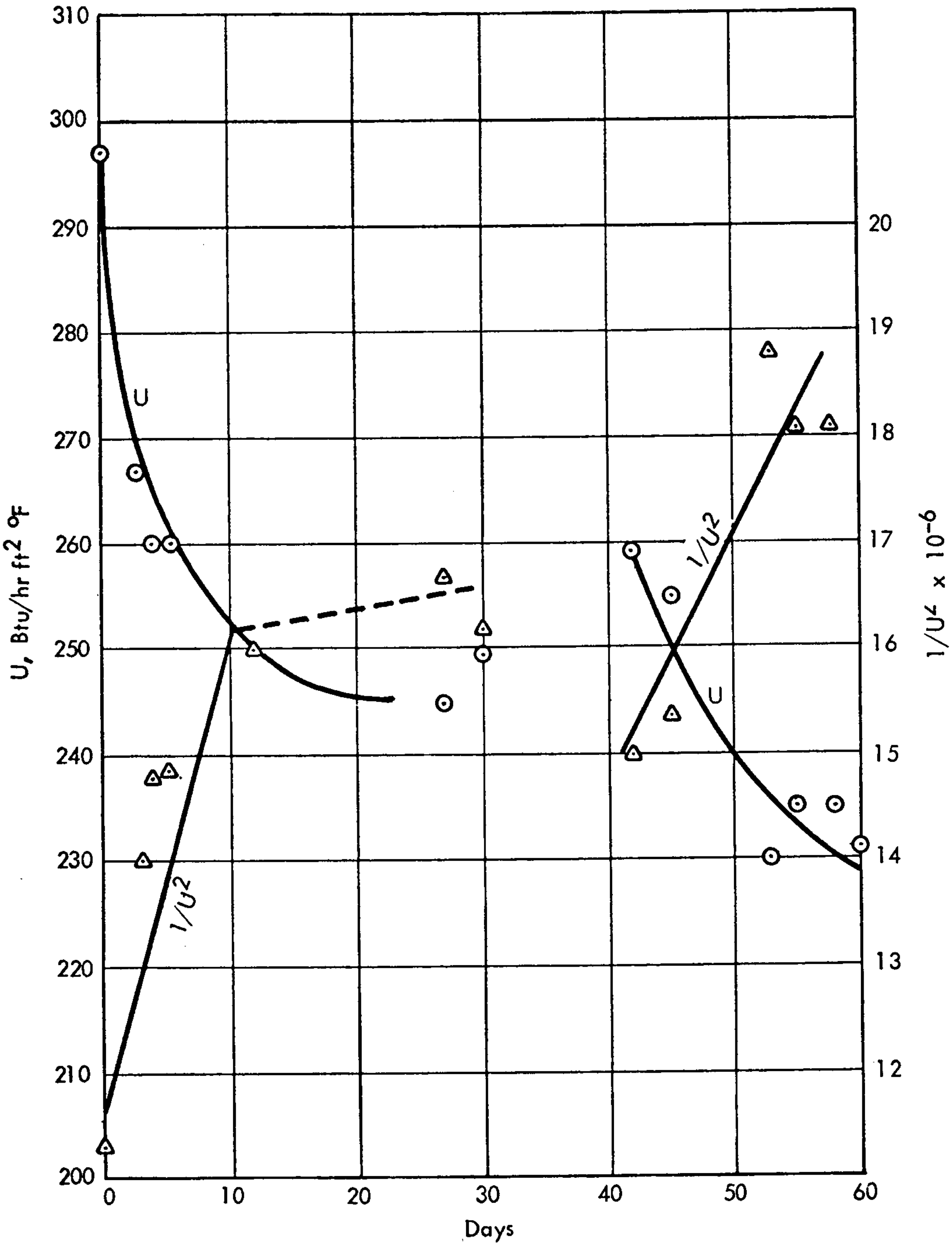


Figure 5.7. Scaling versus Apparent U Coefficient, Evaporator XI.

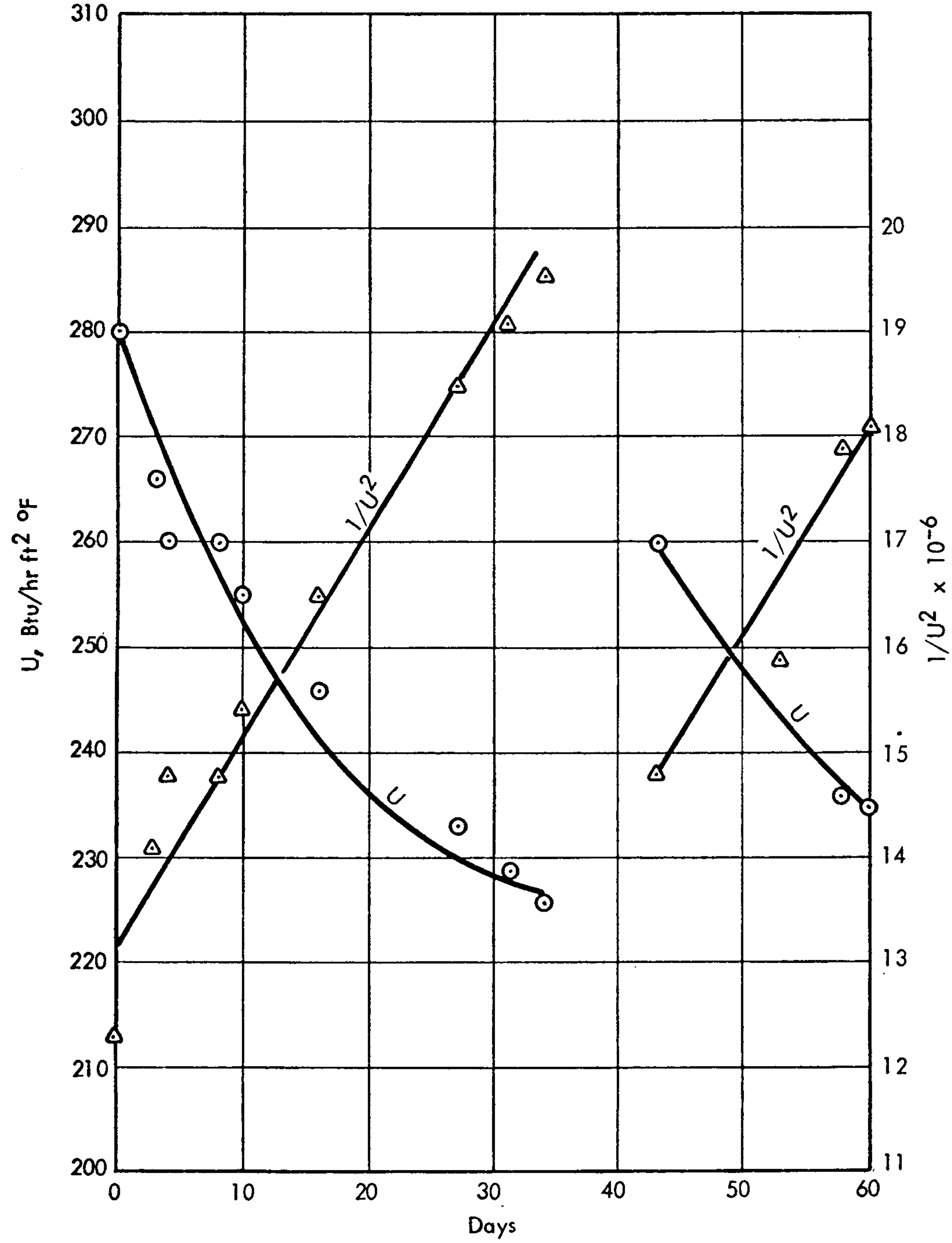


Figure 5.8. Scaling versus Apparent U Coefficient, Evaporator XII.

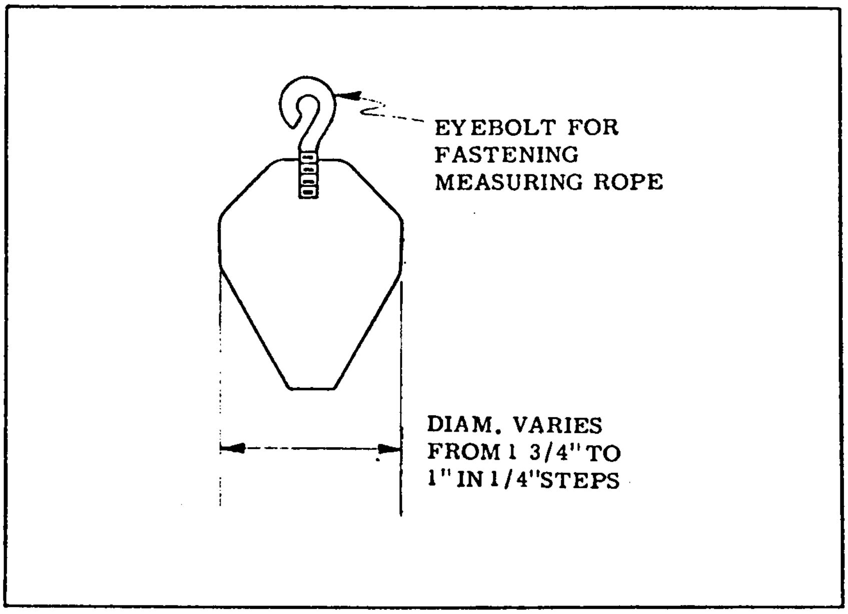


Figure 5.9. Scale Measuring Gage

of scale between stop-point of 1 1/2 inch gage and 1 inch gage was assumed to be 0.25 inch and similarly for other gages. As an example, Table 5.3 shows the scaling rate data for Effects XI and XII during two runs in 1964. Effect XI was purposely not provided with liquid distribution weirs as a control on the results obtained in XIIth Effect (equipped with weirs). Scaling is seen to be much severer in XIth Effect (equipped with weirs). Scaling is seen to be much severer in Effect XI compared to Effect XII. The results for Effect XI were interpreted as indicating that the rate of scale formation is a function of the amount of scale already formed. The scale already formed inhibits the transfer of heat and, therefore, less evaporation and concentration occur and less scale forms. This is the reason behind rapid initial decline of U in Figures 5.6 and 5.7 and then the gradual tapering off to an almost steady value.

Table 5.3

Scaling in Effects XI and XII

RUN PERIOD	OPERATING DAYS	TOTAL SCALE FORMED (LBS) XI	XII	DAILY SCALE FORMATION RATE LBS/DAY XI	XII	TUBES PLUGGED AT END OF RUN XI	XII
10-17 to 1-17	91.42	12,103	5,019	132.4	54.9	144	78
2-6 to 3-6	28.96	5,343	139	184.5	4.8	126	2

F. SCALE REMOVAL TECHNIQUES

Even if the process parameters are designed not to cross the solubility curves of the scale-forming compounds, it was noticed at the Freeport Plant that scale deposition can still take place due to liquid maldistribution and tube starvation. Process upsets can also lead to scaling. It is therefore essential to study the methods available for scale removal. Two broad divisions of these methods can be made: plant shutdown and subsequent scale removal, or on-stream scale removal methods. Obviously, the first method is undesirable but it has to be resorted to when other methods fail to work.

1. Scale Removal after Plant Shutdown

Severe scaling or plugged tubes require the plant to be shut down and equipment opened to remove the hard scale deposit by drilling. The tube may have to be rejected if the drilling operation does not work.

A second method tried for scale removal at Freeport involved the use of high pressure water jets. The Third Annual Report by Stearns-Roger Corporation states (page 28) that a pressure of 3000 - 4000 psi was involved, whereas the Fifth Annual Report by Stearns-Roger indicates (page 40) use of water up to 10,000 psi to remove anhydrite scale from Effect I. This technique consisted of passing a hose with nozzle through the entire length of each scaled tube. Scale in 147 of the 163 tubes scaled was removed in 14 hours. The method was said to be expensive but otherwise the best known until then.

A third method involved use of cold seawater (concentration factor close to 1.0) to dissolve out gypsum scale. This resulted in savings of up to 100 manhours per shutdown. For satisfactory results by this method, the washing must be initiated before the scaled tubes become completely plugged (Third Annual Report, page 29).

Calcium sulfate anhydrite is much less soluble than gypsum so that it is hard to dissolve out the anhydrite scale. Fourth Annual Report by Stearns-Roger contains Tables 3-2 and 3-3 (reproduced here as Tables 5.4 and 5.5) which report the results of rates of dissolution of anhydrite that were achieved by various methods. Nalco 918 is a polyphosphate preparation by Nalco Company while Versene is ethylenediamine tetraacetic acid (EDTA) — a chelating agent. The report concludes that circulation of seawater at 140°F is the most successful method attempted so far.

Table 5.4

Scale Deposited During 10 Day Period of Operation @290°F Inlet Steam

INCREMENT	TOTAL LIN FT OF SCALE	HOURS WASHED	DISSOLUTION RATE FT/HR		METHOD OF WASHING
			INCR	TOTAL	
0	2313	0	--	--	--
1	1606	36	19.64	19.64	120° F Sea water
2	1181	67	13.71	16.90	120° F Sea water + 5 PPM Nalco 918
3	1124	91	2.38	13.07	70° F Sea water
4	1090	111	1.70	11.02	70° F Sea water + Versene
5	940	135	6.25	10.17	140° F Sea water and shocking
6	704	153	13.11	10.52	140° F Sea water and shocking
7	493	173	10.55	10.52	140° F Sea water and shocking
8	380	195	5.14	9.91	10% NaCl solution @ 90° F
9	135	243	5.10	8.96	10% NaCl solution @ 130° F
10	12	263	6.15	8.75	10% NaCl solution @ 130° F
11	0	284	0.57	8.14	10% NaCl solution @ 130° F

Table 5.5

Scale Deposited During 16 Day Period of Operation @285°F Inlet Steam

INCREMENT	TOTAL LIN FT OF SCALE	HOURS WASHED	DISSOLUTION RATE FT/HR		METHOD OF WASHING
			INCR	TOTAL	
0	2521	0	--	--	--
1	2230	20	14.55	14.55	10% NaCl solution @ 130° F
2	1905	58	8.55	10.62	10% NaCl solution @ 130° F
3	1676	102	5.20	8.28	10% NaCl solution @ 130° F
4	1181	146	11.25	9.18	10% NaCl solution @ 130° F with improved water distribution
5	689	189	11.44	9.69	10% NaCl solution @ 130° F with improved water distribution
6	411	257	4.09	8.21	10% NaCl solution @ 130° F with improved water distribution
7	25	298	9.41	8.38	10% NaCl solution @ 130° F with improved water distribution
8	0	322	1.04	7.83	10% NaCl solution @ 130° F with improved water distribution

2. On-Stream Scale Removal

Scale removal with the plant operations shut down is expensive and inefficient since for continuous water supply the product storage capacity has to go up. Also, there is considerable labor involved in start-up and shutdown so that if scale can be removed while the plant is on-stream, it would be a very useful technique.

It was mentioned earlier that alkaline scale due to deposition of $CaCO_3$ or $Mg(OH)_2$ can be removed by lowering the pH of seawater such that the Langelier Index is negative. The danger of equipment corrosion, however exists with this method. This method was successfully applied at the San Diego, California MSF Plant of the Office of Saline Water (OSW R&D Report Number 133, page 100).

For acid scale due to calcium sulfate anhydrite, gypsum or hemihydrate, the acidification (pH lowering) method does not work. The solubility of these compounds is strongly dependent on temperature and/or concentration factor. By lowering the temperature and/or the concentration factor of the brine, operating conditions are shifted to points lying below or to the left of the solubility curves (see Figure 5.4). The deposited salt has then a tendency to go into solution, the rate being dependent primarily on the fluid turbulence and the degree of unsaturation of the circulating brine.

This method of scale removal is particularly suited for freshly deposited scales that are not too dense and/or thick. Scales resulting from process upsets can also be removed by this method very efficiently. Although there is some loss in production due to decreased concentration factor the net overall loss is much less than that encountered in off-stream descaling. This technique was, therefore, applied a number of times at the Freeport Plant (e.g., see page II-20, Annual Report Number 2; page 10, item 2-21, and page 109 in Annual Report Number 3; page 88 of Annual Report Number 8) with good results.

G. PRETREATMENT METHODS

Internal scale prevention techniques, where the scale preventive action is brought about while the liquid stream is flowing in the process equipment, were discussed in Section B, this chapter. A brief summary is presented here of the methods which involve treatment of the liquid (seawater feed stream) before it enters the main processing equipment.

The basic idea underlying these methods is to remove (or exchange) either partially or completely the scale forming ions from the seawater feed so that its scaling potential is significantly reduced or even entirely eliminated. It was seen earlier that two types of scale generally occur in seawater conversion plants: alkaline scale due to the bicarbonate ion HCO_3^- and acid scale due to Ca^{++} and SO_4^{--} ions. The following methods have been suggested and discussed in the literature:

- Chemical methods — Cold and hot lime soda process; pH adjustment with acids, salts of strong acids, e.g., $FeCl_3$; complexation or chelation with EDTA, etc.
- Membrane processes — Microfiltration or reverse osmosis, electrodialysis, Donnan Softening technique.
- Ion exchange techniques — Sodium and hydrogen zeolite softening, chloride-anion exchange technique, organic anion/cation exchange techniques.

- Physical methods — Adsorption, foam-fractionation, heating to a high temperature followed by filtration, etc.

Cold or hot lime soda process is effective for removal of carbonate hardness only. Acidification or pH adjustment method was used at the Freeport Plant for preventing and removing alkaline scales. Handling of acid and corrosion problems exist with this method. Ships and small facilities may prefer the use of $FeCl_3$ for this reason (OSW R&D Report Number 133, page 44). Use of chelants such as EDTA was shown to be very expensive (OSW R&D Report Number 42, also R&D Number 25).

Membrane processes are quite efficient in removing the scale-forming ions but are suitable at present only for brackish water. The Donnan Softening technique is discussed in OSW R&D Report Number 506. A number of process design examples were developed in this report from which costs of water softening were estimated. For Webster, S. D., brackish water (less than 1500 ppm TDS and hardness about 800 ppm as $CaCO_3$), the cost of pretreatment was stated to be 3.66 cents per 1000 gallons for 50 percent removal of calcium and magnesium ions.

With the introduction of weak acid/base resins, the continuous processes ion exchange is becoming a well established technique for large scale demineralization of brackish and seawaters[1]. A paper[2] presented recently at an OSW-sponsored symposium mentions the use of ion exchange technique by a 210 million gallons per day plant of the Metroplitan Water District of Los Angeles. Alameda County Water District of California was also stated to be using continuous ion exchange units to soften 25 million gallons seawater per day.

Finally, a number of physical methods of seawater pretreatment to prevent scaling have been proposed. Foam fractionation is one of the most recent techniques proposed for salt removal from seawater[3]. Another method

---

(1) OSW R&D Report Number 616 (1971) by Control Systems Research, Inc.

(2) Higgins, I. R., and R. C. Chopra, "Chem-Seps Continuous Countercurrent Ion Exchange for Removal of Calcium from Sea Water to Prevent Scaling of Distillation Equipment", paper presented at OSW Symposium on Treatment of Feed Water for Distillation Processes, Washington, D. C. (April 1971)

(3) OSW R&D Report Number 248

suggested in a patent[1] involves heating seawater by steam or flue gases to a temperature such that supersaturation with respect to the scale-forming component is reached and causing this component to precipitate partly by itself and partly on a suspension of fine particles.

A comprehensive technical-economic study of the pretreatment methods would be most helpful in optimizing the best combinations of pretreatment methods and desalination techniques.

---

(1) Mayfield, A. L., and R. J. Staffa, "Method and Apparatus for Treating Seawater to Remove Objectionable Hardness", U. S. Patent Number 3,026,261 March 20, 1962.

Chapter 6

MATERIALS AND CORROSION HISTORY

It is a well-established fact that marine environments are extremely corrosive. In a desalination plant handling seawater, as at the Freeport Plant, almost all equipment is exposed either internally and/or externally to hot seawater and a saline atmosphere. Since economic water costs can only be achieved by smooth plant operation with minimum downtime (due to corrosion failures, etc.) and by utilizing cheap materials of construction, the study of materials and their corrosion history was one of the prime goals of the Freeport Saline Water Conversion Plant program. The relevant experiences at this plant over its eight-year period of operation are discussed in this section under the following headings:

A. General Materials Philosophy

B. Application of Corrosion Test Data to Materials Selection

C. Summary of Plant Corrosion History

D. Most Critical Areas of Corrosion

E. Effect of Deaeration and pH Control on Corrosion Rates

F. Effectiveness of Cladding and Protective Coatings

G. Effectiveness of Cathodic Protection System

A. GENERAL MATERIALS PHILOSOPHY

To reduce unit water costs it is essential to reduce the initial capital costs and the fixed charges. If cheap materials of construction are utilized, capital costs can be significantly reduced; but the operating and fixed charges may go up due to frequent material failures, the resulting downtime, costly repairs, and reduced plant life. Selection of the materials of construction has, therefore, to be based on a number of factors, the principal ones being the corrosion resistance and the initial cost.

1. Basic Aspects of Corrosion

For structural strength and easy workability (including repairs) most of the equipment in any plant is made from metals and alloys. The thermodynamically more stable state of these metals is in the form in which these metals occur in nature as their ores, e.g., as oxides, carbonates, sulfates, etc. It is obvious, then, as to why metals exposed to corrosive atmospheres tend to return to their original stable state. This deterioration of metals and alloys in contact with hostile environment is known as corrosion. Corrosion of metals and alloys is undesirable since the product has no structural strength, it is not heat conductive, the metal strength is reduced, and finally the equipment may fail due to reduced mechanical strength. Corrosion is undesirable from several other standpoints too — corrosion products increase fluid flow resistance and hence aggravate pumping power requirements; corrosion products block flow passages and generate faulty process control signals so as to upset proper process regulation; accumulation of corrosion products on normally non-corroding metal surface initiates crevice or pitting type corrosion; liquid streams become polluted due to corrosion products; spillage and product loss, and finally danger to safety of operating personnel exist when corrosion occurs. There are, therefore, several reasons, both technical and economic, why corrosion should be avoided.

Corrosion can be classified in a number of different ways. Nature of corrodent determines the "wet" or "dry" type corrosion. If a metal is corroding due to chemical reaction with gases it is dry corrosion, while moisture-laden gases and corrosive liquids cause wet type of corrosion. A second classification is based on the type of mechanism involved in corrosion — either an electrochemical or a direct chemical reaction mechanism. A third classification of corrosion is based on the appearance of the corroded metal — uniform corrosion over the entire corroding surface or localized corrosion of small areas on the surface. Localized corrosion is further subdivided into two classes — macroscopic and microscopic. The former involves galvanic, erosion, crevice, pitting, exfoliation, and selective leaching types of corrosion while the microscopic localized corrosion involves intergranular and stress corrosion cracking.

a. Uniform corrosion — This can arise from chemical or electrochemical action. Iron and its various forms (mild steel, cast iron, etc.) readily undergo the following oxidation reaction in contact with air or seawater containing dissolved oxygen:

$$Fe + O_2 \longrightarrow Fe_2O_3$$

$$Fe + O_2 \longrightarrow Fe_3O_4 \text{ (magnetic oxide)}$$

In the presence of an electrically conducting fluid or electrolyte such as seawater plus the heterogeneity of metallic surfaces arising from inclusions, microstructural composition differences, localized stresses and mechanical imperfections, the metal surface develops minute anodic and cathodic areas which shift constantly so that a uniform electrochemical corrosion results. General corrosion probably causes the greatest amount of damage to metals. This type of corrosion has been well researched and documented so that equipment life can generally be estimated accurately. By proper selection of resistant metals, coatings, or linings, this form of corrosion can be reduced or even eliminated. Control of environmental conditions such as temperature, pressure, velocity, composition, etc., should be considered in the case of general corrosion. Finally, use of corrosion inhibitors may solve some immediate and long-range problems. With saline water conversion plants, where the product is water for human consumption, one is restricted in the employment of such inhibitors for corrosion because of the danger of product water contamination.

b. Localized corrosion — As mentioned earlier two main subdivisions of localized corrosion are the macroscopic corrosion and the microscopic corrosion. Galvanic, erosion, crevice, pitting, exfoliation, and selective or parting corrosion are examples of macroscopic corrosion. Galvanic corrosion takes place when two different metals in contact are exposed to a conductive corrodent solution. A difference in electrical potential exists between different metals, and forces an electrical current to pass through the solution. This current flow causes the anodic metal to corrode. Obviously, the greater the potential difference between the two metals the greater the rate of corrosion. In galvanic corrosion, the active metal corrodes more while the cathodic or noble metal corrodes less, relative to the corrosion rates when each is left by itself in the solution. For a reduction in galvanic corrosion rates, the area of the noble metal should be small relative to that of the active metal. For example, brass or bronze valves and fittings in a steel pipe experience much less localized corrosion of steel at the fittings, whereas a steel fitting in a copper line generally corrodes faster. Although galvanic corrosion is undesirable, the same principle can be applied to protect a large amount of equipment. Typical of such application are galvanized (zinc plated) steel sheets and the use of magnesium and zinc anodes to protect buried steel pipelines[1].

There are three possible ways for prevention of galvanic corrosion: at the design stage, avoid the use of dissimilar metals with large e.m.f. differences; provide for insulation if dissimilar metals are used; and finally, avoid large noble-metal-to-active-metal area ratios.

Erosion type of corrosion occurs when the corroding fluid has high velocity and turbulence which removes the protective (oxide) surface films. Pump impellers, agitator blades, and pipeline fittings (elbows in particular) are specially subject to this form of corrosion. Elimination of solid particles from the flowing fluid, reduction in its velocity and turbulence, and use of harder or more corrosion-resistant material not dependent upon protective film action are some of the preventive measures that can be taken for this form of corrosion. Aeration of the fluid to help in quick build-up of the protective film or use of inhibitors to reduce corrosiveness of the solution are also occasionally practiced.

Cavitation damage in pumps is a special form of erosion corrosion. Formation and collapse of vapor bubbles in centrifugal pumps cause severe mechanical damage to the impeller and pump casing. The removal of the protective oxide film starts normal corrosion. Cavitation damage can be avoided by selecting pumps of adequate design and ensuring that NPSH requirements are fully met.

Crevice corrosion is the result of concentration differences between the surrounding area and a crevice existing normally at gaskets, lap joints, bolts, rivets, etc. Those materials that depend on an air-formed oxide film to achieve their corrosion resistance, such as stainless steel and titanium, are especially susceptible to crevice corrosion.

Pitting corrosion is the formation of depressions (and even holes) on a surface due to concentrated local electrochemical action. Pitting is the most serious of the localized cell types of corrosion. Detection is difficult in early stages but metal perforation occurs rather rapidly with thin walled equipment components. This is so because the anodic areas do not shift as in

---

(1) "Chemical Engineers' Handbook" edited by Perry et al; 4th edition, pages 23-3 (1963).

general (uniform) corrosion and loss of metal occurs at only a number of isolated spots. Aluminum and stainless steel alloys in chloride containing aqueous environments (even a few ppm) show strongly the pitting type of corrosion although other metals can have similar problems. Treatment with dichromates is frequently effective where chlorides are the cause of the trouble. A clean, homogeneous and polished metal surface is less likely to develop pitting compared to one with many inclusions, defects, and a rough surface.

Exfoliation is subsurface corrosion that begins on a clean surface but spreads below it to cause whole layers of materials to rust away. The attack is usually recognizable by a flaky and occasionally blistered surface. It is well known in aluminum alloys and can be combatted by heat treatment and alloying.

Selective leaching or parting corrosion involves the removal of one element from an alloy leaving the material spongy and structurally weak. The dezincification that occurs with some copper-zinc alloys containing more than 15% zinc is typical of this type of corrosion. Remedial action involves the use of nonsusceptible alloys.

Metal failures can sometimes occure without apparent loss of metal through uniform corrosion. This is due to stress-corrosion cracking resulting from the combined action of residual or applied stress and chemical corrosion. It is usually preceded by a fine pitting, with the cracks starting in a pit. Transgranular or intergranular cracking is observed in almost all metals or their alloys but each requires certain environmental conditions to produce this form of corrosion cracking. Typical examples include the cracking of austenitic stainless alloys in the presence of chlorides and the caustic-embrittlement cracking of steel in caustic solutions. Intensity of stress, and levels of temperature and concentration are very important factors in stress-corrosion cracking. Temperatures below 50°C are safe for austenitic stainless alloys regardless of chloride concentrations. However, at higher temperatures, even a few ppm of chlorides can bring about metal failure if it is conjointly subject to a certain minimum stress.

Intergranular corrosion results from localized electrochemical attack along the grain boundaries of an alloy, usually because the grain boundary regions contain material that is anodic to the central region of the grains. Austenitic stainless steels, high-nickel alloys, and aluminum alloys are especially susceptible to this form of corrosion. Annealing after welding operations, use of columbium — or titanium-stabilized grades of stainless steel, and reduction of carbon content (0.03 percent maximum) solve the corrosion problem in general.

Finally, deterioration of steel or iron may occur directly or indirectly as a result of the metabolic activity of microorganisms. Coatings are generally used as a preventive means where biological corrosion is suspected. Chlorination of water bearing these microorganisms is also an effective treatment to prevent growth of marine organisms, formation of slimes, etc.

B. APPLICATION OF CORROSION TEST DATA TO MATERIALS SELECTION

On the basis of the general materials philosophy as discussed above and experience reported by salt-making and marine water distillation plants, the following materials were selected for corrosion tests at the Wrightsville Beach pilot plant of the OSW[1]:

- Low carbon steel tubes.
- Copper, admiralty, aluminum brass, 90-10 cupro-nickel, and aluminum tubes.
- Steel, cast iron, admiralty, copper, aluminum brass, 90-10 cupro-nickel, and aluminum test spools in the vapor head and blowdown tank of the LTV evaporator.

The tubes and test spools were subjected to the following conditions:

- Temperatures — 75°F to 250°F.
- Brine concentrations — normal seawater to 4 times the normal salinity.
- High oxygen concentrations at high temperatures.
- Exposure to humid marine atmospheres during shutdown periods.
- Corrosive conditions imposed by acid injection and erosion imposed by sludge recirculation.

---

(1) "Operation of Pilot Plant LTV Evaporator at Wrightsville Beach, North Carolina" by W. L. Badger and Associates, Inc., OSW R&D Progress Report Number 26.

Experts of the International Nickel Company and engineers of the W. L. Badger Company inspected and examined these test spools and tubes after sufficient exposure to the corrosion conditions described above. The following conclusions were drawn and reported in OSW R&D Report Number 26 on page 92:

> "From the accumulated data and observation of the materials of construction used in the evaporator and tube equipment, it may be concluded that steel can be used as a material for the construction of the major portions of the evaporators, pumps, piping, and perhaps heating equipment. Aluminum may be completely ruled out. Of the less expensive nonferrous alloys, admiralty metal, copper, aluminum brass, and 90-10 cupro-nickel are entirely satisfactory on the basis of corrosion results. As a broad general conclusion it is highly probable that steel will turn out to be the best all around material for economical operation. It is essential, however, that further work be done in this regard and, accordingly, we recommend that individual heating bundles in the demonstration plant be made of admiralty, aluminum brass, and cupro-nickel in addition to mild steel."

As can be seen from Table 2.12, Chapter 2, "Plant Design", these recommendations were adopted in toto by the Freeport original plant constructors, Chicago Bridge and Iron Company.

C. SUMMARY OF PLANT CORROSION HISTORY

Performance of the various materials under actual plant-type operating conditions at Freeport over an eight-year period is discussed below in the following subsections:

1. Evaporator effects
2. Preheat exchangers
3. Final condensers, HX-312 and HX-318
4. Deaerator-decarbonator
5. Pumps
6. Piping

1. Evaporator Effects

There are four major components of an evaporator effect — body or outer shell of the effect with its lower part acting as brine sump, top waterbox, tubesheets, and heating element tubes. Bodies of all the effects were made of carbon steel A-285C. Annual Report Number 1 by Stearns-Roger Corporation contains the following conclusions on page II-15:

> "The carbon-steel bodies of the evaporators appear to be relatively free of corrosion and capable of functionally performing their intended service. However, it is recommended that a good coating of anti-corrosion material be applied to the inside of the vessel to minimize formation of iron oxide scale that is carried by the brine and deposited on the tube sheets downstream."

Considerable pitting-type corrosion was noticed in the second year of operation at Freeport. The following observations are reported in the Second Annual Report (OSW R&D Number 100) on page III-7:

> "Tube Materials. As pointed out earlier in Section II, it was necessary to remove all carbon-steel evaporator tubes after eighteen months' service and to replace them with aluminum-bronze tubes.
>
> "All of the copper-base evaporator tubes are in excellent condition and have operated satisfactorily (no failures) to date. The copper-based alloys include 90-10 copper-nickel, aluminum-bronze, and inhibited Admiralty Bronze.
>
> "Corrosion (Pitting). The carbon-steel shells and tube sheets, particularly in the high-temperature effects, are definitely pitting. The greatest pitting has been observed on the top tube sheets of the effect I and V evaporators. This pitting has not reached the critical stage as yet, but should and will be periodically examined. In addition to the obvious disadvantage

> of loss of metal, the corrosion of carbon-steel components has the added disadvantage of introducing flakes of iron (corrosion products) into the brine stream. Many of these flakes become lodged in the holes of the evaporator distribution plates, thereby inhibiting the brine circulation.
>
> "It is felt that an original installation coating of a protective material (such as baked-on phenolic) would have minimized this problem."

It is important to note that this accelerated corrosion of the steel tubes, tubesheets, and shells could have been the result of operating the plant at very low pH's, which was attempted in order to dissolve the anhydrite scale from the high temperature effects.

The Third Annual Report (OSW R&D Number 123), covering the operations for the period July 1963 through June 1964, contains the following corrosion history of the evaporator effects:

- First Effect — This effect was exposed to temperatures above 240°F. The waterbox and tubesheet experienced considerable corrosion and erosion damage. The vapor body corroded to a lesser extent than the waterbox. Two protective coatings were tested at Freeport: Dampney Apexior #1 appeared partially successful while an epoxy coating of Brutem 30 failed within a month. Since the baked-on phenolic material in brine piping from this effect provided effective service it may also be suitable for service inside of the evaporator.

- Second Effect — This effect was in about the same condition as the first effect. Pitting was a little deeper in the waterbox. The carbon steel distributor plate failed and was replaced.

- Third Effect — This effect has also experienced corrosion to almost the same extent as the first two. The Brutem 30 coating was in good condition after 90 days in service.

- Fourth through Ninth Effects — These effects experienced progressively lesser corrosion in their waterboxes and vapor bodies.

- Tenth, Eleventh and Twelfth Effects — These were essentially in good condition.

During the fiscal year 1965, the corrosive deterioration of the evaporator waterboxes had advanced so far that liners had to be installed for safe operation of the plant. A 90-10 cupro-nickel lining was installed in Effects I, II, and III. Monel welding electrode 140, 180, and Incoweld A electrodes were utilized for compatibility with the lining and the base metal. A 1/4 inch thick carbon steel liner was installed on the inner walls of the waterboxes of Effects IV and VI, followed by a coating of epoxy.

During the fiscal year 1966, the waterbox of Effect IV was lined with a welded-in 316L stainless steel liner to replace the badly corroded and perforated carbon steel liner installed only a year ago (November 1964). A 90-10 cupro-nickel liner was also installed in the waterbox of Effect V, although no leaks had yet developed in the original material. In April 1966, while air-testing all evaporators during a shutdown, approximately 110 leaking admiralty brass tubes in Effect IX were discovered. Sections of the failed tubes were sent to Anaconda American Brass Company, Chase Brass and Copper Company, and Cyrus W. Rice and Company for evaluation. The major cause of failure was attributed to intercrystalline corrosion. Symptoms of impingement attack were also present.

During the fiscal year 1967, a random eddy-current testing (Probolog) of tubes in selected evaporators and heat exchangers was conducted by Dr. Frank C. Parker and a full report was submitted by him to OSW. The following excerpt is taken from the Sixth Annual Report, page 55:

> "In general, the inspection divulged that both 90-10 Cupro-Nickel and Aluminum-Brass are performing well, but admiralty is deteriorating. A 'sonoray' sonic test was performed on carbon steel vessel walls, with inconclusive results; however, no serious problems were encountered."

Dr. Frank Motoh of the Stearns-Roger Home Office was assigned to the plant for two months during the fiscal year. He prepared detailed initial materials evaluation report forms (MERF).

In the fiscal year 1967, a number of equipment modifications and replacements were introduced when converting the plant from 12 to 17 effects.

During the fiscal year 1968, sumps in Effects VIII and IX were lined with concrete. Except Effect VIII all other effects had their top waterboxes lined with 90-10 cupro-nickel, 316L stainless steel, or monel. The five modular effects XIII through XVII, had fiberglass-reinforced polyester domes. After careful examination of the Freeport Plant, Cyrus William Rice and Company concluded that[1] "Corrosion of the steel tubesheets in the evaporators, while greater in the hotter effects, was more general and much less severe than in the more turbulent (velocity effect) heat exchangers. Carbon steel waterboxes, unlined or lined with steel, were extensively rusted, pitted, and scaled."

During the first part of fiscal year 1968, considerable time was spent preparing input for the new Materials Evaluation Program. The isometric sketches and detailed data regarding materials of construction were assembled in a manual, "MERF Manual - Appendix 1, Sea Water Desalting Demonstration No. 1," and 424 Material Evaluation Report Forms were prepared for input into the Material Evaluation Report System.

At the shutdown between Development Runs Number 14 and 15, selected tubes in several bundles were probolog tested (for wall thickness). About 18 percent of the titanium tubes in Effect IX showed gross welding defects. The five Type 304 and 316 stainless steel tubes in Effect I showed general minor attack while 75 percent of the 40 stainless steel tubes (316L) tested from Effect XIV showed signs of shallow pitting. Mild pitting attack was also noticed in the four titanium tubes in Effect I.

Revere Copper and Brass, Inc., of New York examined two 90-10 cupro-nickel tubes from Effect XIII and reported (OSW R&D Number 4, page 233) that these tubes had suffered no measurable corrosion in their 130 days of operation. No preferential attack was observed in the weld area.

Corrosion history of the evaporators during the final fiscal year, 1969, was reported in the Eighth Annual Report as follows:

- At the beginning of Run 15, Corrater probes were located in the first effect sump and the sump-to-sump transfer line between Effects XIII and XIV. The calculated corrosion rate was found to be approximately 45 MGPY (mils per year) for dissolved oxygen levels around 40 ppb. The corrosion rates

---

(1) Seventh Annual Report by Stearns-Roger Corporation, published as OSW R&D Report Number 479, page 229 (1969).

and pitting indices at both points followed each other closely.

- A number of holes were found and repaired in the steam chest of Effects II and III just above where the monel expansion joint is welded to the carbon steel evaporator steam chest.

- After more than 8500 operating hours, the cement-mortar lined brine sumps of Effects VIII and IX appeared to be in good condition.

- CDA 194 copper alloy tubes in Effect XIII showed no signs of corrosion.

2. Preheat Exchangers

Shells, tubes and pass partitions were the three main components of the heat exchangers from the materials and corrosion standpoint. The original shells were constructed of A-285, grade C flange quality steel, the tubes were varied in the different exchangers between carbon steel, aluminum brass, admiralty, and 90-10 cupro-nickel, and the pass partitions were made of carbon steel in all exchangers except HX-212, -213, -214 and HX-311, -312 where cupro-nickel cladding was provided. Tubesheets and waterboxes in these latter exchangers were also cupro-nickel clad to provide protection against the non-deaerated seawater they handled. The following is a yearwise study of corrosion in the heat exchangers.

a. First year — Besides observing the corrosion status of heat exchangers and other equipment, test coupon corrosion study was conducted by the International Nickel Company. It was concluded in the First Annual Report that "carbon steel does not give satisfactory service in heat exchanger tubes, tube sheets, baffles, and piping when exposed to non-deaerated seawater."

The corrosion test report from the International Nickel Company stated (in Annual Report Number 1, Appendix B), that "austenitic stainless steels are highly susceptible to localized pitting and crevice corrosion... use of such materials should be restricted to low velocity seawater... copper-nickel alloys (70-30, 90-10) and 2% aluminum brass gave excellent performance under all test conditions. The relative general superiority of the copper-base alloys over steel is amply illustrated in the attached graph [Figure 6.1], steel

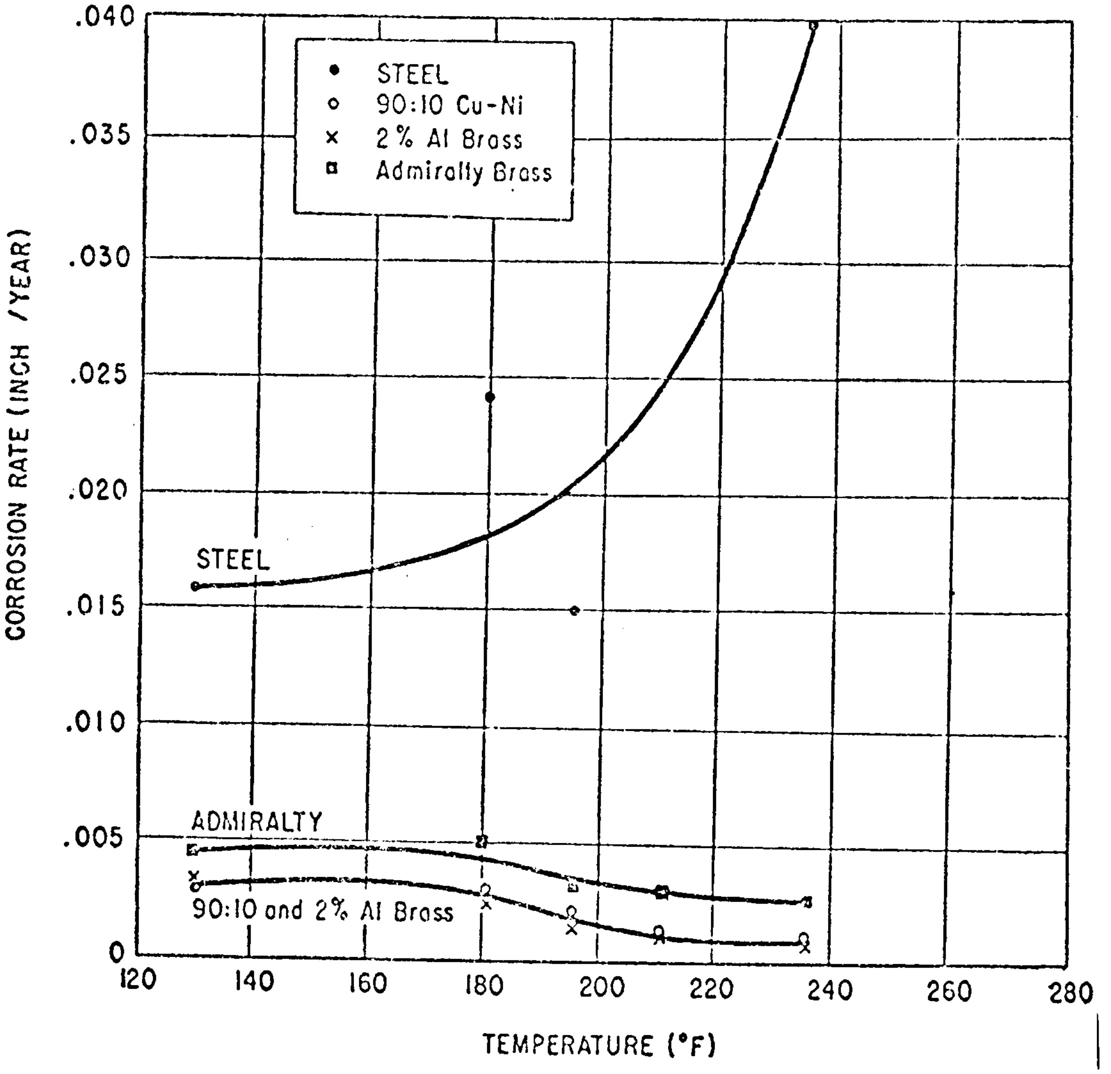


Figure 6.1. Corrosion Rate of Metals in Hot Seawater

being a poor performer under the conditions existing at the test locations. The performance of Admiralty brass is regarded marginal to poor."

An inspection carried out to locate the leakage of salt water in HX-202 in August 1961 revealed extensive corrosion on the carbon-steel tube plugs. Other exchangers showed similar problems and some 1700 plugs were removed and replaced with plugs of material identical to that of the tubes in each particular exchanger.

In September 1961 Hurricane Carla hit the Freeport area and caused extensive damage to the plant. The hurricane water level averaged six feet at the plant site. In March 1962 a general inspection of the heat exchanger tubes and waterboxes was carried out. This inspection indicated that the corrosion rates in the carbon-steel tubes were a function of the temperature range of the heat exchangers — HX-201 showed the worst corrosion while HX-211 the least. It was also found that the major portion of the corrosive action was confined to those components that are exposed to non-deaerated seawater. The corrosive action had reduced wall thicknesses by about 30%. Carbon steel tubes from HX-201, -208, -211, and HX-301, -305, and -308 were replaced with aluminum brass tubes.

b. Second year — A microbiological corrosion survey was conducted by Sharpley Laboratories, Inc., of Fredericksburg, Virginia. The results of the survey revealed no detrimental microbiological corrosion; however, the presence of marine growths in specific areas of the plant were noted as a possible source of trouble in continued operations. Chlorine injection (about 10 ppm) in the feed seawater line was successful in killing marine growth in HX-215.

c. Third year — The waterbox on HX-201 showed considerable corrosion. HX-208 through 211 displayed slight corrosion in their waterboxes. HX-209 developed a leak to the atmosphere in its waterbox. Waterboxes and pass baffles in the 300 Series heat exchangers were deteriorating, especially in the high temperature end of the plant. HX-301, -304, and -306 had pass baffle failures.

d. Fourth year — The Fourth Annual Report contains the following remark on page 47:

> "The waterboxes in all heat exchangers had deteriorated to such an extent that the Plant was losing its on-stream reliability. Holes as large as one-half-inch diameter had penetrated the heat exchanger shell in many places. Alloy and coated linings were, therefore, installed in the waterboxes of most of the exchangers. HX-301 and -305 were lined with monel; HX-302 through -304 with 316L

stainless steel; HX-201, -209, and -306 through -310 with carbon steel; and HX-201, -210, -309, and -310 <u>lower</u> waterboxes with carbon steel and protected with coating of air-dried epoxy. Equipment was installed in three heat exchangers to evaluate the effects of impressed current on corrosion rates.

e. <u>Fifth year</u> — The Fifth Annual Report (pages 192-195) by Stearns-Roger contains the following observations regarding the heat exchanger corrosion status in the fiscal year 1966:

> "There have been no failures of the 70-30 copper-nickel cladding provided at tube sheets and waterboxes of non-deaerated seawater feed heat exchangers. It is possible that 70-30 copper-nickel cladding was a contributing factor to the failure of the admiralty brass tubes in Heat Exchanger 311. The tube failures occurred close to the tube sheet and the tubes were anodic to the clad surfaces and of a lessor exposed surface area."

Monel liners and pass partitions in HX-301 and HX-305 did not show deterioration even in areas of high-velocity impingement on the partitions.

Although most published data discourage the use of stainless steel in chloride media because of pit susceptibility and stress-cracking, the 316L stainless liners and pass partitions installed in the upper waterboxes of HX-302 through -304 provided satisfactory protection. The protection provided by 316L liners compared favorably with those of cupro-nickel and monel.

Aluminum-brass and 90-10 cupro-nickel tubes provided excellent corrosion protection with no failures.

f. <u>Sixth year</u> — As mentioned above in the subsection on evaporators, the major activity during this year was centered around the probolog and sonoray testing of tubes and vessel walls. In general, the inspection showed that both 90-10 cupro-nickel and aluminum-brass were performing well, but admiralty was deteriorating.

Mr. Frank Motoh of the Stearns-Roger Home Office prepared detailed materials evaluation report forms.

g. Seventh year — CDA-194 tubes in HX-307 showed two new pits and some general shallow pitting. HX-303a upper waterbox developed a leak which was traced to a hole in the stainless steel weld that bound the liner to the baffle. Sandblasting of an apparently unaffected tubesheet revealed considerable attack, especially in the area directly adjacent to the weld binding the alloy liner to the carbon steel tubesheet. Later, some selected exchangers were sandblasted to check if there was any corrosion. The results are presented in Table 6.1. Examination of this table revealed that HX-302, -303a, -303b, -306b, and 307 had serious tubesheet corrosion. The all-alloy exchangers (HX-304, -305, and -312) were not seriously affected. Those tubesheets where the liner was installed in 1964 showed the most serious attack relative to the tubesheets where the lining was done in 1967. This would indicate that corrosion was occurring uniformly but the extent of the problem was masked by the heavy accumulation of corrosion products.

The repair to the worst of these tubesheets was accomplished by filling the pits using carbon steel welding rod and then rebonding the liner to the carbon steel tubesheet with alloy welding rod.

HX-307 tubes of CDA-194 (proprietary copper-iron-phosphorus alloy) showed very minor, generally uniform attack after 160 days of operation at pH 6.5 to 7.5 and dissolved oxygen level of between 10 and 20 ppb.

h. Eighth year — Parting corrosion (denickelification) was noticed in HX-304 and -305 top waterbox liners of 70-30 cupro-nickel.

In October 1968, HX-200 was removed from the plant as its waterboxes were perforated and the admiralty tubes severely corroded. Carbon steel liners in upper and lower waterboxes failed and were replaced with carbon steel 285C plate.

The General Electric RTV-112 silicone rubber coating in the inlet waterbox of HX-207 separated from the tubesheet.

Table 6.1

Inspection of Preheat Exchanger Upper Waterboxes — January 1968

HEAT EXCH NO.	TUBE MATERIAL	LINER INFORMATION		APPROXIMATE OPERATING TEMPERATURE °F	SAND-BLASTED	INSPECTION REMARKS
		MATERIAL	DATE INSTALLED			
302a	Al-Brass	Monel	3/64	260	Yes	Not too bad, tube sheet attack about 3/8-inch deep
302b	Cu-Ni 90-10	316L	3/64	255	Yes	Bad, typical attack about 7/8-inch deep
303a	Admiralty	316L	3/64	255	Yes	Bad, typical attack about 1-inch deep with considerable undercutting of liner
303b	Al-Brass	316L	3/64	250	Yes	Bad, typical attack about 1-inch deep with considerable undercutting of liner
304	Al-Brass	70-30 Cu-Ni	Original	240	No	No attack on tube sheet. 1-1/4-inch diameter by 3/32-inch deep pit in east side of baffle
305	Al-Brass	70-30 Cu-Ni	Original	230	No	No attack on tube sheet
306a	Al-Brass	316L	5/67	225	No	Bad, typical attack about 5/8-inch deep. Tube sheet uniform attack about 1/4-inch
306b	Al-Brass	Monel	3/64	220	Yes	Bad, typical attack about 1-inch deep
307	Cu-Fe	316L	9/67	205	No	Bad, typical attack but not as deep. Tube sheet uniform attack about 1/2-inch
308	Al-Brass	316L	5/67	195	No	Not too bad; pits starting around outside but weld attack has not started. Tube sheet uniform attack about 1/4-inch
309	Al-Brass	316L	9/67	185	No	About the same as HX-308
310	Admiralty	316L	9/67	175	No	About the same as HX-308
311	Al-Brass	Carbon Steel	1/68	165	Yes	About the same as HX-308
312	Cu-Ni 90-10	70-30 Cu-Ni	Original	155	No	No attack on tube sheet

About 200 material performance forms (MPRS) were submitted to the Materials Test Center of the Oak Ridge National Laboratory.

A summary of conclusions regarding the suitability of the metals and alloys tested at Freeport in different services is given at the end of the Subsection C. 6, "Piping".

3. Final Condensers — HX-312 and HX-318

Trouble-free operation of these condensers is very essential for maintaining the rear-end temperatures (which control plant capacity) and the plant production rate. In the original 12-effect plant, HX-312 condensed vapor from Effect XII using raw seawater (non-deaerated) as a coolant. Originally, this unit had tubes of aluminum brass while the tubesheets were lined with 70-30 cupro-nickel. Magnesium anodes (DOW Type 24W) were also installed in both the top and bottom waterboxes as a cathodic protection system.

During the first three years of its operation, the final condenser had no serious corrosion problem except for the fact that the sacrificial magnesium anode dissolution rate was rather high.

After the Freeport Plant conversion to 17 effects, the original heat rejection condenser HX-312 was retired and a new unit HX-318 was designed for the modified conditions. This unit had both aluminum brass and 90-10 cupro-nickel tubes and Gaco EHS-55 Epoxy coating in the upper and lower waterboxes. This coating in the inlet waterbox had failed during the final year of operations (FY-1969).

4. Deaerator-Decarbonator

The important components of the deaerator at Freeport were: the shell, the packing, and the seawater spray nozzle. Originally these parts were made of carbon steel, ceramic, and Saran plastic respectively. The deaerator body (i.e., shell) was coated internally with epoxy to provide resistance to corrosion from the oxygen-rich and acidified raw seawater feed.

As long as the coating was intact, there was no corrosion problem in the deaerator. But, sometime during the second year the spray nozzle dropped off and the feed seawater stream impinged on the ceramic packing, the movement of which caused the paint to come off and initiated severe pitting corrosion of the shell. A 15 to 30 mil coating of coal-tar epoxy was applied to the shell after sandblasting it. Inspection in the third fiscal year revealed that damage and minor failure of the coating had occurred at the ceramic packing — shell contact points. The ceramic raschig rings were later replaced with polypropylene "Maspac FN-90" manufactured by the Dow Chemical Company.

In fiscal year 1967, the deaerator-decarbonator internal coating exhibited numerous penetrations at the spray impingement zone near the top of the packing. After sandblasting the complete interior of the vessel, a new Brutem 110 epoxy coating was applied. The impingement zone was further protected with a buildup of fiberglass mat and Brutem 130 epoxy.

In the final year of plant operation (FY-1969), Brutem coal-tar epoxy coating of the upper 4 feet of the deaerator required repairs. The service life of this coating has been noted to be about 18 months.

5. Pumps

There were mainly seven categories of pumps used at the Freeport Plant — condensing water pumps (P-1, P-2), seawater feed pumps (P-3, P-4, P-5), brine blowdown pumps (P-11 through P-22), condensate pumps (P-31 through P-43), flashed condensate pump (P-44), product water pumps (P-50 and P-51), and seal water pump (P-47). The pumps utilized iron and steel construction except the condensing and seawater feed pumps which utilized IB Ni-Resist impellers. The wear ring, lantern ring, gland nuts, and shaft sleeve were made of monel in all pumps, except the condensing water pump (see Table 2.12, Chapter 2, "Plant Design").

Because of the continuous duty of the pumps and the corrosive nature of the fluid handled at high velocities, corrosion was a continuing problem with pumps at the Freeport Plant.

a. <u>First year</u> — The capacity of pumps P-1 and P-2 reduced by 30% within ten months of operation. Inspection revealed severe cavitation damage in the second stage propellers and in the bowls adjacent to the propeller tips. Straightening vanes in the pump casings were also damaged. It was concluded that, in addition to the improper pitch of the propellers as installed, the materials of construction of the impellers and casings were not suitable for the severe corrosive service involved.

Severe cavitation in pumps P-4 and P-5 resulted in recurrent impeller and shaft failure problems. The impeller leading edges on the P-11 through P-22 brine pumps showed indications of slight cavitation damage. Graphitization of the cast iron impellers of the condensate pumps P-31 through P-43 was noticed. With the removal of heat exchangers 202 through 207, the corresponding pumps P-32 through P-37 were also removed from service.

b. Second year — Material failures toward the end of the first year necessitated certain changes in the pumps P-1 and P-2. The bowls of both pumps were fiberglass-epoxy coated. Ni-Resist epoxy-coated propellers in P-1 and bronze propellers in P-2 were the additional changes introduced. Special heavy-duty, stress-relieved, Ni-resist impellers were installed in pumps P-4 and P-5 discharging the deaerated-decarbonated seawater from the deaerator. After 14 months, this special impeller showed corrosion pits which penetrated through the metal in some spots. NPSH available at these pumps is extremely low so that cavitation conditions existed constantly. Brine pumps had performed well with the original materials so far. Both the condensate and product water pumps performed with normal maintenance. Some graphitization of the cast iron impellers handling low pH (5.0 to 5.5) was evident.

c. Third year — Bowls of pumps P-1 and P-2 were again epoxy-coated. Pump P-5a, capable of operating at very low NPSH, replaced pump P-5. After six months in operation, this vertical, can-type pump developed a crack on the casing and graphitization of the suction bell. Cavitation damaged the impeller of P-44. The ruined impeller had to be replaced with a spare. Condensate and product pumps worked satisfactorily with only bearing and packing replacements. Acid feed pump P-46 (Lapp Pulsafeeder), which handles 94 percent sulfuric acid solution, required no excessive maintenance.

d. Fourth year — Pump preventive maintenance was reported (Fourth Annual Report, page 51) to have been put on a regular schedule during this fiscal period. Severe cavitation in pump P-2a damaged the second stage bronze propeller blades. There was no corrosion damage to the stainless steel shaft, bowl castings, or cap screws. This pump was reassembled using one monel and one Ni-Resist propeller. The seal cage in P-3 had completely corroded away, and cracks were found in both the suction cover and the discharge flange. Brine pumps P-11 through P-22 displayed very little corrosion damage.

Because of the low pressure on the suction side of the condensate pumps P-31 through P-43, cavitation conditions always existed and new impellers were installed in four of these pumps.

e. Fifth year — A suction drum, with automatic level control, was installed to provide a constant suction head to pump P-3. Cavitation in pumps was significantly reduced. A new horizontal, centrifugal pump P-11b was installed as Effect I blowdown pump (replacing P-1) to test an improved design equipped with a mechanical seal instead of the original asbestos packing.

In late October 1965, pumps P-1 and P-2 were disassembled and inspected. The carbon steel shaft on pump P-1 was severely corroded and the bronze propellers in pump P-2 (which had only been in service for six months) were badly deteriorated. As a result of this inspection, materials in these pumps were upgraded. A new 316L stainless steel shaft and two new 316L stainless steel propellers were installed in pump P-2. Both pumps were returned to service in November 1965.

In February 1966, a blade on one of the monel propellers failed; this was the second such failure. A used, 316L stainless steel propeller was installed in place of the broken monel propeller, and pump P-2 was again returned to service.

Operation of these pumps indicates that 316L stainless steel is a good material for propellers and shafts in seawater service. Although monel resists corrosive effects of seawater, failures have been encountered with monel propellers due to an apparent difficulty in attaining a homogeneous casting of this material.

The housing around the stuffing-box busing in pump P-3 had corroded to such an extent that it was necessary to replace this pump with an iron pump equipped with bronze impellers. The suction cover was coated with epoxy for corrosion protection. The standby pump P-4 experienced numerous shaft and impeller failures. Both stainless steel and carbon steel shafts were used.

In November 1965, after only two weeks of operation, the mechanical seal in P-11b began to leak. Subsequent inspection showed a groove had been cut, presumably by silt, in the carbon seal ring. Replacement of this ring stopped further leakage. Cast iron suction covers of P-21 and P-23 corroded extensively. The cast iron impellers on P-40, -41, and -42 deteriorated to the point that design capacity could not be achieved. These were replaced with bronze impellers. Acid pump P-45 was disassembled and new spare parts (plunger, yoke, seals, bearings, and check valves) were installed.

f. Sixth year — Cast iron casings of the original brine pumps were failing due to graphitization. Ni-Resist bowl of P-5a cracked.

g. Seventh year — Pump maintenance continued to be generally high because of casing, suction cover, and impeller failures. The raw seawater intake pumps P-1 and P-2 required considerable maintenance.

h. Eighth year — Centrifugal brine and condensate pumps required only normal attention during this year. The pumps equipped with mechanical seals required frequent attention. Some failures occurred due to brine corrosion of seal compression springs. The Eighth Annual Report concludes on page 80: "At present, mechanical seals are not recommended for desalination plant service." Corrosion of carbon steel suction covers was a continuing problem. P-3 had numerous failures — most were due to mechanical (and not corrosion) reasons. Some of the failures were due to foreign material entering the pump body along with the feed. Recommendations for corrosion resistant pump materials for desalting plant service are given following Subsection C. 6.

6. Piping

Piping in the Freeport Plant transported six different fluids — seawater, condensate, vapor, brine blowdown, acid, alkali, and raw seawater (underground). Specifications for the original design are given in Table 2.12 in Chapter 2. High velocity flow of corrosive fluids like hot seawater subjects the internal surface of the piping to considerable corrosion. The Third Annual Report states, on page 49: "Repair and replacement of seawater and brine piping is the greatest, single maintenance item of the Freeport Plant."

a. First year — Some sections of carbon steel pipe lost as much as 30 percent of wall thickness, due to corrosion. Carbon steel pipes that were carrying deaerated seawater did not suffer excessive corrosion. PVC and fiberglass epoxy pipe sections were used in the line carrying acidified seawater to the deaerator. The original coating of aluminum paint used for covering the exterior of all piping began failing shortly after the plant was put into operation.

b. Second year — All the brine pump discharge lines of the first eight effects failed. These failures were in the same spot on each line — i.e., at the first weld seam downstream of the control valve. On the first three effects, carbon steel piping with a baked-on phenolic interior coating were substituted for the failed lines. Other lines were repaired by welding. Although pure condensate is considered extremely corrosive, only one noteworthy failure occurred in this piping, which involved the drain line from condenser 304. Cathodic protection that had been installed to protect the travelling screen and the end-of-pipe gate valves proved satisfactory.

c. Third year — Carbon steel piping failures continued as before. All piping failures which required maintenance were located adjacent to weld seams or on the bottom of the pipeline. Two stress-relieved sections of pipe were installed adjacent to the weld seams of the brine pump P-15 discharge. Numerous failures had occurred at this location in the past. Dissolved oxygen, high temperatures, and high velocities (turbulence) are the biggest factors that contribute to carbon steel piping failures. At Freeport, it was clearly established (OSW R&D Report Number 123, pages 50-51) that considerable buildup of corrosion products existed in nondeaerated seawater piping whereas the deaerated line was essentially free of such buildup. The condensate, steam, and vapor lines displayed little corrosion in the third year of operations at the Freeport Plant.

d. Fourth year — The Fourth Annual Report again commented (on page 55) that: "The largest single item of maintenance is in piping repair and replacement." Brine pump suction and discharge sections failed most frequently throughout the plant. These were replaced with phenolic-coated piping. Corrosion was severe in the high temperature end of the plant. Several failures occurred in the reducing elbows and horizontal piping around HX-215. Condensate piping in the high temperature end of the plant began to show corrosion and experienced some failures.

A historical summary (to 1966) of the seawater feed piping is presented in Table 6.2. A similar summary of brine blowdown piping is given in Table 6.3. Operating conditions in the blowdown piping are severe and numerous failures were encountered. Turbulence in pipe-size reducers (at brine pump suction and discharge) and at level control valves accelerates corrosion attack on carbon steel and results in short material life.

Table 6.2

Summary of Seawater Feed Piping

LOCATION		PRESENT MATERIAL	REPLACEMENT SPOOLS		REMARKS
FROM	TO		NO.	MATERIAL	
HX-201	Effect 1	Carbon steel internally coated with baked-on phenolic	2 1	Carbon steel Present material	1 to 2-year life Unknown life
HX-301	HX-201	Dow-Smith Chemline	1 1	Carbon steel Present material	1 to 2-year life Unknown life
HX-302	HX-301	Fibercast CS-300	1 1 1 1	Carbon steel Fibercast J-300 Fibercast J-300 Present material	1 to 2-year life Leaks at flanges due to improper assembly Failed due to poor quality control of piping material Caustic-resistant grade for additional tests
HX-303	HX-302	Carbon steel	2	Carbon steel	2 to 3-year life
HX-304	HX-303	Carbon steel	2	Carbon steel	2 to 3-year life
HX-305	HX-304	Carbon steel	2	Carbon steel	2 to 3-year life
HX-306	HX-305	Carbon steel	2	Carbon steel	2 to 3-year life
HX-307	HX-306	Carbon steel	1	Carbon steel	4 to 5-year life
HX-208	HX-307	Carbon steel	1	Carbon steel	4 to 5-year life
HX-308	HX-208	Carbon steel	1	Carbon steel	4 to 5-year life
HX-209	HX-308	Carbon steel	none		Original material - no leaks
HX-309	HX-209	Carbon steel	none		Original material - no leaks
HX-210	HX-309	Carbon steel	none		Original material - no leaks
HX-310	HX-210	Carbon steel	none		Original material - no leaks
HX-211	HX-310	Carbon steel	none		Original material - no leaks
Deaerator LCV	HX-211	Carbon steel	1	Carbon steel reducer	Reducer failed in weld area
Deaerator	Deaerator LCV	Carbon steel	1	Carbon steel	Failed at P-5a suction - 4 to 5-year life
HX-212	Deaerator	Carbon steel	1 3	Carbon steel elbow Carbon steel reducer	Failed at elbow out of HX-212 - 3 to 4-year life Reducer failed at FCV - 1-year life
HX-311	HX-212	Carbon steel	1	Carbon steel	4 to 5-year life
HX-213	HX-311	Carbon steel	none		Original material - no leaks
HX-214	HX-213	Carbon steel	none		Original material - no leaks
HX-215	HX-214	Carbon steel	1	Carbon steel	4 to 5-year life
Pump P-3	HX-215	Carbon steel	none		Patched leaks in P-3 discharge elbow
Pumps P-1 and P-2	HX-312	Carbon steel	none		Original material - no leaks
Pump P-53	Barometric Condenser	Carbon steel	none		Original material - no leaks
Pump P-53	Screen	Carbon steel	2	Carbon steel	Runs on ground - lies in surface water

NOTE: Certain spool pieces have operated for longer durations than the approximate life indicated above, however, this was accomplished by patch-welding type repairs to the outside of the pipe

Table 6.3

Summary of Brine Blowdown Piping

PUMP NO.	PUMP SUCTION - RISERS		PUMPS SUCTION - ELBOWS		REMARKS
	ORIGINAL MATERIAL AND LIFE	REPLACEMENT	ORIGINAL MATERIAL AND LIFE	REPLACEMENT AND LIFE	
P-11	Carbon steel - 5-year	1 - Carbon steel	Carbon steel - 3-year	Carbon steel with baked-on phenolic	Phenolic coating must pit free
P-12, P-13, P-14, P-15, and P-16	Carbon steel	none	Carbon steel - 3-year	Carbon steel with baked-on phenolic	Phenolic coating must pit free
P-17, P-18, P-19, P-20, P-21, P-22, and P-23	Carbon steel	none	Carbon steel	none	Original suction piping still in service; no maintenance or replacement has been required
	PUMP DISCHARGE - PIPE AND FITTINGS		PUMP DISCHARGE - TO EVAPORATORS		
P-11	Carbon steel - less than 1-year	1 - Carbon steel, 1-YR 2 - Carbon steel with baked-on phenolic, 2-YR	Carbon steel, 4 to 5-YR	1 - Carbon steel	Life of phenolic coating in all cases is a function of quality of material and quality control of application
P-12 and P-13	Carbon steel - less than 1-year	1 - Carbon steel, 1-YR 2 - Carbon steel with baked-on phenolic, 2-YR	Carbon steel, 4 to 5-YR	none	Reparable leaks in riser piping encountered at present
P-14	Carbon steel - 1 to 2-year	1 - Carbon steel, 1 to 2-YR 1 - Carbon steel with baked-on phenolic	Carbon steel, 4 to 5-YR	none	Phenolic coated piece still in service
P-15	Carbon steel - 1 to 2-year	1 - Carbon steel, 1 to 2-YR 1 - Carbon steel with baked-on phenolic	Carbon steel, 4 to 5-YR	none	Riser leaks occurred immediately downstream of P-15 orifice; phenolic coated piece still in service
P-16	Carbon steel - 1 to 2-year	1 - Carbon steel, 1 to 2-YR 1 - Carbon steel with baked-on phenolic	Carbon steel, 4 to 5-YR	none	Riser in service with no leaks, phenolic coated piece still in service
P-17	Carbon steel - 1 to 2-year	1 - Carbon steel, 1 to 2-YR 1 - Carbon steel with baked-on phenolic	Carbon steel, 4 to 5-YR	none	Leaks in riser patched and one section replaced, phenolic coated piece still in service
P-18 and P-19	Carbon steel - 1 to 2-year	2 - Carbon steel 1 to 2-YR	Carbon steel, 4 to 5-YR	none	Riser leaks patched, leaks occurred immediately downstream of P-19 orifice
P-20 and P-21	Carbon steel - 3-year	1 - Carbon steel	Carbon steel	none	Discharge pipe replacement still in service
P-22	Carbon steel - 3-year	1 - Carbon steel	Carbon steel	none	One leak in riser patched
P-23	Carbon steel - 3-year	none	Carbon steel	none	

e. Fifth, sixth, seventh, and eighth years — The same corrosion failure pattern existed as described above. Corrosiveness of untreated and treated product water was also studied in considerable detail during this period and the following conclusions were drawn:

- High-purity product water from distillation-type desalting plants can be expected to be corrosive to unprotected carbon steel pipelines.
- Minor amounts of dissolved oxygen in high-purity product water will significantly accelerate corrosion rates.
- Inorganic complexing agents such as polyphosphates and silicates can significantly reduce corrosion rates and "red water" conditions.

The corrosiveness of brine as measured by the Corrosometer probes located in Effect I sump and the brine transfer line between Effects XIII and XIV indicated that the corrosion rates at both the locations (averaging 35 to 50 mils per year) were similar and they increased or decreased together. The Pitting Index showed a similar behavior.

The Third Annual Report contains the following comment (page 135) on materials and corrosion:

> "Primary emphasis has been placed on using the least exotic metals in this plant. This policy was adopted as a design criteria with full knowledge that some materials would prove unsatisfactory, and require early or frequent replacement. The process of evaluation thus established would result in a final economic list of materials for future design considerations. The following evaluations are presented as recommendations for optimum use of various metals."

The following observations and conclusions have been collected from the Third, Fourth and Fifth Annual Reports by the Stearns-Roger Corporation.

7. Materials

a. Carbon steel — This material will serve satisfactorily in the deaerated seawater circuits. Although carbon steel performance is not perfect, this material is economical when compared to the cost of known corrosion-resistant materials. A glass-reinforced plastic pipe may eventually prove to be less expensive and a more satisfactory replacement for carbon steel. Carbon steel performs very well at temperatures up to 200°F. However, at higher temperatures, a protective coating or a more resistant material is required. When baked-on phenolic coatings are applied, carbon performs well in seawater concentrate service up to 250°F. For satisfactory service in seawater concentrates below 200°F, carbon steel weld joints should be stress-relieved. Carbon steel exposed to steam and condensate above 200°F may contribute to the formation of magnetic iron oxide. This is aggravated by alternate exposure to air and steam, therefore, it may not be a problem during long term, continuous operation. When used as a tubesheet material, the performance of carbon steel can only be termed as minimal. Cupro-nickel (70-30) cladding of tubesheets is recommended to lessen possible galvanic action and erosion, and minimize corrosion. Carbon steel is not satisfactory as a heat exchanger tube material for seawater service.

The evaporator cone, vapor body, steam chest, heat exchanger shells, brine pump casings, deaerator shell, flash tanks, and chemical and product tanks of carbon steel displayed negligible metal loss due to corrosion.

b. Cupro-nickel — This material has performed excellently when used as a cladding for carbon steel in the non-deaerated feed water circuit. Cupro-nickel should be considered as a cladding or solid material for service in the high temperature preheat exchanger waterboxes and tubesheets and as cladding for the evaporator tubesheets and waterboxes in the high temperature effects. This metal's resistance to erosion exceeds that of either carbon steel or aluminum brass; consequently, local high velocity areas and impingement areas (as occur at the upper tubesheets) would be better protected by using cupro-nickel. An economic evaluation, for some applications, would be required to determine the relative merits of cupro-nickel versus carbon steel with baked-on plastic coating.

c. Aluminum-brass — This tube material has provided very good service wherever it has been applied in this plant. In all cases, velocities are less than the 7 FPS upper limit for this material. Aluminum-brass demonstrates some susceptibility toward pitting where fouling is present, or when marine growth becomes attached to the material (as in heat exchanger 215). This material has the capability to withstand shock chlorination when this method is used to control biological growth. Aluminum-brass is superior to the stainless steels for withstanding the effects of inhibited hydrochloric acid cleaning. Aluminum-brass heat transfer capabilities are superior to those of cupro-nickel, but are inferior to those of Admiralty brass. Observation of the two materials has indicated that Admiralty brass has shown somewhat less resistance to corrosion than has aluminum-brass.

d. 300 Series stainless steels — These materials are subject to pitting-type corrosion and stress corrosion cracking. These alloys are austenitic in that they are solid solutions of iron carbide in gamma iron (face centered cubic form). They are non-magnetic and are very corrosion resistant while in this form. Since the structure is a super-cooled solution, this form is rather unstable. Cold work of the metal causes precipitation to occur, resulting in work-hardening and a much lower resistance to corrosion. Work-hardening, however, provides the excellent resistance to cavitation damage exhibited by these alloys.

The mechanism of corrosion resistance of these materials is the formation of an oxide film which passivates the metal. The nickel content is less reactive and supplements the chromium under more severe conditions. When the media becomes reducing at the surface of the material, these alloys may shift up the galvanic series to a vulnerable position approximating carbon steel. This action, in conjunction with work-hardening, is the cause of rapid deterioration of stainless steel pump impellers and shafts in stagnant seawater. Also, contributing to failure of these materials is their susceptibility to chloride stress corrosion cracking, primarily from the presence of calcium chloride, magnesium chloride, and sodium chloride.

Considering these factors, all of which have been proved in the Freeport Plant, it is recommended that the use of 300 Series stainless steels be limited to pump impellers and pump shafts only when the flow of seawater is continuous. If continuous seawater flow cannot be obtained, provisions should be made, during pump outages, for removing the pump from contact with seawater or flushing the pump with fresh water. This would prevent impeller or shaft contact with seawater during non-operating periods. The resistance to cavitation damage warrants giving special consideration to pump removal and/or flushing during shutdown periods despite the minor additional expense involved.

e. Ni-Resist Iron — This material has exhibited good corrosion resistance when used as non-rotating castings in vertical turbine pumps. However, like stainless steel, it has a tendency to corrode in stagnant seawater. Magnetic forms of this material are subject to graphitization, as exhibited by the suction bell on pump P-5a. Therefore, use of this material should be limited to types that are non-magnetic, because magnetic properties indicate either insufficient alloying element or precipitation, and loss of corrosion resistance. This (non-magnetic) material has a lesser tendency to work-harden than 300 Series stainless steel, but demonstrates a better resistance to cavitation damage than does carbon steel or cast iron. If weld repairs are made on Ni-Resist castings, proper heat treatment must be applied to maintain the non-magnetic structure throughout the casting.

f. Monel — Monel was used as brine pump shaft material and as lining for heat exchanger waterboxes and pass partitions in the deaerated seawater circuit. Monel was also used for the evaporator heating element expansion joints. There were no failures in these services.

g. Hastelloy C and titanium — Only slight corrosion was noticed in tubes of these metals used in the heating elements of the evaporator effects. Although suitable from purely a corrosion standpoint, these metals are quite expensive for normal seawater desalting service.

h. Dow-Smith Chemline — Chemline pipe is comprised of continuous glass fibers embedded in thermoset epoxy resin. The glass fibers are filament-wound and there is a resin-rich liner in the inner pipe wall. This material is recommended for continuous use at temperatures up to 300°F. An eight-foot spool piece, including flanges, was fabricated and installed between Heat Exchangers 301 and 201 in February 1966. Since that time, the spool piece has been in service for over 1000 hours at temperature range of 200°F to 265°F with no weakening, leaking, or deformation.

454-997 O - 72 - 9

i. PVC, ABS and other plastics — Test spools of epoxy-lined asbestos cement, unlined asbestos cement, Type I PVC, Type II ABS, fiberglass reinforced epoxy; and fiberglass reinforced polyester were installed in the non-deaerated seawater feed to the deaerator (in temperatures no higher than 150°F). In two years' operation, none of these materials failed. Saran-lined pipe proved quite satisfactory in non-deaerated seawater feed piping from the point of sulfuric acid injection downstream to the deaerator. Bis-phenol-type FRP plastic "Corite" as a distributor plate in the evaporators was not successful since very careful handing of this brittle material is required. Reinforced concrete pressure pipe could provide adequate service as gravity seawater feed intake piping.

j. Coatings — A number of new exterior paints were tested, of which Napko latex-base exterior paint was the most acceptable.

A baked-on phenolic coating proved to be a practical means of extending the life of carbon steel brine piping and other equipment. A 5 to 20 mil thick coating can be shop-applied by spray techniques at 400°F temperature. Pinhole-free coating is essential to avoid accelerated corrosion.

A number of epoxy coatings were tried for protection of evaporator waterboxes. A hot air-dried, catalyzed epoxy with a heat reflective layer has been tried in Effects I, II, III, and V, and in certain heat exchangers. This material will provide good protection to metal in temperatures as great as 240°F but only if it is completely holiday-free. The slightest pinhole will be undercut with the resulting failure of the entire coating. An ambient air-dried epoxy cement was also tried as a coating in Effect I waterbox. Two layers of this material were applied and after six weeks the first layer had failed; the second layer failed after an additional 30 days of service.

D. MOST CRITICAL AREAS OF CORROSION

Every chemical plant is subject to nature's destructive force of corrosion acting on the materials used in its construction. This is especially true of metallic materials. A more important fact, however, is that every plant has certain sections, or components of certain sections, that are most severely under corrosive attack. If the very operation of the process depends on the proper functioning of these sections or components then it becomes imperative to design and construct these sections with great care so that the component failures are minimized. To achieve lower water costs in distillation-type plants, it is necessary to aim at 90 to 95 percent on-stream time. It is possible to reach this goal with a good preventive maintenance program and special attention to the most critical areas of corrosion. A detailed discussion of the corrosion experiences at the Freeport Plant was given above in Section C. The purpose of this section is to briefly point out the critical areas of corrosion.

If it were economically feasible to use the expensive (noble) metals or alloys, there would be no corrosion other than the wear and tear of the metal. For practical reasons, the use of such materials as carbon steel is inevitable. Presence of such factors as the intense turbulence areas, high fluid velocities, slurries, acidic or alkaline conditions, cavitation, and sudden changes of direction or velocity lead to the existence of critical areas of corrosion. The following is a brief list of such areas in a Freeport-type desalting plant.

- Evaporator effect and heat exchanger waterboxes and tubesheets — There was both uniform and pitting-type corrosion in these components when made of carbon steel. High flow rates, intense turbulence, presence of solid corrosion products, and flashing of the brine feed all contribute to the severity of corrosion in the waterboxes. Alloy cladding (cupro-nickel, monel) significantly reduces the corrosive attack.

- Heat exchanger pass partitions — Erosion corrosion caused wide holes to appear in some of the waterboxes. The resulting short-circuiting of flow adversely affects the plant economy. Monel cladding provides adequate protection.

- Non-deaerated seawater circuit — Because of the high oxygen content of the seawater, ferrous metals and alloys are subject to serious attack. Product coolers, final condenser, and other auxiliary equipment need either alloy cladding or a cathodic protection system.

- Brine pumps — Cavitation damage was the most recurrent problem with these pumps. NPSH requirements must be met or even exceeded to reduce the cavitation damage. Impellers, shafts, and pump casings must be made of hard, erosion-resistant metals (e.g., Ni-Resist or 316L stainless steel).

- Piping — As mentioned earlier, the piping repair and replacement represented the biggest single maintenance item at the Freeport Plant. High temperature-high velocity fluid flow, sudden changes of direction, and turbulence were the causes for carbon steel piping failures. Non-deaerated sections of straight-run piping suffered greater corrosion than the deaerated sections of straight-run piping. Suction and discharge piping of the brine blowdown pumps were particularly vulnerable to this attack. Suction piping failed repeatedly in areas of weld joints, size reduction or enlargement, and directional change. Shop applied, baked-on phenolic coating satisfactorily reduced metal failures of pump discharge piping.

- Acid injection into raw seawater — Maximum use of plastic or plastic-lined components is needed in this area.

E. EFFECT OF DEAERATION AND pH CONTROL ON CORROSION RATES

The original purpose of the deaerator was to remove dissolved gases — mainly, oxygen and nitrogen — from the raw seawater feed so that carbon steel equipment could be used throughout the plant in order to reduce capital costs. The deaerator reduces the oxygen concentration of seawater from about 6 ppm to 0.1 ppm. If in-leakage of atmospheric oxygen is prvented by proper design and sealing materials, then a significant reduction in carbon steel equipment is achieved. This was effectively demonstrated at Freeport when two straight runs of pipe in the deaerated and nondeaerated sections of the plant were examined (OSW R&D Number 123, Figures 3.22 and 3.23, page 51). Considerable buildup of corrosion products was seen in the nondeaerated seawater feed piping while none was present in the piping carrying the deaerated feed.

Later, the deaerator also acted as the decarbonator for removing CO_2 from the scale-forming bicarbonates in SWF. Efficient removal of CO_2 in the deaerator reduces corrosion in the equipment downstream of the deaerator.

A second factor affecting corrosion in the deaerated SWF circuit is the addition of sulfuric acid to the raw seawater feed for scale control. At the deaerator inlet, pH is as low as 4.0 while at its outlet it is about 5.5 to 6.0, which is still on the acidic side of the pH scale. Initially, caustic soda was added to bring this to the neutral range. However, with the use of aluminum-brass tubes (instead of the original carbon steel) in the preheaters, the caustic neutralization step could be eliminated without any serious corrosion problem.

Addition of catalyzed sodium sulfite to the deaerated SWF is effective in scavenging the residual oxygen but does not result in measurable reduction in corrosion rates (Eighth Annual Report, page 87).

F. EFFECTIVENESS OF CLADDING AND PROTECTIVE COATINGS

Carbon steel waterboxes and tubesheets in the heat exchangers and evaporator effects showed rapid metal loss, and perforations were found in carbon steel pass partitions that were in contact with the seawater. Initially, only the non-deaerated seawater circuit equipment was alloy-clad with the expectation that once the seawater was deoxygenated its attack on carbon steel would be at a tolerable level. But the turbulence and high temperature conditions accelerated the carbon steel attack to such an extent that a few plant shutdowns were necessitated by the corrosion failures. As described earlier, monel and cupro-nickel cladding provided adequate protection to these carbon steel waterboxes and tubesheets.

Protective coatings are not as durable as the metal claddings but they reduce corrosion rates at low initial costs. "Brutem" coal-tar epoxy coating was effective in reducing the corrosion found in the deaerator tower interior and in the travelling screen housing. A baked-on phenolic coating (shop-applied) provided excellent protection for brine piping at temperatures up to 250°F. A hot air-dried catalyzed epoxy with a heat reflective layer worked well in Effects I, II, III, and V and in certain heat exchangers at temperatures up to 240°F when it was applied completely holiday-free.

For plant exterior painting, a chlorinated rubber-based paint was satisfactory. Other recommended systems are vinyl or epoxy applied over an inorganic based paint. Use of aluminum-based paints is not recommended (OSW R&D Report Number 123, page 140).

G. EFFECTIVENESS OF CATHODIC PROTECTION SYSTEM

The basic corrosion reaction can be written as:

at the anode $\mathrm{Fe} \rightarrow \mathrm{Fe}^{++} + 2e$

at the cathode $2\mathrm{H}^{+} + 2e \rightarrow \mathrm{H}_2$ (in acid media) and $O_2 + 2H_2 + 4e \rightarrow 4OH^-$ (in neutral or alkaline solutions)

or $O_2 + 4H^+ + 4e \rightarrow 2H_2O$ (in acid solutions)

Iron atoms go into solution to become iron ions while releasing electrons — this is the basic step in iron corrosion. Since this reaction takes place only at anodes, two methods of reducing it become evident:

- sacrificial anode method, and
- impressed-current method.

Both of these methods aim at making the metal to be protected a cathode in the electrolyte involved. Examples of the sacrificial-anode method include the use of zinc, magnesium, or aluminum as anodes in electrical contact with the metal to be protected — e.g., zinc anodes buried in ground to protect underground pipe lines, or aluminum anodes as attachments to ships' hulls. The current required is generated in this method by corrosion of the sacrificial-anode material. In the case of impressed-current method, the required direct current is provided by external sources (e.g., rectifiers) through a non-corroding anode such as platinum. The requirement with respect to current distribution and anode placement varies with the resistivity of soils or electrolyte. It is generally impractical, for instance, to use cathodic protection where dilute acids are the corroding solution or electrolyte.

At the Freeport Plant, the original design called for installation of DOW Type 24W coated magnesium sacrifical anodes in the non-deaerated seawater service in HX-212, -213, -214, -311 and -312. These anodes were installed in the top and bottom waterboxes of each exchanger. In the various annual reports or development reports, very little was mentioned regarding the experience with this sacrificial-anode system. However, in a meeting with Mr. D. I. Dykstra of OSW and Mr. D. D. Kays (previously) of Stearns-Roger, it was mentioned that the rate of the magnesium anode consumption was so high that the method was abandoned.

The Second Annual Report contains this remark (page III-12): "The cathodic protection installed to protect the travelling screen and end-of-pipe gate valves has proven to be eminently satisfactory."

The second method of cathodic protection, the impressed-current system, was installed during fiscal year 1965 in heat exchangers HX-201, -306, and

-310 to test the feasibility of this method for retarding the corrosion rate of the carbon steel waterboxes. The system included: a direct-current rectifier, platinum-coated anodes, silver-silver chloride reference electrodes, and associated control equipment. The anodes were immersed in the seawater flow at the upper and lower waterboxes of the above-mentioned exchangers. The rectifier maintained a positive potential on the platinum anode and a negative potential on the carbon steel waterbox wall. By proper adjustment of the potential difference between the electrode and the wall, the entire carbon steel wall surface can be made to act as a cathode and thus the dissolution of metal is prevented.

Cathodic protection was accomplished in HX-310 (at a temperatures of 135°F) with a current density of 0.5 to 0.6 ampere per square foot. (This current density is extremely high. Normally it is in the milliampere range.) In 18 months of operation, the carbon steel waterbox showed negligible corrosion rates. HX-306 (at 200°F) and HX-201 (at 265°F) required current densities of 0.8 to 1.0 ampere per square foot to achieve the desired reference electrode-waterbox potential difference of 0.85 volt. For the first six months, corrosion continued in these exchangers due to insufficient current densities. A larger capacity rectifier was installed in January 1966. Later reports by Stearns-Roger contain no information regarding this cathodic protection system.

Chapter 7

MAINTENANCE AND EQUIPMENT PERFORMANCE HISTORY

A. MAJOR EQUIPMENT SPECIFICATIONS

To better assist the reader in the proper evaluation of the maintenance activities and equipment performance at the Freeport Plant, a number of tables have been prepared and included in this subsection. Table 7.1 lists the original evaporator effectwise details. Table 7.2 lists similar details for the 17-effect evaporator. The 5-effect module was added to the original plant in FY-1967.

Table 7.3 contains design parameters for the preheat exchangers Series 200. Table 7.4 is for the vapor condensing preheaters Series 300 exchangers. For quick performance check, the overall heat transfer coefficients obtained for both these series of exchangers are listed along with the design values. The average coefficients were calculated from values reported for Development Runs 6 through 11 only.

Specifications for flash tanks and vessels, pumps and drivers are also given in Tables 7.5 and 7.6 respectively.

B. MAJOR EQUIPMENT MAINTENANCE

The annual reports were the source of information and data for the tabulation of maintenance experience given in Table 7.7. Often, a general statement of a maintenance problem would be given, such as "50% of the tubes in Effect M were scaled," or "Corroded at the turbulent sections of piping." Generally, these items of maintenance were counted as "one event." The comment given would be that a pump "failed." If it failed, then it must have been repaired, but the event was listed as "failed . Occasionally, the statement would

Table 7.1

Original Evaporator Specifications

Equip. No.	SERVICE	MATERIAL				Vessel Code	TUBES				SHELL		
		Vessel	Nozzles	Tube Sheets	Tubes		Number	O.D.	Gage	Length	Diameter	Design Temperature	Design Pressure
I	Evaporator Effect	CS/A258C	CS/A285C	CS/A201B	CS/Corten	ASME 1959 & TEMA C	490	2"	14	24 ft.	10'	300°F	75 PSIG
II	Evaporator Effect	CS/A285C	CS/A285C	CS/A201B	Admiralty	ASME 1959 & TEMA C	467	2"	16	24 ft.	10'	300°F	75 PSIG
III	Evaporator Effect	CS/A285C	CS/A285C	CS/A201B	AL-BR	ASME 1959 & TEMA C	467	2"	16	24 ft.	10'	300°F	60 PSIG
IV	Evaporator Effect	CS/A285C	CS/A285C	CS/A201B	90-10 Cu-Ni	ASME 1959 & TEMA C	383	2"	16	24 ft.	10'	300°F	60 PSIG
V	Evaporator Effect	CS/A285C	CS/A285C	CS/A201B	Carbon Steel	ASME 1959 & TEMA C	400	2"	14	24 ft.	10'	300°F	60 PSIG
VI	Evaporator Effect	CS/A285C	CS/A285C	CS/A201B	CS/Corten	ASME 1959 & TEMA C	535	2"	14	24 ft.	10'	300°F	60 PSIG
VII	Evaporator Effect	CS/A285C	CS/A285C	CS/A201B	AL-BR	ASME 1959 & TEMA C	425	2"	16	24 ft.	10'	300°F	60 PSIG
VIII	Evaporator Effect	CS/A285C	CS/A285C	CS/A285C	CS/Corten	ASME 1959 & TEMA C	490	2"	14	24 ft.	12'	300°F	15 PSIG
IX	Evaporator Effect	CS/A285C	CS/A285C	CS/A285C	Admiralty	ASME 1959 & TEMA C	467	2"	16	24 ft.	12'	300°F	15 PSIG
X	Evaporator Effect	CS/A285C	CS/A285C	CS/A285C	AL-BR	ASME 1959 & TEMA C	425	2"	16	24 ft.	14'	300°F	15 PSIG
XI	Evaporator Effect	CS/A285C	CS/A285C	CS/A285C	Carbon Steel	ASME 1959 & TEMA C	430	2"	14	24 ft.	16'	300°F	15 PSIG
XII	Evaporator Effect	CS/A285C	CS/A285C	CS/A285C	Admiralty	ASME 1959	390	2"	16	24 ft.	18'	300°F	15 PSIG

Table 7.2

Final Evaporator Specifications

Equip. No.	SERVICE	MATERIAL				Linings	TUBES				SHELL		
		Vessel	Nozzles	Tube Sheets	Tubes		Number	O.D.	Gage	Length	Diameter	Design Temperature	Design Pressure
I	Evaporator	CS A 285C	A285C	A201B	AL-BR,	Water box	310,5,5,4	2"	16,18,16,	24'	10'	300°F	75 PSIG
					304SS	90-10 Cu-Ni			.041"				
II	Evaporator	CS A 285C	A285C	A201B	306SS, Ti	Water box	291	2"	16	24'	10'	300°F	75 PSIG
					Admiralty	90-10 Cu-Ni							
III	Evaporator	CS A 285C	A285C	A201B	AL-BR	Water box	315	2"	16	24'	10'	300°F	60 PSIG
						90-10 Cu-Ni							
IV	Evaporator	CS A 285C	A285C	A201B	Cu-Ni, 90-	Water box	314	2"	16	24'	10'	300°F	60 PSIG
					10	1/4" CS +							
						epoxy							
V	Evaporator	CS A 285C	A285C	A201B	AL-BR		315	2"	16	24'	10'	300°F	60 PSIG
VI	Evaporator	CS A 285C	A285C	A201B	AL-BR	Water box	315	2"	16	24'	10'	300°F	60 PSIG
						1/4" CS +							
						epoxy							
VII	Evaporator	CS A 285C	A285C	A201B	AL-BR		316	2"	16	24'	10'	300°F	60 PSIG
VIII	Evaporator	CS A 285C	A285C	A201B	AL-BR	Sump mortar	291	2"	16	24'	12'	300°F	15 PSIG
IX	Evaporator	CS A 285C	A285C	A201B	Titanium	Sump mortar	290	2"	25	24'	12'	300°F	15 PSIG
X	Evaporator	CS A 285C	A285C	A201B	AL-BR		315	2"	16	24'	14'	300°F	15 PSIG
XI	Evaporator	CS A 285C	A285C	A201B	AL-BR		316	2"	16	24'	16'	300°F	15 PSIG
XII	Evaporator	CS A 285C	A285C	A201B	Cu-Ni 90-		282	2"	18	20'	18'	300°F	15 PSIG
					10								
XIII	5 Effect Modular Addition to												
	Evaporator Train	CS A 285C	A285C	SA516GR70	CDA 194		150	3 1/8"	16	11'	N/A	250°F	14.9 PSI &
				Cu B-152									Full Vac
				clad									
XIV	5 Effect Modular Addition to												
	Evaporator Train	CS A 285C	A285C	SA516GR70	316SS		360	2"	20	22'	N/A	250°F	14.9 PSI &
				Cu B-152									Full Vac
				clad									

Table 7.2 (cont'd)

Final Evaporator Specifications

Equip. No.	SERVICE	MATERIAL					TUBES				SHELL		
		Vessel	Nozzles	Tube Sheets	Tubes	Linings	Number	O.D.	Gage	Length	Diameter	Design Temperature	Design Pressure
XV	5 Effect Modular Addition to												
	Evaporator Train	CS A 285C	A285C	SA516GR70	AL-BR		366	2"	16	22'	N/A	250°F	14.9 PSI &
				Cu B-152									Full Vac
				clad									
XVI	5 Effect Modular Addtion to												
	Evaporator Train	CS A 285C	A285C	SA516GR70	Cu-Ni 90-		252	2 1/2"	18'	22'	N/A	250°F	14.9 PSI &
				Cu B-152	10								Full Vac
				clad									
XVII	5 Effect Modular Addition to												
	Evaporator Train	CS A 285C	A285C	SA516GR70	Cu-Ni 90-		223	3"	18'	22'	N/A	250°F	14.9 PSI &
				Cu B-152	10								Full Vac
				clad									

Table 7.3

Series 200 Heat Exchanger Specifications

Equip. No.	SERVICE	SHELL						TUBES							Design U	Average U	Outside Total heat transfer surface	Tube Velocity	No. of Passes Tubeside
		Diam.	Length	Material	Thickness (Inches)	DESIGN		Diam./Length	Gage	Pitch	No. of Tubes	Material	DESIGN						
						°F	PSIG						°F	PSIG					
#201	Seawater/1st Effect																		
	Condensate	24" ID	10'-	A285C	3/8	300	115/Full	1" OD/	14	1 1/4"	208	Steel	300	250	112	153	435.6	5 ft/sec	2
			3 1/2"				Vac	8'-0"											
#202	Seawater/2nd Effec																		
	Condensate	24" ID	10'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	178	Admiralty	300	250	227	*	372.8	5 ft/sec	2
			3 1/2"				Vac	8'-0"											
#203	Seawater/3rd Effect																		
	Condensate	24" ID	10'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4	178	AL-BR	300	250	209	*	372.8	5 ft/sec	2
			3 1/2"				Vac	8'-0"											
#204	Seawater/4th Effect																		
	Condensate	24" ID	10'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	223	90-10 Cu-	300	250	219	*	467.1	10 ft/sec	2
			3 1/2"				Vac	8'-0"				Ni							
#205	Seawater/5th Effect																		
	Condensate	24" ID	10'-	A285C	3/8	300	115/Full	1" OD/	14	1 1/4"	208	Steel	300	250	202	*	435.6	5 ft/sec	2
			3 1/2"				Vac	8'-0"											
#206	Seawater/6th Effect																		
	Condensate	24" ID	10'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	178	Admiralty	300	250	195	*	372.8	5 ft/sec	2
			3 1/2"				Vac	8'-0"											
#207	Seawater/7th Effect																		
	Condensate	24" ID	10'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	228	AL-BR	300	250	78	*	1193.8	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#208	Seawater/8th Effect																		
	Condensate	28" ID	22'-	A285C	3/8	300	115/Full	1" OD/	14	1 1/4"	224	Steel	300	250	191	216	1172.9	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#209	Seawater/9th Effect																		
	Condensate	28" ID	22'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	291	Admiralty	300	250	185	143	1523.7	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#210	Seawater/10th																		
	Effect Condensate	28" ID	22'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	311	AL-BR	300	250	177	157	1628.4	5 ft/sec	2
			3 1/2"				Vac	20'-0"											

* 'U' values for HX-202 through 207 not reported. These were removed early in plant life.

Table 7.3 (cont'd)

Series 200 Heat Exchanger Specifications

Equip. No.	SERVICE	SHELL						TUBES							Design U	Average U	Outside Total heat transfer surface	Tube Velocity	No. of Passes Tubeside
		Diam.	Length	Material	Thickness (Inches)	DESIGN °F	DESIGN PSIG	Diam./ Length	Gage	Pitch	No. of Tubes	Material	DESIGN °F	DESIGN PSIG					
#211	Seawater/11th																		
	Effect Condensate	32" ID	22'-	A285C	3/8	300	115/Full	1" OD/	14	1 1/4"	360	Steel	300	250	165	137	1884.9	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#212	Deaerator Feed/																		
	Condensate from																		
	#211	32" ID	22'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	355	90-10 Cu-	300	250	172	166	1858.8	10 ft/sec	2
			3 1/3"				Vac	20'-0"				Ni							
#213	Raw Seawater/12th																		
	Effect Condensate	32' ID	22'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	410	Admiralty	300	250	161	146	2146.8	10 ft/sec	2
			3 1/3"				Vac	20'-0"											
#214	Raw Seawater/Con-																		
	densate from #312	32" ID	22'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	440	AL-BR	300	250	168	133	2303.8	10 ft/sec	2
			3 1/3"				Vac	20'-0"											
#215	Raw Seawater/Pro-																		
	duct Water to																		
	Freeport	36" ID	22'-	A285C	3/8	300	115/Full	1" OD/	18	1 1/4"	580	AL-BR	300	250	142	65	3036.9	10 ft/sec	2
			3 1/3"				Vac	20'-0"											

Table 7.4

Series 300 Heat Exchanger Specifications

Equip. No.	SERVICE	SHELL						TUBES							Design U	Average U	Outside Total heat transfer surface	Tube Velocity	No. of Passes Tubeside
		Diam.	Length	Material	Thickness (Inches)	DESIGN °F	DESIGN PSIG	Diam./ Length	Gage	Pitch	No. of Tubes	Material	DESIGN °F	DESIGN PSIG					
#301	Seawater/Steam																		
	from 1st Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	14	1 1/4"	288	Steel	300	250	245	482	1508.	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#302	Seawater/Steam																		
	from 2nd Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	18	1 1/4"	265	Admiralty	300	250	261	421	1387.5	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#303	Seawater/Steam																		
	from 3rd Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	18	1 1/4"	274	AL-BR	300	250	257	524	1434.7	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#304	Seawater/Steam																		
	from 4th Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	18	1 1/4"	300	90-10 Cu-	300	250	267	368	1570.8	10 ft/sec	2
			3 1/2"				Vac	20'-0"				Ni							
#305	Seawater/Steam																		
	from 5th Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	14	1 1/4"	274	Steel	300	250	194	433	1706.9	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#306	Seawater/Steam																		
	from 6th Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	18	1 1/4"	326	Admiralty	300	250	247	338	1706.9	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#307	Seawater/Steam																		
	from 7th Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	18	1 1/4"	314	AL-BR	300	250	229	352	1644.1	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#308	Seawater/Steam																		
	from 8th Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	14	1 1/4"	343	Steel	300	250	206	193	1795.9	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#309	Seawater/Steam																		
	from 9th Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	18	1 1/4"	276	Admiralty	300	250	235	226	1445.1	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#310	Seawater/Steam																		
	from 10th Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	18	1 1/4"	332	AL-BR	300	250	217	160	1738.4	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#311	Seawater/Steam																		
	from 11th Effect	28" ID	22'-	A285C	3/8	300	90/Full	1" OD/	18	1 1/4"	307	AL-BR	300	250	227	60	1607.5	5 ft/sec	2
			3 1/2"				Vac	20'-0"											
#312	Seawater/Steam																		
	from 12th Effect	48" ID	22'-	A285C	7/16	300	Full Vac	1" OD/	18	1 1/4"	1000	AL-BR	300	35/	196	227	4188.8	8 ft/sec	4
			3 1/2"					16'-0"						Full Vac					

Table 7.5

Centrifugal Pump and Driver Specifications

Equip. No.	SERVICE	PUMP					MOTOR			
		GPM	TDH (ft.)	Impeller Material	°F	Case Material	RPM	Enclosure	Volts/ Frequency	HP
P1	Condensing Water	3500	22	1B Ni Resist.	88	Iron	1750	Open	440/60	
P2	Condensing Water	3500	22	1B Ni Resist.	88	Iron	1750	Open	440/60	
P3	Seawater Feed	955	55	1B Ni Resist.	88	Iron/Steel	1750	Open	440/60	
P4	Seawater Feed	855	275	1B Ni Resist.	134	Iron/Steel	1750	Open	440/60	
P5	Seawater Feed	855	275	1B Ni Resist.	134	Iron/Steel	1750	Open	440/60	
P11*	Blowdown I to II	850/210	45/65	1B Ni Resist.	251	Iron/Steel	1750	Open	440/60	
P12	Blowdown II to III	790/250	45/65	1B Ni Resist.	243	Iron/Steel	1750	Open	440/60	
P13	Blowdown III to IV	725/310	45/65	1B Ni Resist.	235	Iron/Steel	1750	Open	440/60	
P14	Blowdown IV to V	660/365	45/65	1B Ni Resist.	226	Iron/Steel	1750	Open	440/60	
P15	Blowdown V to VI	600/425	45/65	1B Ni Resist.	216	Iron/Steel	1750	Open	440/60	
P16	Blowdown VI to VII	550/480	45/65	1B Ni Resist.	208	Iron/Steel	1750	Open	440/60	
P17	Blowdown VII to VIII	480/550	45/65	1B Ni Resist.	198	Iron/Steel	1750	Open	440/60	
P18	Blowdown VIII to IX	425/600	45/65	1B Ni Resist.	185	Iron/Steel	1750	Open	440/60	
P19	Blowdown IX to X	365/660	45/65	1B Ni Resist.	172	Iron/Steel	1750	Open	440/60	
P20	Blowdown X to XI	310/725	45/65	1B Ni Resist.	157	Iron/Steel	1750	Open	440/60	
P21	Blowdown XI to XII	250/790	45/65	1B Ni Resist.	138	Iron/Steel	1750	Open	440/60	
P22	Blowdown XII to T-1	210/850	45/65	1B Ni Resist.	119	Iron/Steel	1750	Open	440/60	
P31	Condensate	47-61	70	Iron	261	Iron	1750	Open	440/60	
P32	Condensate	47-61	70	Iron	250	Iron	1750	Open	440/60	
P33	Condensate	47-61	70	Iron	242	Iron	1750	Open	440/60	
P34	Condensate	47-61	70	Iron	234	Iron	1750	Open	440/60	
P35	Condensate	47-61	70	Iron	225	Iron	1750	Open	440/60	
P36	Condensate	47-61	70	Iron	215	Iron	1750	Open	440/60	
P37	Condensate	47-61	70	Iron	206	Iron	1750	Open	440/60	
P38	Condensate	47-61	70	Iron	196	Iron	1750	Open	440/60	
P39	Condensate	47-61	70	Iron	183	Iron	1750	Open	440/60	
P40	Condensate	47-61	70	Iron	170	Iron	1750	Open	440/60	
P41	Condensate	47-61	70	Iron	154	Iron	1750	Open	440/60	
P42	Condensate	47-61	70	Iron	135	Iron	1750	Open	440/60	
P43	Condensate	47-61	70	Iron	115	Iron	1750	Open	440/60	
P44	Flashed Condensate	60-360	60	Iron	134	Iron	1750	Open	440/60	
P47	Gland & Purge Seawater	50	175							
P50	Product to DOW	425	190	Iron	100*	Iron	1750	Open	440/60	
P51	Product to Freeport	425	150	Iron	100	Iron	1750	Open	440/60	

* P11 through 22 have forward feed/backward feed data.

Table 7.6

Tanks and Vessels

Equip. No.	SERVICE	Height (feet)	Diameter (feet)	Capacity (gallons)	Pressure (PSIG)	Temperature °F	Material	Internals
FT #1	Flash Tank	9'1"	23 1/4" I.D.	----	76 & FV	300	A-283-GR C; A-7; A-181-GR I; A-105-GR II	----
FT #2	Flash Tank	9'1"	23 1/4 I.D.	----	61 & FV	300	A-283-GR C; A-7; A-181 GR I; A-105 GR II; A-106	A-7
Split Tank	Product Water Storage	27'6 5/16"	10'6" I.D.	15,000	----	----	A-283-GR C; A-7; A-181-GR I; A-105-GR II; A-53-GR B; A-307-GR B	----
Deaerator	Deaerator	46'7 3/4"	5'11" I.D.	440,000 lb/hr	15 & FV	150	A-285-GR C-FBX; A-7	Raschig rings
FT #3	Flash Tank	9'1" OA	23 1/4" I.D.	----	61 & FV	300	A-283-GR C; A-7; A-181 GR I; A-105 GR II; A-106	----
FT #4	Flash Tank	9'1" OA	23 1/4" I.D.	----	61 & FV	300	A-283-GR C; A-7; A-181-GR I; A-105 GR II; A-106	----
FT #5	Flash Tank	9'1" OA	23 1/4" I.D.	----	61 & FV	300	A-283-GR C; A-7; A-181 GR I; A-105 GR II; A-106	----
FT #6	Flash Tank	9'1" OA	23 1/4" I.D.	----	61 & FV	300	A-283-GR C; A-7; A-181-GR I; A-105-GR II; A-106	----
FT #7	Flash Tank	9'1" OA	23 1/4" I.D.	----	61 & FV	300	A-283-GR-C; A-7; A-181-GR I; A-105 GR II; A-106	----
FT #8	Flash Tank	9'5" OA	2'6" I.D.	----	16 & FV.	300	A-283-GR C;A-7; A-181-GR I;A-105-GR II; A-106	----

Table 7.6 (cont'd)

Tanks and Vessels

Equip. No.	SERVICE	Height (feet)	Diameter (feet)	Capacity (gallons)	Pressure (PSIG)	Temperature °F	Material	Internals
FT #9	Flash Tank +++	9'5" OA	2'6" I.D.	----	16 & FV	300	A-283-GR C;A-7; A-181-GR I; A-105-GR II; A-106	----
FT #10	Flash Tank	9'1" OA	3'0" I.D.	----	16 & FV	300	A-283-GR C;A-7; A-181-GR I; A-105-GR II; A-106	----
FT #11	Flash Tank	11'1" OA	4'0" I.D.	----	16 & FV	300	A-283-GR C; A-7; A-181-GR I; A-105-GR II; A-106	----
	Clarifier-Thickener	15'0"	35'0" I.D.	----	----	----	A-283-GR C; A-7	A-106-GR B; A-7

NOTE:
FV = Full vacuum
ID = Inside Diameter
OA = Overall

Table 7.7

Maintenance History of the Freeport Plant, 1961-1969

	FY-1962	FY-1963	FY-1964	FY-1965	FY-1966	FY-1967	FY-1968	FY-1969
Acid Tank	Drained, cleaned, and inspected.				Opened and cleaned - sludge removed.	Corrosion a problem.	Corrosion continued as a problem.	
Caustic Tank		Cleaned sediment from tank and strainer.						
Deaerator	Repaired level control valve. Repaired inlet nozzle.	Repaired and inspection. PVC section replaced by Saran lined pipe and orifices. Air leak. Locate and repair leaks. Opened-removed Raschig rings, innercoating replaced. Saran spray nozzles separated from steel pipe. Raschig rings broken. Replaced grating supporting Raschig rings. Mixer disassembled and repaired. Sandblast and coat interior. Mixer removed from service - leaking.	Opened, inspected and cleaned of silt in bottom. Epoxy coating moderately damaged in packed area.	Interior coating of epoxy needed repair. Packing material changed from Raschig rings to polypropylene. New grating and support installed.	Vacuum leak in instrument tap - August 1965.	Internal coating failed, replaced.	Internal coating failed, removed, and replaced.	Repaired upper 4 feet of epoxy coating, inside. Capped out of service all nozzles not used.
Electrical	Tripping of circuit breakers on overload. Shorted main feed to Pump P-4. Reconditioned equipment following Hurricane Carla. Switch contacts corroded. Motor Starter for P-5 failed. Clean & restore all switch gear. Motor for P-4 shorted.	Shorted P-5 motor. Control switches stick; continuous maintenance required.			Motors for P-50 & 51 completely overhauled. Lighting circuits rewired.		Several switch gear failures - overloads due to high ambient temperatures; control transformer insulation breakdown. Dust and moisture accumulation noted on switch gears - assumed to contribute to problem. Cleaned.	Several motors and starter transformers burned out. All starters removed, inspected, and cleaned. All bus bars, structure, and internal wiring cleaned. After above cleaning, 440 volt short circuit occurred on lower bus insulator. Considerable fire damage to switch gear. Entire plant wiring system reworked. Motor leads, switch gear to pump motors, failed. New wiring installed.
Evaporators	Scaling of Effect I. Tubes drilled in March 1962. Scaling in Effects I, II, III, IV, X & XII in June, 1961. Scaling (again) in Effect XII, XI & X, by June 30, 1961. Poor water distribution. Installed new distributor plates in Effects X, XI, & XII. Carbon-steel baffle installed in Effect XII. Reinsulation. Repainting. Cleaned evaporator tubes - I, X, XI & XII. Installed atmospheric vents on water box of I. Scaling - heavy in I and XII, progressively less toward center Effects. Corrosion pits on carbon-steel tube sheets, distributor plates, vapor body cones and brine discharge lines	Air tested all Effects. Distribution plates pickled and hot-dipped galvanized. Acid clean seawater side of I, V, VI, VII & XII and condensing side of I. Baffle plate installed in each effect. Scale in XII & I. Inspect bottoms of evaporators; minimal deposits I through X, 76 tubes of XI scaled; 80% tubes of XII scaled. Tubes drilled and cleaned. Sandblast and coat water box XII. Major leaks found in I and V. Retubed all carbon-steel tubes. Expansion joints installed in Effects I, V, VI and VIII. Air tested Effects I, VIII and XI. Cleaned water boxes and exterior of shells. Install glass gage on Effect I. Effect XII badly scaled - 75%. Effect XI scaling - about 50%. Very little scale in remaining Effects.	Scaling, Effects X, XI, XII, removed by cold sea water wash. Scaling a persistent problem. Corrosion erosion found in first three Effects due to hot seawater. Effects and piping need to be coated. Carbon steel replaced. Less corrosion/erosion from IV through IX. Few carbon steel plug failures. Poor brine distribution flow contributes to scaling. Retubed all evaporators originally tubed with carbon steel.	Carbon steel plugs were replaced in Effects I, II and V. Liners were installed in upper water boxes. Effects I through XI were hydrostatic tested.	Leaking tubes found by air testing - June, 1966. Water for liner of Effect IV replaced - Nov. 1965. Waterbox liner of V replaced as precaution - Nov. 1965. 110 leaks found in IX by air testing - April 1966.		Corite distributor plates broken several times in Effect 1 through V requiring extensive repairs. Hold down brackets installed. Water boxes relined with various materials.	Corrosion pits repaired in II, III and VI. Effect XII demister dislodged - tie down wires replaced. External corrosion in I & II. Pressure tap failures on IX, X and XI. Carbon steel liner, Effect IV, failed - replaced by carbon steel 285C liner - was later repaired.
Flash Tanks				External corrosion of a tank bottom.		Condensate level gages installed.	3 Tanks fail due to corrosion. 2 Tanks fail from inside - impingement corrosion.	External repair to FT-7. Internal and external corrosion repair on FT-8. External patch repair 18" wide by 50% of circumference.
Heat Exchangers - 200 Series	Leakage due to corrosion of carbon-steel. Reinsulation. Repainting. Leaks. Replace carbon-steel tube in HX-201. Flush condensate side, bottom tube sheet, of all exchangers. General inspection, via top beads, of all tubes and water boxes. Leaks in HX-205 - March 1962. Leak in HX-205 - April, 1962. Removed HX-205 from condensate system. Carbon-steel tubes of 208 and 211 replaced.	Air tested all exchangers. Marine life in water boxes HX-214 & 215. Slimy substance on seawater side HX-212, 213, and 214. Remove barnacles and pluggage HX-215. Steel plates of HX-202, 203, 204, 205, 206, and 207 were leak tested. Barnacles in inlet water box HX-215.	Corrosion in water box HX-201, protective coating applied. Some evidence of corrosion in HX-208 through 211. HX-209 had leak to air, repaired. HX-212, 213 and 214 required frequent cleaning. HX-215 accumulates shells and marine life, needs periodic cleaning. HX-202 through 207 bypassed to reduce maintenance.	Repair of both water boxes was major item of maintenance. Liners of 316L stainless steel, monel, and carbon-steel, epoxy coated were installed.	Leaks occurred in both upper and lower water boxes of units in deaerated seawater feed circuit.	Leaks common in carbon steel water boxes. New liners installed in water boxes. Heavy magnesium hydroxide buildup in HX-201 was removed. Minor maintenance - leaks.		Several carbon steel drain plug, upper tube sheets failed.
Heat Exchangers - 300 Series	Reinsulation, repainting, leaks. Leaking tubes in HX-301 plugged - April, 1962. Leaking in HX-301 - April, 1962 (note- 2nd time). Carbon-steel tubes of 301, 305 and 308, were replaced. Modified vents on HX-307, 308, 309 and 310.	Air tested all exchangers. Tube leak in HX-311. Leak in HX-312. Leaking tube in HX-311 replaced.	Water boxes and pass baffles in high temperature end are deteriorating. HX-302 shell failed. Welded patch on outside, put liner on inside.	Repair of both water boxes was major item of maintenance. 316L liner installed in water box HX-302. Other HXs were so lined. Carbon-steel, epoxy coated and monel liners installed.	Leaks occurred in upper and lower water boxes of units in deaerated seawater feed circuit. In HX-301, 302 and 303 for first time. Severe seawater leaks, tube-side to shell-side, HX-311. Admiralty tubes failed in tube roll area.		Water boxes relined - used alloy carbon-steel, 316L stainless, and monel. Three times tubes of HX-312 failed in June, 1966. HX bypassed to permit operation. HX retubed in September 1966. Hydrostatic pressure test of HX-308 & 309 found leaks in welds and liner. Liners repaired in 5 heat exchangers because of galvanicaction - Jan. 1968. Liner removed & repairs made in HX-303b, Feb., 1968. External vessels continued corroding and failed.	Repaired epoxy coating HX-318. Several failures of heat exchanger external steam chests. Replaced carbon steel gasket etaining strips of HX-311. Repaired tube sheet weld liner of top and bottom water boxes of HX-31. Plugged leaking tube of HX-310. Silicone rubber coating of inlet water box, HX-307, separated. Not replaced.
Instrumentation	Pneumatic system flushed with water after Hurricane Carla. Damage resulted from operation before becoming dry. Outside lines frozen. Clean and calibrate all pneumatic instruments.	Thermowells and nipples installed throughout pipe system. Flow orfice and thermowell installed in brine line from XII. Thermowell installed in steam line. Replace desicant in dryer. Repaired all leaks. Checked system. Replaced transfer valve. Corrosion of seawater feed valve body with brass.	Normal cleaning of conductivity cells. Normal maintenance on automatic level controls. Evaporator level control requires purge cleaning and gear lubrication. Valve stem - gear drive is worn and needs replacement. Original thermowells required modification (deepen) to improve usefulness. Pressure gage calibration frequency varies with environment: more often for salt water service than fresh water. Two instrument air compressor required much maintenance: rod and ring replacements, valve overhauls. Complete overhaul needed soon.	New and longer rack gears installed in evaporator level control valves. Pistons and valves changed twice a year on air compressors. Crankshaft bearing changed on one compressor.	Routine maintenance. Conductivity cell sensor heads replaced.		Frequence of calibration increased for accuracy. Routine maintenance.	Metering pumps required considerable maintenance: motor bearings to stroke linkage adjustment. Frequent maintenance, cleaning and calibrating, of instrumentation. Frequent installation of special instrumentation, hence increased maintenance effort. High oil and water content of instrument air caused extra maintenance effort. Brine level controls of Effects I - XII required extra maintenance. Mainsteam control valve and pressure reducing valve overhauled. Instrument air compressors required considerable attention. Number 3 instrument air compressor failed - stripped threads on compressor block, low pressure cylinder,

Table 7.7 (cont'd)

	FY-1962	FY-1963	FY-1964	FY-1965	FY-1966	FY-1967	FY-1968	FY-1969
Instrumentation (cont'd)								resulting in broken piston rod, rings, and a bent crank shaft. Compressor repaired.
Piping & Tubing	Corrosion of carbon-steel. New seawater feed line to Freeport product water cooler thence to HX-214. Fresh water cooling line from Freeport cooler to barometric condenser. Line from barometric condenser discharge to suction of P-44 to put fresh water cooling into split tank. Vent line bypass of Effect XII. Reinsulation. Repainting. Pipes & fittings froze. Line leaks.	Hot well/seawater pit piping reconditioned and coated. Slimy substance on seawater piping up to deaerator. Cross-connected the acid and caustic tanks. Piping from P-11, 12 and 13 replaced with phenolic coated sections. Weld failures, Effects I through VIII, on discharge line down stream from control valve. Failure in cooling water line from vent condenser. Failure in seawater line, HX-201. Corrosion failure drain line from HX-304.	New piping installed to move brine discharge from point of seawater intake. Repair and replacement of seawater and brine piping is greatest single maintenance item. Major points are: suction piping at brine pumps P-11 through P-17; discharge piping at brine pumps P-14 through P-19; seawater feed piping from HX-301 to Effect I; seawater feed piping from HX-212 to deaerator. Piping failures were adjacent to weld seams or on the bottom of the pipe line. Cnrbon steel piping carrying acid entirely replaced twice.	Repair and replacement - largest single maintenance item. Failures in brine pump discharge pipe sections; failure a function of temperature. Failures in suction pipe section of brine pumps. Brine piping between HX-201 and Effect 1 failed. Failures in reducing elbows and horizontal piping adjacent to HX-215. Cemented joints, brine piping between Effects I & II leaked. Failures in condensate piping; a function of temperature. Pipe sizes increased to reduce flow bottle necks.	Carbon steel pipe failures largest single item. Carbon steel good for condensate and acid (H_2SO_4) but last only 5 years carrying seawater or blowdown at temperatures significantly above ambient. Numerous leaks in seawater feed piping; a function of temperature. Welds and elbows predominate. Numerous leaks in pipe sections of brine pump discharge; a function of temperatures. Welds and elbows predominate. Few and minor failures in condensate piping; usually in threaded areas of unions, valves, and nozzles. Acid feed line was replaced. Life span 18 - 24 months.	Numerous carbon-steel pipe failures; replaced	Cooling line, HX-318, ruptured upon startup - high internal pressure. Random failures in carbon steel pipe in brine pumps suction and discharge sections.	Carbon steel hot brine piping continued to fail at elbows, reducers, control valves, etc. Carbon steel piping in hot end of plant, generally replaced with glass reinforced plastic pipe. Some glass reinforced fittings failed due to supporting and fabrication problems. External corrosion caused replacement of galvanized instrument and utility air piping. Reworked and repaired major pipe support hangers.
Pumps - General							Maintenance continued generally high due to failures of casings, suction covers, and impellers.	Mechanical seals require frequent attention; not recommended. Causes - seal ring corrosion and plugging of seal water system. Corrosion of carbon steel suction covers continues as a problem.
Pumps - P-1 & P-2, Circulating Water	Leaks caused by pitting of carbon-steel. Cavitation damage in second stage propellers and bowls. Improper propeller pitch. Install suction bells. Install wafer-type gate valves. P-2 disassembled and repaired - sleeve bearing, impeller, bowl casting pitted, shaft worn, and corrosion.	Repaired and inspected. Disassembled P-1: propeller cavitation, pitted of bowl, bowl relined. Micarta bearings and bronze impellers installed in P-2. Bowls coated.	Improper vertical alignment due to bearing protruding through bowl housing of P-2. Diffuser cone worked loose during operation of P-2. Sever shaft vibrations due to broken blade; bronze impellers installed in P-2. Overhauled in Oct., 1963 - metallizing & machining shaft, fabrication and installation of stainless steel shaft coupling and sleeve, repair casing, epoxy coating, install cutlass bearings. P-2 replaced by P-2a in Dec., 1963. In 3 months, propeller blade broke. Two new bronze propellers installed.	New stainless steel shaft installed in P-1. New cutlass bearing installed in P-1. Second stage impeller slipped, damaged impeller and bowl. Second stage impeller, P-2, worn apparently due to cavitation. Bottom flange corroeded after coating failure.	Carbon-steel shaft corroded, replaced - Oct., 1965. Bronze propeller deteriorated - six months use - Oct., 1965. Monel propeller failed - Feb., 1966 (2nd occasion).	P-2 converted to use stainless steel shaft and impeller.	Replaced shaft, shaftsleeves, stuffing box bearings, and four shaft marine type bearings in Aug., 1967. Repaired discharge transition section. Pump shaft metalized. Motor reepoxyed - above in Aug., 1967. New transition section and new shaft bearings in Feb., 1968. New Impeller in March 1968. New bearings in March 1968. Epoxy coated impeller housing discharge transition and elbow. Replaced elblow and propeller locking collar.	P-1 propeller failed - blade broke and damaged bowl assembly. Replaced propeller, shaft, propeller locking collar, and all marine type bearings.
Pumps - P-3, Deaerator Feed	Impeller failures - 2 - due to foreign object. Shaft failure - due to foreign object.	Repaired and inspected. Impeller loosened - new impeller and shaft. New bearings and seals.	Replaced inboard and outboard bearings. Epoxy coat suction bell.	Bronze impeller installed - Sept., 1964. Bearings replaced. Seal cage corroded away - Nov., 1964. Crack in in suction cover and discharge flange repaired - Nov., 1964. Maximum size bronze impeller installed - Jan., 1965.	Replaced after 5 years' service - not repairable.	Shaft broke - complete overhaul.		Replaced shaft, bearings, and seals. Built up impeller with bronze spray metal. Impeller vibrated - either out of balance or not center line bored. Shaft failed again, due to improper fitting of impeller to shaft. Installed new shaft, reconditioned and balanced impeller, and installed new bearings. Pump lost performance. Impeller undersize and damaged. Installed new, larger impeller.
Pumps - P-4 & P-5, Seawater Feed	Cavitation caused impeller and shaft failures. Impeller failure. Cracked impeller. Broken shaft. Install new impeller and shaft.	Repaired and inspected. Replaced impeller by one of original specifications. Shaft, stainless steel, broke. P-5 casing coated with epoxy. Replaced P-5 with P-5a, a centrifugal with a vertical pump. P-4 is to be standby.	Excessive vibration after 6 months to P-5a. Wear in to bowl bearing. Defects found in Ni-resist bowls and suction bell. Shaft sleeve worn at packing. Impeller shaft extensively worn at top bowl bearing - less at other 3 bearings. First stage impeller showed some cavitation damage. Graphite bearings show impact damage and were replaced. (Page 130-131.) Snap rings failed, sleeve rode shaft. Installed set screws to correct (Page 45). Slight vibration required shut down. On disassembly, bowl cap screws missing, top casing cracked, middle bowl casing repaired before original assembly, suction bell graphitized, and a chipped graphitar bearing. Pump returned to manufacturer for repair.	For P-5a: Two new stage bowls; Teflon line shaft bearing; 316 stainless steel bowl bolts; graphitar bearings; shaft bearing arears and stuffing boxes ceramic coated.	Numerous shaft and impeller failures - P-4. P-5a failed in Sept., 1965 (see FY-1965) bearings and bonding of ceramic coating failed. P-5a failed in Jan., 1966 for same reason-3rd failure. P-5a failed in May, 1966 for same reason plus two bowls cracked, bearing housings enlarged, and severe corrosion.	P5a: Graphitar bearings and ceramic coating failed July 1966. Similar failure in Aug., 1966. Ni-resist bowl cracked in Nov., 1966. P-4 rebuilt as standby for P-5a.	P-5a replaced with stainless steel pump. P-4 finally worn out - epoxy lining to bearing housing failed, causing shaft misalignment and breakage. Impeller housing failed. New double-volute pump installed in March 1968. Packing gland cored out rapdily on all pumps. Various packing materials tried. Shafts and packing sleeves were rebuilt and sized using Rokide.	P-5a shaft failure; repaired with Rokide coating. Marine type bearings failed; replaced. Repacked gland with Teflon. Serious pitted occurred under epoxy coating on lower column, coating removed.
Pumps, P-11 through P-22, Brine	Cavitation, slight damage to impeller leading edge. Impellers worked loose on 3 occasions. Installed new propellers in Pumps P-17 through P-22, to increase capacities. Iron oxide flake build up in suction of P-21. Replace bearings in P-17. Loose impeller on P-13. Suction lines removed and inspected. Repacked all pumps.	Repaired and inspected. Pitting in discharge piping - P-11 through P-17. Nozzles and valves on each pump for a standby. Flow from P-17 interrupted. Replace bearing on P-18. Air leak in P-15. Replaced bear and seals in P-16. Install rubber seats on P-11, 12, and 13. New bearings and seals on P-12 & 16. Installed maximum size propellers on P-17 and 18.	Mainly bearing replacements and repack stuffing box. Maximum size impellers installed to increase capacity. Cavitation a problem. Has eroded a suction bell. Repaired.	Bearing replacement (nominal). Suction covers corroded.	Ni-resist impeller cracked and loose - April, 1966. Mechanical seal leakage - Nov., 1965. Mechanical seal leakage occurred in June, 1966 - replaced carbon seal rings. Yokes of 2 pumps cracked - May 1966. Maximum size impeller installed - May 1966. Suction cover leak, cover replaced - Jan., 1966. Pump overhaul suction cover corroded - July, 1965. Suction cover corroded - July, 1966.			Normal maintenance.
Pumps, P-17 through P-22, Condensate				Nominal maintenance.				Normal maintenance.
Pumps, P-31 through P-43, Condensate	Graphitization of cast iron impellers. (Use of P-32 through P-37 discontinued when HX-202 - 207 were bypassed.) Overload failure. Motor	Repaired and inspected.	Replaced in board and out board bearings on P-40. Replaced impeller on P-41.	Four new impellers; some cavitation.	Motor hold down bolts rusted off, replaced. Cast iron impellers, 3 pumps, deteriorated. Replaced with maximum sized bronze impellers.			

Table 7.7 (cont'd)

	FY-1962	FY-1963	FY-1964	FY-1965	FY-1966	FY-1967	FY-1968	FY-1969
Pumps, P-31 through P-43, Condensate (cont'd)	bearings on P-40 failed. Installed couplings on discharge line of all pumps. Repacked all pumps.							
Pump, P-44			Cavitation ruined impeller, replaced.	Some cavitation damage. Capacity increased, larger motor and impeller.				
Pump, P-45, Caustic Feed	Stroke adjustment mechanism failure.	New pump.			Completely overhauled – first time in two years.			
Pumps, P-46 & P-47, Acid Feed	Plunger failure. New plungers. Repacked.	New pump. Cleaned. Installed new plunger. Teflon seat repaired.	Suction and discharge valves would freeze up. Cleaning of liquid-end of pump required.		Replaced diaphram in P-46.	Minor maintenance, mainly plugging of suction strainer and check valves.		
Pumps, P-50 & P-51, Condensate			Occasional replacement of packing.					
Pump, P-53				Stuffing box corroded (same as P-3).				Ni-resist impeller failed, replaced. Repaired shaft. Suction on foot valve failed, replaced.
Split Tank			Protective coating failed, walls rusting.		Liner, Freeport side failed in Nov., 1965.	Corrosion a problem. Coating failed and was replaced.	Corrosion a problem.	
Travelling Screen				Minimum maintenance required.				
Valves	Keystone valve linings had deteriorated, 80 valves were modified. Valves froze. Brine line valve liners replaced. Iron parts of sea water pit valves corroded.	Hotwell/seawater pit gates valves repaired or replaced. Install disc on level control, Effect VIII. Replace steel body of level control valve. Replace gate valve with butterfly on P-5. Butterfly valve maintenance: seats on P-31; disc on P-21, and new shaft. Replaced bottom blindhead of steam pressure reducing valve. Replaced plug type valve with butterfly on Pump P-3.		Gates valves, intake pit, required frequent inspection. One broken.	Condensate control valve body corroded; replaced. Flash tank control valve disc stuck – failed to operate. Cleaned and lubricated. Minor steam valve maintenance. Replaced gate valves with globe valves. First overhaul of desuperheated steam flow control valve – April, 1966.			Failed rubber seats and disc liners in control valves.
Miscellaneous	Gland seal lines froze. Gland seal pump failed. Heat tracing. Insulating of all impulse lines. Major corrosion of components exposed to non-deaerated seawater. Major damage from Hurricane Carla generally not included.	Seawater pit drained and cleaned. Low seawater feed flow due to piece of broken impeller. Replaced lighting transformer. Removal of large jelly fish from seawater pit. Inspect and repair seawater feed lines,	Seawater intake pit requires cleaning annually to prevent silt accumulation throughout plant. Freeze protection equipment installed. Clarifier-thickener converted to a pre-settling basin. Vacuum pump vibrates, sheared foundation tie-bars. Repaired. Chlorine solution from HX-215 corrodes butterfly valve, requiring excessive maintenance.		Cleaned yearly. Silt removed. Vacuum equipment: barometric condenser leaked and repaired. Suction and discharge valves replaced. Steam feet ejectors and barometric condenser replaced for larger capacity. Minor maintenance on Utility Air System. Instrument Air System – extensive compressor maintenance; yearly replacement of high – and low pressure cylinder assemblies; air drier system repaired.	External corrosion continued as a problem.	Corrosion continued a problem – air compressor. Extensive maintenance on steam jet ejectors. Diffuser was eroded – replaced and steam received was over design pressure and temperature. Replaced suction and discharge valves on mechanical vacuum pump. Air Drier required abnormal maintenance.	Internal corrosion on muffler on vacuum pump, repaired. Low performance, noisy operation of vacuum pump. Large accumulation of silt and Scale in four suction valves. Oil silt accumulated in discharge valves of vacuum pump. Vacuum pump (above) repaired. Needed: Adequate means to remove moisture and debris from line ahead of vacuum pump suction. R&D Maintenance: instrument installation and calibration, sample coil fabrication, sample tube pulling, sample taking, hydrostatic testing, etc. Sandblasting and painting continued. Replacement and renewal of gratings, stairway treads, rails, etc., of general plant maintenance. New and replacement insulation was installed. Entire plant hydrotested in May and June, 1969. All leaks found were repaired. Selected items were thickness tested.

be, in effect, "pump repaired, xxx, xxx, xxx, etc." listing several components used in the repair. This item was listed under "Pumps: — Repaired" and if shaft, bearing and/or impeller or propeller were mentioned, this event was counted under the listed components. For these reasons the tabulations of maintenance on the plant as discussed in the annual reports cannot be taken as "accurate": there were too many generalized statements of maintenance actions to permit a reasonable degree of accurate tabulation of the work.

C. RECURRING PROBLEMS

The maintenance problems experienced at Freeport seemed to recur, in many instances, several times. A list of the maintenance events occurring more often than 10 times is given below.

Problem	Occurrences
Corrosion	63
Pumps: repair/overhaul/replacement	48
Air tests	46
Leaks (air, hydro)	45
Pumps: propeller-impeller broken	44
Pumps: bearings	27
Coatings (epoxy)	26
Heat exchangers — liners	23
Pumps: shaft	23
Pipe: replacement/repair	22
Evaporators: scale formation	21
Heat exchangers: cleaned	19
Heat exchangers: inspection	18
Valves: repair/replacement	15
Pipe: failed	13
Baffle plates: installed	13
Evaporators: inspection	12
Evaporators: cleaned	11
Total	489

Rearranging the maintenance events mainly be principal item of equipment, but including the occurrence of problems involving corrosion, leaks, and coating, yields the following list:

Problem	Occurrences
Pumps	162
Evaporators and scale	88
Heat exchangers	86
Corrosion	63
Tests (air, hydro)	54
Pipes, fittings	52
Leaks	45
Electrical	28
Instrumentation	27
Coatings	26
Valves	20
Deaerator	7
Pneumatic system	4
Miscellaneous	31
Total	693

The maintenance events attributable to corrosion, leaks and coatings could have involved several major equipment items. For example, corrosion can cause leakage in the evaporator, pumps, piping, or heat exchangers. Scale formation, generally, occurred in Effects I, II, X, XI and XII (or XVII and XII as the case may be) and was included with the maintenance of evaporators.

D. SUMMARY OF EQUIPMENT PERFORMANCE

Table 7.7 identifies a total of 86 problem areas in which at least one maintenance event occurred during the eight years of operation. Seven of these problem areas accounted for 79% of the maintenance effort. These areas are: pumps, evaporators (including scaling), heat exchangers, corrosion, tests (air and hydrostatic), pipes and fittings, and leaks. Hence, these areas are logical points of special consideration in the design, development and operation of new LTV plants.

During the review of the annual reports, it was noticed that occasionally a comment was made on a specifically troublesome maintenance area. In FY-1964 it was noted that, "Repair and replacement of seawater and brine piping was the greatest, single maintenance item." In FY-1963, sticky control switches seemed to plague the plant operations; however, this problem was recorded only in this fiscal year. Carbon steel was a material that presented many problems (corrosions and leaks) and was noted as the item requiring the most maintenance during one year. Another maintenance problem was presented by the mechanical seal that was used. The comment specifically rejected this seal for use in future plants. Two comments indicated that a considerable maintenance effort was being expended in two problem areas — leaks (110 repaired at one time) and valve modifications (80 repaired). Each of these instances of multiple repair was listed as "one" maintenance event.

Inasmuch as the Freeport Plant was the first of a kind or a demonstration plant, much of the maintenance effort may be equated to a learning process. It is now known, for example, how to operate an LTV to minimize, if not eliminate, scaling in the various effects. Also, much of the scale that occurs can be washed out before it becomes a serious maintenance problem, so that in a true production plant, maintenance due to scaling of the evaporators should be almost, if not entirely, non-existent. Therefore, the operation of the Freeport Demonstration Plant will have made it possible to operate a full-scale production plant in such a way as to minimize the maintenance required.

Chapter 8

OPERATING HISTORY

A. OPERATING PROCEDURES

Every operating facility has a standard operating procedure (SOP). The objective of such procedures are maximum efficiency at minimum cost, and the elimination of danger to operating personnel and equipment. Preparation of an SOP is a combination of experience and understanding of the man-machine interface. A good SOP just does not happen; it is developed over a period of time, and it is always kept current.

The construction contractor prepared an SOP for the initial start-up of the Freeport Plant (see Chapter 3). It is reasonable to assume this manual was used as the guide to plant operations until the operating contractor (Stearns-Roger Corporation) assumed responsibility. Quite probably these procedures were observed for some time afterward by the operating contractor while gaining experience with the plant. Certainly the construction contractor's manual of Operating Instructions was the point of departure when the operating contractor began preparing his own SOP. In June 1961, a comprehensive "Start-Up and Operating Procedure" was submitted by the operating contractor (First Annual Report, page II-1).

Minor changes in procedures occurred frequently in the time period July 1961 to July 1967. In this period 12 Development Runs were conducted, each one having its own objectives. Hence, some variations were necessary in operating procedures. For example, the first, or at least very early, change was the abandonment of the magnesium hydroxide scale prevention technique and the initiation of acid injection process. A second procedural change arose with the introduction of flash tanks and removal of HX-202 through HX-207. Procedural variations were indicated when changing equipment and chemicals, or upon the introduction of the chlorine shock treatment to control marine growth.

A significant change in operating procedures occurred concurrently with the installation of the 5-effect module increasing the plant to 17 effects. In anticipation of this event, the operating contractor updated and expanded the operating procedures and the operator training program. The manual entitled "17-Effect Freeport Test Bed Center Plant Description and Operating Procedures" was prepared and issued during the summer of 1967. This manual was used until updated early in 1969.

In January 1969, the operating contractor rewrote the operating manual, and expanded the scope extensively by including such separate procedures as the "Chlorination Procedure" under a general heading of "Preventive Maintenance". Certain parts of this manual were revised in April 1969. Inasmuch as the plant operation was termined early in May 1969, this manual with these revisions was the last manual of operating procedures used at Freeport.

The outline of the final manual, "Operating Manual, 17-Effect VTE Falling Film Evaporator, Freeport Test Facility," is:

I. Introduction

II. General Plant Operation
  A. Process Description
  B. Changes in Plant
  C. Factors Governing Performance

III. Plant Operation
  A. General Control
  B. Detailed Control

IV. Start-Up and Shutdown Procedures
  A. Start-Up
  B. Shutdown (fresh water)

V. Trouble-shooting
  A. Seawater and Feed and Brine Cycle
  B. Steam Cycle
  C. Condensate and Product Cycle

VI. 17-Effect Equipment Designations

VII. Panel Mounted Instrumentation Descriptions

VIII. Preventive Maintenance
- A. Freeze Protection
- B. Air Compressor and Vacuum Pump Oil Change
- C. Pump Greasing
- D. Chlorination
- E. Acid Tank Drier
- F. Instrument Air Drier Desiccant Change
- G. Valve Stem Greasing
- H. Instrument Calibration
- I. Instrument Air Blowdown Traps
- J. HX-318 and Intake/Discharge Pit Valve Switching

X. Appendix
Includes individual chemical procedures, various tables and calibration charts, and diagrams.

B. EMERGENCY PROCEDURES

Very little on the subject of emergency procedures was found among the available documents and paper pertaining to the Freeport Plant. The First Annual Report, page II-5, states:

> "On September 18, the Plant was started up under emergency conditions...Emergency operating procedures were adopted since very little instrumentation (automatic) was functional...the Plant's equipment was manually operated and controlled."

This start-up occurred 5 days after Hurricane Carla.

The CB&I manual, "Operating Instructions,...", contains a section entitled "Emergency Shutdown — a. Start-up from Emergency Shutdown; b. Pump Failures." The quotation in the preceding paragraph, at least in part, refers to this section of the CB&I manual.

The report, "Summary of Performance Sea Water Desalting Demonstration Plant Number 1, Freeport, Texas," page 4-2, paragraph 406, states that a freeze protection procedure had been developed for operator use as a result of operating through the winter of 1961-62.

The "Operating Manual, 17-Effect VTE Falling-Film Evaporator, Office of Saline Water, Freeport Test Facility" contains instructions for freeze protection, including a chekclist of steps the operator is expected to take. The freeze protection is to be initiated "approximately one month before freezing temperatures are expected...." This manual omitted the section in the CB&I document relating to operation in the event of failure of certain pumps.

C. PLANT OPERATING REQUIREMENTS

The operating requirements for the Freeport Demonstration Plant have been divided into production requirements and facility services and protection. The former category includes those requirements necessary for production; the latter includes those necessary for servicing the facility.

The production requirements given below are, in general, average for the development runs completed prior to final shutdown. Inasmuch as the plant sometimes was operated at less than design capacity, the values below are believed to be on the conservative side.

The nominal capacity of the Freeport Plant was 1,000,000 gallons per day which is approximately 350,000 pounds of water per hour. The fresh water output from this plant during the Development Runs varied from an average low of 278,000 pounds per hour to an average high of 403,000 pounds per hour. The requirements to obtain this indicated production are given below.

PRODUCTION REQUIREMENTS

Seawater:

  Input — 416,000 to 522,000 lbs/hr
  Concentration factor — 0.745 to 1.02
  Temperature, °F — 55 to 90

Utilities:

  Electricity — 341 - 525 Kw connected operating load
  Steam to Effect I
    Input — 24,650 - 38,200 lb/hr
    Temperature, °F — 260 - 285
    Pressure, psig — 20 - 30

Chemicals (averaged over the Development Runs):

H_2SO_4	—	49 - 76 lb/hr
Polyphosphate	—	0.74 - 3.5 lb/hr
Caustic	—	0.18 - 7.5 lb/hr
Sulfite	—	0.58 lb/hr
Chlorine	—	4 - 10 lb/hr; 2 hour injection period/week
Silicate	—	4 lb/hr approximately

Personnel:

Varied from 13 to 20. (See subsection D, Manpower Levels.)

Fresh Water:

Fresh water was required for start-up (pumped to various seals) and for the fresh water shutdown. If this water was not available from the "Split Tank", it was taken from the City of Freeport.

The existence of a facility means certain requirements for services and protection. These requirements are merely listed for identification.

- Sewage
- Telephones
- Fresh water for personnel and laboratory
- Fire protection
- Hurricane protection
- Freeze protection

D. MANPOWER LEVELS

The Freeport Demonstration Plant was staffed for operation by the operating contractor in April 1961. This initial crew totaled 16 people as follows:

- Plant Superintendent
- Plant Engineer

- Plant Operators — 2 men per shift, 4 shifts
- Accountant
- Maintenance Technician
- Instrumentation Technician
- Operator-Technician
- Chemist
- Janitor

The responsibilities of the staff were to operate and to maintenance the plant, to evaluate and recommend improvements in the plant process, equipment and materials, and to make detailed cost evaluations.

Experience in plant operations and the passage of time inevitably lead to changes in work loads and assigned responsibilities. These changes are reflected in the number, titles and functions of plant personnel. Table 8.1 depicts the personnel changes from 1961 through FY-1969. The number of personnel on the plant crew varied from a low of 13 in FYs-1964 and 1965 to a maximum of 20 in FYs-1968 and 1969. The annual reports for these latter years reflect the growth of research and development efforts undertaken at the Freport Plant and the depth to which this work was pursued.

An important operational change was made in FY-1964, when the plant control was automated (Third Annual Report, page 16). This change permitted a 50% reduction in operator personnel, from two men per shift to 1 man. Automated control proved to be very successful. A corollary to this operational change was the need to train, and the reliance upon, the maintenance personnel for the relief operators. Accordingly, these individuals assumed a secondary function as required.

During FYs-1962, 1963, and 1964, the Plant Superintendent, with the Plant Engineer in a "staff" capacity supervised both operational and maintenance functions in the plant. In FY-1965, the position of "Operations Supervisor and Chemist" was authorized. The following year, FY-1966, the position of "Maintenance Supervisor" was established. This action was the first step towards a larger operating crew.

In FY-1967, as may be seen in Table 8.1, the position of a full-time chemist was reestablished, the Operations Supervisor made full-time, a Development Engineer was added, and the position of Plant Engineer abolished. Among the

Table 8.1

Plant Organization — 1961-1969

	ANNUAL REPORTS: 1	2	3	4[1]	5[1]	6	7	8[1]
	FISCAL YEARS: 1962	1963	1964	1965	1966	1967	1968	1969
POSITION:								
Superintendent	1	1	1	1	1	1	1	1
Plant Engineer	1	1	1	1	1			
Development Engineer						open	1	1
Chemist	1		1	1/2	1/2	1	1	1
Process Engineer								1
Operations Supervisor				1/2	1/2	1	1	1/2
Maintenance Supervisor					1	open	1	1/2
Accountant	1	1	1	1	1	1	1	1
Stenographer/Receptionist							1	1
HOURLY-PAID:								
Plant Operators	8	8	4	4	4	4	4	4
Technicians:								
Maintenance	1							
Instrument	1							1
Operator-Technician	1							
Operator-Instrumentation		1	1	1	1	1	1	
Maintenance Clerk							1	1
Operator-Chemist		1						
Utility			1	1	1	4	3	
Operator-Welder			1	1	1	1	2	
Operator-Maintenance			2	2	2	2	1	2
Operator-Millwright							1	
Welder								2
Millwright								1
General Maintenance								2
Janitor	1	1						
NUMBER POSITIONS	16	14	13	13	14	18	20	20
POSITIONS FILLED	16	14	13	13	14	16	20	20

NOTES: (1) Annual Reports Nos. 4, 5, and 8: Positions marked "1/2" indicate these functions performed by one individual.

hourly-paid employees, the number of Utility Technicians was increased from one one to four. The total number of crew positions changed from 14 to 18, although only 16 employees were actually on the payroll in FY-1967.

Also in FY-1967, a shift of responsibilities was made between the Plant and the Contractor's Home Office. The Plant Superintendent became responsible for all in-plant activities, including plant engineering, development, data collection, operations, maintenance, modifications, and employee relations. The Home Office would provide overall direction and planning, design engineering and technical evaluations.

The authorized position/personnel changes in FY-1968 were the addition of a Stenographer/Receptionist and a Maintenance Cler, the latter to provide assistance in warehousing, purchasing, job accumulation cost, and machinery documentation, as well as clerical functions.

The final organization chart for the Freeport Plant is shown in Figure 8.1. This chart was taken from the Eighth Annual Report.

E. COMPARISON OF MANUAL AND AUTOMATIC CONTROL

Initial construction of the Freeport Plant anticipated manual control by plant operators wherever a fluid level was critical, e.g., in the cone of each evaporator. Two operators per shift were provided. It was possible to maintain control by this operational method for nearly a year.

Prior to January 1962, an automatic level controller was installed in Effect XII. The operating success of this controller led to the installation of a similar device on Effect XI, for the purpose of observing the action of two automatic controllers in series. The result of this test clearly indicated that more automation could be utilized in this plant.

By the end of FY-1962, automatic level controllers had been installed on Effects I through X, thus completing the installation of these devices. Incorporation of these controllers into the routine operation was not accomplished until FY-1963. This modification to the Freeport Plant proved to be extremely valuable to the overall operations. Although the operators had become skilled in maintaining the brine levels in the various evaporators, that control method had been very time-consuming and had restricted the use of these operators for other tasks.

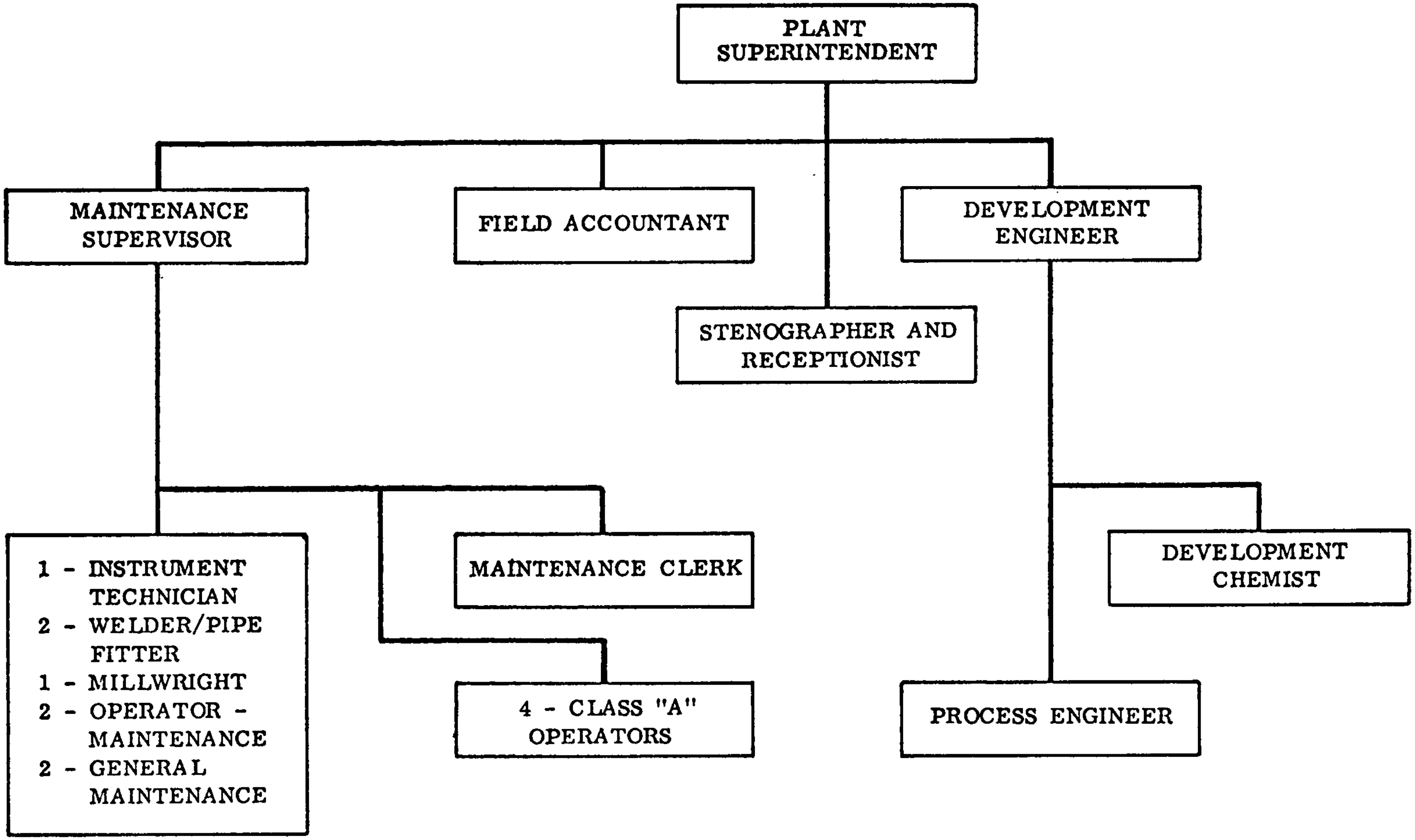


Figure 8.1. Organizational Chart of the Freeport Test Facility, 8th Annual Report

In FY-1963, there were 12 other major modifications to the total instrumentation system of the Freeport Plant, which, as a whole, greatly improved the collection of data and improved the overall operation.

The success of automatic brine level controllers and the other instrumentation modifications permitted a personnel readjustment in FY-1964. The number of plant operators was reduced from 2 men per shift to one man per shift, thereby permitting these positions to be used for other skills and to effect an overall reduction of one position on the plant crew.

The Third Annual Report (FY-1964) states on page 30, paragraph 3-48:

> "The incorporation of automatic level controls into the Plant process equipment has been a significiant improvement to operating efficiency. In addition to facilitating operation, maintenance of the level controls and attendant instrumentation is now confined to routine cleaning, calibrating, and adjusting."

F. NON-PROCESS OPERATIONS

The operations of the Freeport facility involved certain work not directly involved with the LTV process. This effort included:

1. Product Water Treatment
2. Probolog Testing of pipes, heat exchangers and evaporators
3. Materials Testing

1. Product Water Treatment

The Product Water Treatment Program was directed toward the stabilization of the Demonstration Plant product to prevent corrosion in the Freeport public water distribution system, and to respond to consumer complaints of "red water".

The analysis of the product water, four-test runs, prior to treatment is given in Table 8.2.

Table 8.2

Composition of Product Water, Runs 1 through 4

TEST RUN:	1		2		3	4
WATER FOR:	FRPT	DOW	FRPT	DOW	FRPT	DOW
PROPERTY						
Temperature, ^{o}F	107	98	118	104	120	80
Hydrogen Ion Concentration, pH	6.8	7.5	6.3	7.4	6.3	6.6
Total Dissolved Solids, TDS (ppm)	150	20	265	20	140	17
Total Hardness (ppm)	20	3	45	3	25	3
Calcium (ppm)	4	1	9	1	5	1
Magnesium (ppm)	16	2	36	2	20	2
Total Alkalinity as $CaCO_3$ (ppm)	5	2	4	2	3	3
Dissolved Oxygen (ppm)	0.90	nil	0.60	0.05	0.50	0.11
Langlier Saturation Index	-3.37	-3.74	-3.54	-3.78	-3.78	-4.57
Ryznar Stability Index	13.54	15.68	13.38	14.94	13.86	15.74
Classification by Indices	V.C.	V.C.	V.C.	V.C.	V.C.	V.C.

NOTES: V.C. — Very corrosive
Reference — Fourth Annual Report, Table 5-16 and Fifth Annual Report, Table 6-20

Samples of representative materials usually found in water distribution systems were exposed to the above water samples. For certain tests the water was untreated and for other tests the water received the specified treatment. In each case, the extent of corrosion was determined and recorded in mils per year.

The following tables (8.3, 8.4, 8.5, and 8.6), reproduced from the Fourth and Fifth Annual Reports (in which they are Tables 5-17, 5-18, 6-21, and 6-22 respectively), give the test results of exposure of representative distribution system materials to the Freeport product water.

The Product Water Treatment Program described above may be found in the Fourth and Fifth Annual Reports. This work continued successfully, and in the Seventh Annual Report, pages 68-70 and 113-144, the work during Development Runs 14 and 15 was reviewed.

The test during Development Run Number 14 (pages 68-70) included probes located at three points shown in Figure 8.2 (Seventh Annual Report, Figure 2-8) and the point of injection of sodium silicate in the Freeport product header for the first 46 days of the test and at pump 43 (from Effect I) for the next 60 days of this test. The test objectives were to determine:

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

The results of this test are given in Table 8.7 (Seventh Annual Report, Table 2-23).

It is to be noted that the product water immediately after the split-tank is more corrosive than before the tank or before delivery to the Freeport distribution system.

During Development Run Number 15 (page 113-114) the test set up for the first 46 days was repeated with the following changes:

"1. Reading frequency was decreased from three to two times per week.

Table 8.3

Treated Product Water Properties for Test Runs Number 1 and 2

TEST LOOP	UNITS	A		B		C		D	
TEST RUN NO.		1	2	1	2	1	2	1	2
TREATMENT		Na_2CO_3-CaC12	Untreated	Untreated	Untreated	Polyphosphate	Sodium Silicate	$CaCO_3$ Slurry	$CaCO_3$ Slurry
Water		Freeport	Freeport	Dow	Dow	Dow	Dow	Dow	Dow
Chemical addition	ppm	14 as(Ca CO_3)	None	None	None	6as PO_4	2 as(SiO_2)	10 as ($CaCO_3$)	10 as ($CaCO_3$)
Temperature	°F	107	118	98	104	98	104	98	104
Hydrogen Ion Concentration	pH	9. 1	6. 3	7. 5	7. 4	8. 3	10. 1	7. 9	9. 4
Total Dissolved solids	ppm	165	265	20	20	40	60	30	30
Total Hardness	ppm as $CaCO_3$	35	45	3	3	3	3	13	13
Calcium		19	9	1	1	1	1	11	11
Magnesium		16	36	2	2	2	2	2	2
Total Alkalinity	ppm as $CaCO_3$	20	4	2	2	2	34	12	14
Dissolved Oxygen	ppm	0. 90	0. 60	nil	0. 05	nil	0. 05	nil	0. 10
Ryznar Stability Index		8. 84	13. 38	15. 68	14. 94	14. 20	9. 82	10. 78	9. 16
Langelier Saturation Index		+0. 13	-3. 54	-3. 74	-3. 77	-2. 45	+0. 14	-1. 44	+0. 12

Table 8.4

Summary of Test Runs Number 1 and 2

TEST LOOP	A		B		C		D	
TEST RUN NO.	1	2	1	2	1	2	1	2
TREATMENT	Na_2CO_3-$CaCl_2$	None	None	None	Polyphosphate	Sodium Silicate	$CaCO_3$ Slurry	$CaCO_3$ Slurry
Water	Freeport	Freeport	Dow	Dow	Dow	Dow	Dow	Dow
Carbon Steel	8.6	13.8	5.5	5.5	4.3	1.0	6.9	7.4
Galvanized Carbon Steel	13.6	10.3	4.0	3.9	11.6	0.2	3.8	4.5
Cast Iron	8.3	14.7	7.4	7.0	6.0	0.6	9.0	9.1
Copper	0.5	1.3	0.4	0.4	0.3	0.2	0.2	0.2
PVC	None	None	None	None	None	None	None	None
Asbestos-Portland Cement	28.4	36.1	40.3	53.0	66.3	27.8	58.9	44.3
Ni-resist	4.1	7.6	2.2	2.0	2.1	0.7	2.9	2.5
Steam Bronze	0.2	0.9	0.4	None	0.3	0.3	0.4	0.3
Silicon Bronze	0.2	0.9	0.2	None	0.1	None	0.1	0.1
Corrosometer	7.1	10.4	4.2	5.0	3.1	0.3	6.5	5.6

All corrosion rates are expressed in mils per year.

Table 8.5

Treated Product Water, Test Runs Number 3 and 4

TEST LOOP	A		B		C		D	
TEST RUN NO.	3	4	3	4	3	4	3	4
TREATMENT	Oyster Shell	Oyster Shell	none	none	Sodium Silicate	Caustic Soda	Poly-phos'te	Sodium Silicate
WATER	Fr'port	Dow	Fr'port	Dow	Fr'port	Dow	Fr'port	Dow
CHEMICAL ADDITION (PPM)	10 as $CaCO_3$	11 as $CaCO_3$	none	none	20 as SiO_2	pH 8.5 to 9.0	6 as PO_4	6 as SiO_2
Temperature ($^{\circ}$F)	120	80	120	80	120	80	120	80
Hydrogen Ion Concentration (pH)	9.2	9.5	6.3	6.6	10.1	8.9	7.8	9.1
Total Dissolved Solids TDS (PPM)	150	28	140	17	200	20	160	23
Total Hardness as $CaCO_3$ (PPM)	35	14	25	3	25	3	25	13
Calcium (PPM)	15	12	5	1	5	1	5	11
Magnesium (PPM)	20	2	20	2	20	2	20	2
Total Alkalinity as $CaCO_3$ (PPM)	13	14	3	3	63	6	10	9
Dissolved Oxygen (PPM)	1.50	0.60	0.50	0.110	0.50	0.110	0.50	0.110
Langelier Saturation Index	+0.06	-0.64	-3.78	-4.57	+1.20	-2.03	-1.89	-1.68
Ryznar Stability	9.08	10.78	13.86	15.74	7.70	12.96	11.58	12.46

Table 8.6

Summary of Test Runs Number 3 and 4

TEST LOOP	A		B		C		D	
TEST RUN NO.	3	4	3	4	3	4	3	4
TREATMENT	Oyster Shell	Oyster Shell	none	none	Sodium Silicate	Caustic Soda	Poly-phos'te	Sodium Silicate
WATER	Fr'port	Dow	Fr'port	Dow	Fr'port	Dow	Fr'port	Dow
COUPON MATERIAL								
Carbon Steel	1.20	17.2	10.8	15.2	0.9	12.6	12.6	6.8
Galvanized Carbon Steel	7.9	1.2	16.6	11.2	none	*	2.8	4.5
Cast Iron	13.5	27.2	17.5	12.9	1.5	31.6	14.4	9.9
Copper	none	0.7	0.7	0.5	0.1	0.3	0.4	1.0
Polyvinylchloride	none	none	none	none	none	none	none	none
Asbestos-Portland Cement	29.6	18.9	39.0	25.7	18.0	12.1	22.0	13.4
Ni-resist	22 3	6.6	4.6	7.8	0.3	8.8	3.5	5.2
Steam Bronze	0.5	0.6	0.2	0.7	0.2	0.2	0.3	0.6
Silicon Bronze	0.5	0.3	none	0.4	none	0.2	none	0.2
Corrosometer	20.5	11.0	9.3	7.3	0.1	7.1	8.8	negligible

* damaged in cleaning

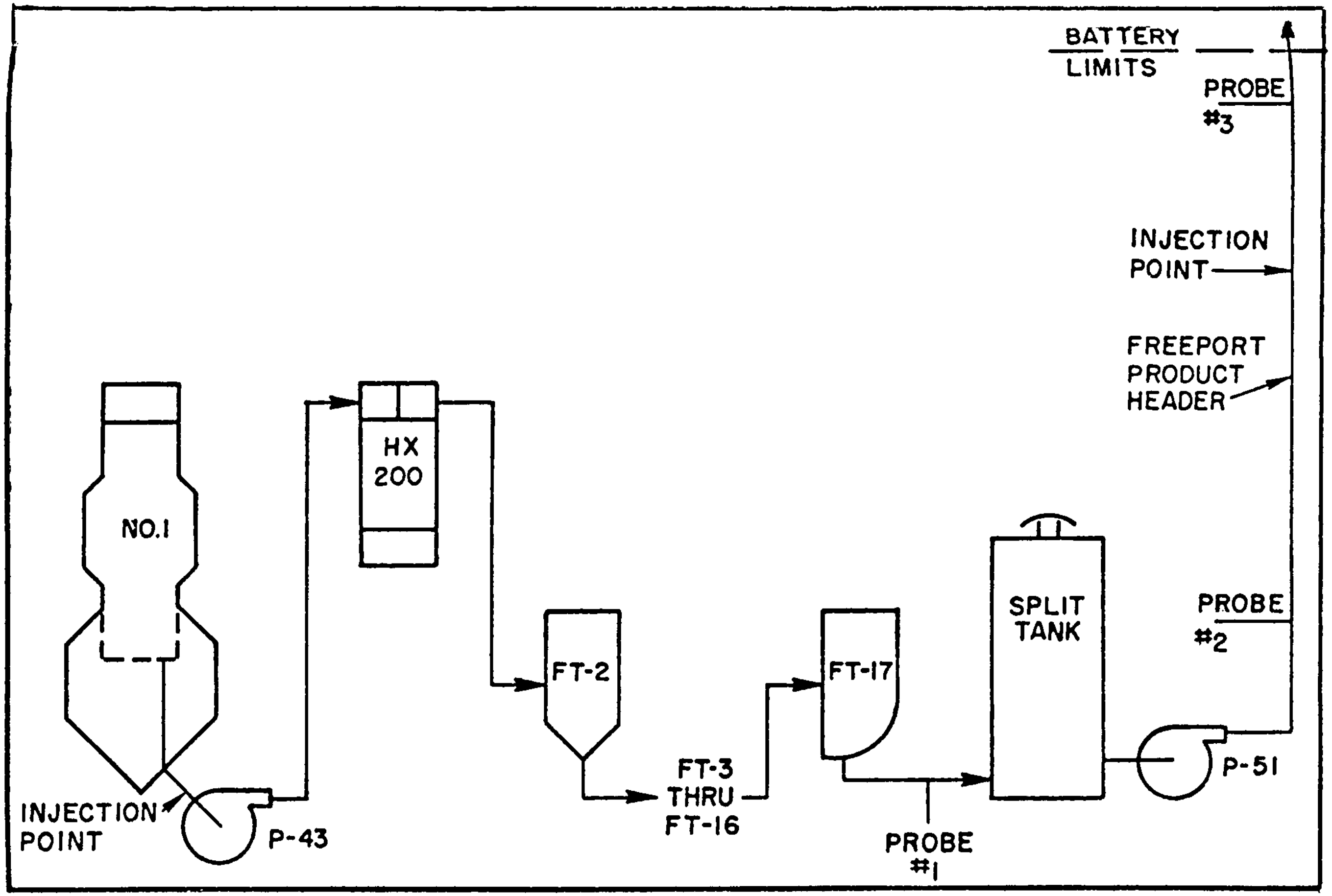


Figure 8.2. Process Sketch Showing Location of Corrosion Probes and Sodium Silicate Injection Points

Table 8.7

Indicated Corrosion Rate in Condensate Circuit — Development Run Number 14

Corrosion Probe Location	Δ Probe Dial Reading	Δ Time (Days)	Calculate Corrosion Rate (MPY)
A. <u>Injection Point —</u> Freeport Product Water Header (Days 0 to 46):			
Inlet to Split Tank	96.0	27	13.0
Outlet to Split Tank	82.7	21	14.4
Freeport Product Line	85.5	26	12.0
B. <u>Injection Point —</u> P-43 Suction (Days 46 to 106):			
Inlet to Split Tank	96.5	46	7.7
Outlet to Split Tank	65.0	21	11.3
Freeport Product Line	42.0	25	6.1

Corrosion Rate, Mils per Year (MPY) =

$$\frac{\Delta \text{ Dial Reading}}{\Delta \text{ Time (Days)}} \times 0.365 \times 10$$

"2. N-Silicate was injected into the Freeport product line only, and just when product delivery was being made to the customer."

The results were given in Table 2-38, Seventh Annual Report, reproduced here:

Probe Location	Time (days)	Corrosion Rate (MPY)
Inlet to split tank	30	5.63 + 0.04
Outlet of split tank	30	8.52 + 0.04
Freeport product line	30	5.40 + 0.04

The results were:

- The corrosion rates were significantly lower than in the previous identical test.
- The corrosion rates were relatively uniform at all locations.
- The corrosion rate at Probe 2 (after split-tank) was slightly higher than at the other two locations.

2. Materials Test Center

The determination of materials of construction for use in a desalting plant was an objective in the operation of the Freeport Plant. The construction of a Materials Test Center was completed at the Freeport Test Facility on April 1, 1969. This facility included a seawater treatment plant, and by August 1969, aluminum and stainless steel test loops, as well as a 3000 GPD test plant to demonstrate the applicability of stainless steel to desalting. The operations of the Test Center include joint industry-OSW projects on a joint funding and benefit sharing basis. Among the initial industrial groups were the Copper Development Association, the Aluminum Association, and the Amercian and Iron and Steel Institute.

Materials contemplated for use in a desalting plant are candidates for testing in this facility. An example of these materials are those contained in Table 8.8, Copper and Copper Alloys under Test (from 1969-1970 Saline Water Conversion Report, page 537, Table 158), which is reproduced here.

Table 8.8

Copper and Copper Alloys under Test

Copper or copper alloy number	Previous trade name	Type of manufacture
122	DHP Copper	Seamless.
142	Arsenical Copper	Seamless.
194	Copper-Iron Alloy.	Seam-welded (SW).
443	Arsenical Admiralty.	Seamless.
608	Aluminum Bronze	Seamless.
613	Aluminum Bronze	Seamless.
687	Arsenical Aluminum Brass.	Seamless.
687	Arsenical Aluminum Brass.	Seamless butt welded (BW).
706	90:10 Copper-Nickel.	Seamless.
706	90:10 Copper-Nickel.	Seam-welded (SW).
706	90:10 Copper-Nickel.	Seamless butt welded (BW).
715	70:30 Copper-Nickel.	Seamless.
715	70:30 Copper-Nickel.	Seam-welded (SW).
715	70:30 Copper-Nickel.	Seamless butt welded (BW).
716	Copper-Nickel + Iron.	Seamless.
Modified 715	70:30 Copper-Nickel + Iron and Manganese.	Seamless.
720	60:40 Copper-Nickel.	Seamless.

The results of material test programs at Freeport are reported for inclusion in the Materials Evaluation Program and stored at the Materials Information Center, Oak Ridge, Tennessee.

3. Probolog Testing

In the realm of non-destructive testing of materials, the Probolog type instrument was used at Freeport to detect variations in the tubes of the evaporators and heat exchangers. The variations could be an actual hole, pits, or general loss or gain of materials.

The Probolog instrument is one that uses electromagnetic energy to induce eddy currents into the test sample for the purpose of detecting discontinuities and/or metallurgical variations. The report "Continuation of Materials Evaluation Program Contract No. 14-01-0001-1657" on page A-39 gives a discussion of Eddy Current Testing. This discussion is principally pertaining to the Probolog instrument and its operation.

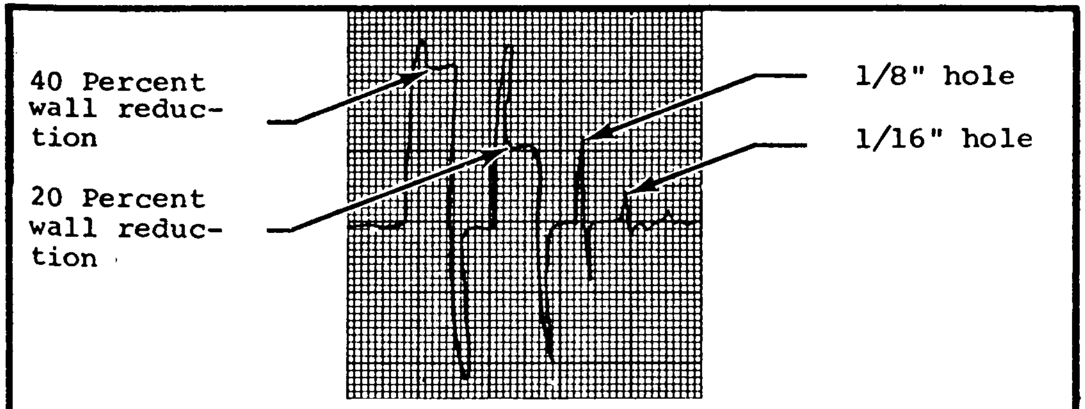
The Probolog instrument is essentially a high precision impedance bridge in which two of the coils are encased in a probe. When the instrument is operated, this probe is drawing through the tube under test at a fixed rate. The tube rate and the recorder speed are synchronized so that a cause for an instrumental response, a defect in the tube, can be exactly located.

Prior to using the Probolog, a reference test must be made, or otherwise be available, on a typical sample of tubing to be tested. The sample would include a section that is undamaged and sections on which various known types of damage are located. A test run is made on this sample to relate the instrument's response to the known damage.

In actual operation, the probe of the instrument drawn through the test tubes and a recording is made of the responses received. By comparing the signals for any defects located with the sample test run, an experienced operator can determine the type of defect and its relative seriousness with respect to further operation of the subject tube.

In a plant such as the Freeport Demonstration Plant where the number of tubes may easily exceed 8000, it is impractical to conduct a Probolog test of each individual tube. By means of statistical testing, it is possible to show that the actual testing of approximately 40 tubes of the bundle in an evaporator or heat exchanger will yield test information pertinent to the entire bundle. The information thus obtained will have a 90% confidence factor.

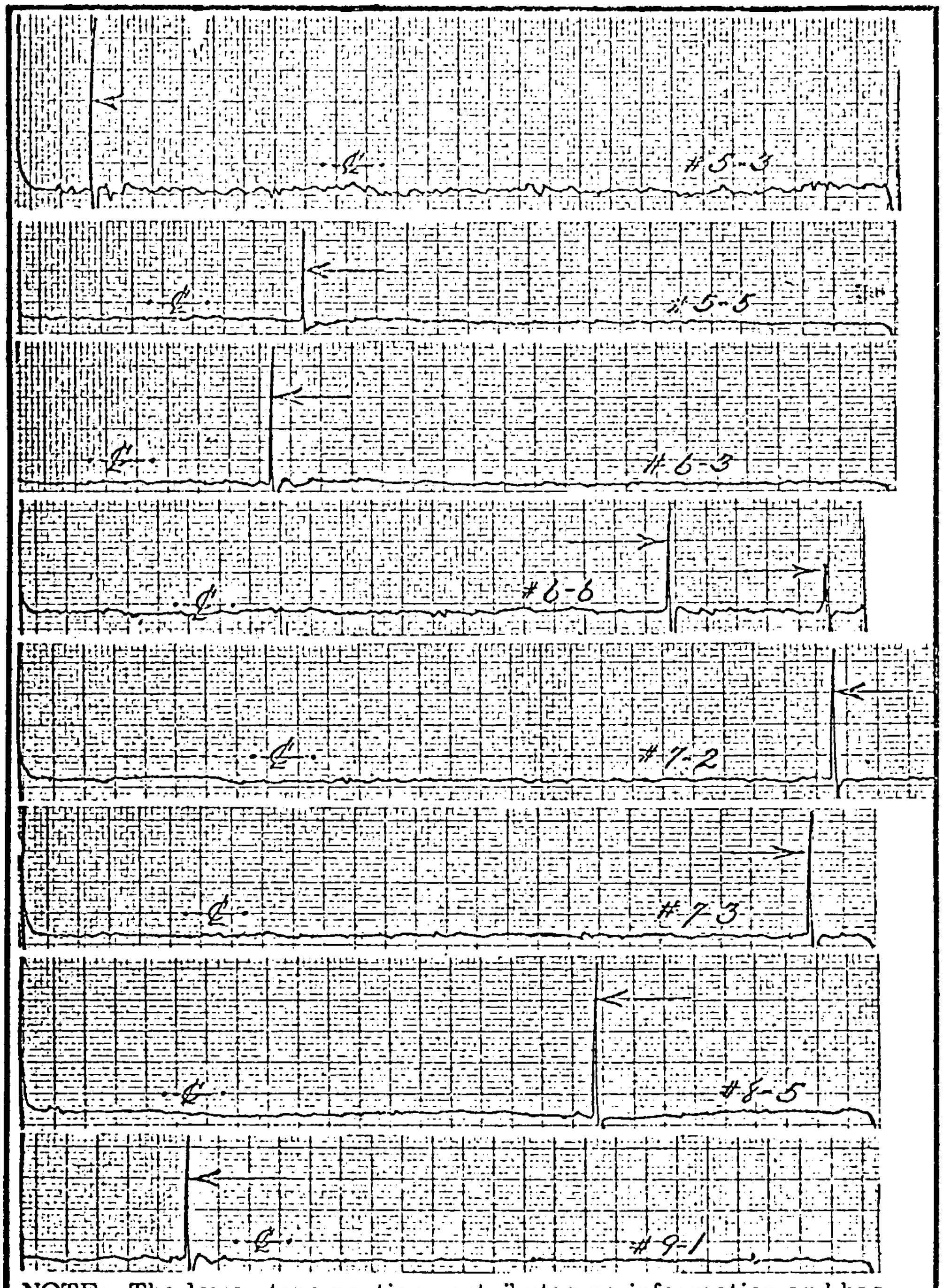
Examples of a Tubing Standard and Probolog recordings of tubes tested are reproduced from the Sixth and Seventh Annual Reports as Figures 8.3 and 8.4.



INSTRUMENT SETTINGS:

Phase 8.5
Frequency 3
Sensitivity 17
Thinning 3
Probe 32 SF

Figure 8.3. Tubing Standard Recording



NOTE: The lower tape portion contributes no information and has been removed.

Figure 8.4. Probolog Recordings of Tubes with Indications of Major Defects

G. WASTE DISPOSAL

The wastes from the Freeport Demonstration Plant were: the brine blowdown from the last effect (XII or XVII, as the case may be); liquids (water) normally associated with a metallurgical type laboratory; and, in the later years principally, the product water that was not delivered to the City of Freeport, to the Dow Chemical Company, or used within the plant.

Qualitywise, these latter two waste liquids were no worse than the seawater feed to the plant, and often better. Therefore, these wastes can be, and were, discharged into an adjacent flume without any serious perturbation to the water in the flume.

Inasmuch as the concentration factor (CF) of the seawater in the flume was, generally, significantly less than that of normal seawater, any recirculation of "waste water" of a better quality would have only further reduce the intake concentration factor. For a demonstration plant, a low CF of the intake water makes an operational comparison of this plant with others more difficult. For this reason, careful consideration of the point of discharge of such wastes is merited.

The discharge of the brine from the last effect presented a greater operational problem and one that could impact an operational as well as a developmental plant.

The initial design of the Freeport Plant had the seawater intake and brine discharge in the same flume. Condenser 312 and the barometric condenser both used raw seawater as the cooling fluid. The warm seawater from these condensers was combined with the brine blowdown from the last effect (XII), which was warmer than flumewater, for discharge into flume adjacent to the plant intake. Thus, the situation that created two operational problems was established.

The possibility that recirculation of the brine discharge was occurring came to light as a result of a sudden and unexplained improvement in the deaerator performance. The investigation showed conclusively that recirculation of the brine was taking place in the flume, which resulted in a localized increase in the concentration factor about the intake and discharge pipes. An attempt to correct this situation was made in May 1962 by installing a barrier between the intake and outlet lines to improve the performance of the intake pumps, and by reworking the valves in the hot well and intake system to reduce the leakage (First Annual Report, page II-13).

Recirculation and a resulting high concentration factor continued to be a problem in FY-1963, necessitating the reduction of product water output to prevent the brine CF from the last effect from exceeding the set point that had been established as a production variable (Second Annual Report, page II-21).

The solution to this problem was the discharge of the brine blowdown into a nearby "waste water canal," an action that apparently was accomplished in FY-1963 (Second Annual Report, page II-21).

The intake seawater continued to be a problem, although the concentration factor was not involved. Some raw seawater was diverted from the intake for use as a coolant in Condenser 312 and then returned to the flume via the alternate intake line. This "coolant" could pick up enough heat to raise its temperature as much as 25°F (Table 3, Development Report Number 2, page 12). Since pressure conditions on the steamside of Condenser 312 were adversely affected by warm condenser water, the overall plant ΔT was seriously affected by a very small change in condenser pressure.

The study conducted during Development Run Number 2 on the question of warm condenser water resulted in the careful closing of the pertinent valves in the intake system whenever an operational change of intake pumps was made. This action reduced the intake of warm seawater that was discharged from Condenser 312. The ultimate solution would be to discharge this condenser water into another canal or otherwise physically separate the intake-discharge locations.

Chapter 9

EVALUATION OF OVERALL PERFORMANCE

A. FREEPORT DEMONSTRATION PLANT

On February 6, 1964, an operating program for the First Demonstration Plant was instituted to fulfill the objectives of the Resolution by the Senate and House of Representative of the United States of America in Congress Assembled; namely, to operate the plant "for the production, from seawater or brackish water, of water suitable for agriculture, industrial, municipal, and other beneficial consumptive uses." (Joint Resolution, PL 85-883, 85th Congress, September 2, 1958.) This program is better identified as "Development Runs" undertaken to accomplish certain engineering, operational, and economic objectives consistent with the cited Joint Resolution.

There were 16 such Development Runs between February 6, 1964 and May 9, 1969, when the plant was finally shut down. Several of the Development Runs were conducted in parts or subruns with their own definite objectives. The operating data generally were tabulated on "Operator Data Sheets" as one run. The exceptions were Development Runs Number 10 and 10A (November 26, 1965 through March 28, 1966), which were reported on separate data sheets. Development Run 15 was reported on two separate Operator Data Sheets due to the fact that it began in one fiscal year and finished in the next — each portion was reported separately. Finally, the operational data (Operator Data Sheet) for Development Runs Number 11A, 12 and 12A were not available, either as an individual Development Report or as an integral part of the Contractor's 6th Annual Report (R&D Progress Report Number 440) of operations. Therefore, these data are omitted. Information on the operations prior to February 6, 1964, was unavailable.

Those sections and columns of the Operator Data Sheet which were used in the analysis of the operation are as follows:

- Operating time — date and hours
- Production (including condensate) — total pounds of product water
- Raw materials — steam (Dow), total pounds; power, total Kw Hrs.; seawater feed, total pounds.
- Process conditions — concentration factor, in and out; temperature, brine discharge from final effect; temperature, steam to Effect I.
- Calculated quantities and ratios — extraction ratio; H_2O/lb Dow steam feed, with condensate; pounds H_2O per Kw Hr.; Kw Hr per pounds SWF.

Those columns under the general headings "Production, including Condensate" and "Raw Materials" imply a time rate. For example, "Pounds, Total" for Production and Seawater Feed are recorded for the date and operating period which suggests a natural division to obtain pounds per hour. Those columns under the general headings "Process Conditions" and "Calculated Quantities and Ratios" do not imply this time relationship.

The Extraction Ratio (ER) recorded in the Operator Data Sheet was calculated by using the salt concentration factor applicable to the Seawater Feed to Effect I (CF_i) and to the Brine Discharge from Effect XII or XVII (CF_o), as the case may be. The formula for this computation is:

$$ER = 1 - \frac{CF_i}{CF_o}$$

It should be noted that this formula <u>does not</u> include the condensate from the steam to Effect I.

The value of CF_i and CF_o was determined each operating day. The ER thus computed can be multiplied by the Seawater Feed to calculate the amount of water separated from the seawater. The result is, of course, a theoretical value, and there were occasions when the metered production quantity was less than this calculated amount.

The next to the last column of the Operator Data Sheet is entitled "Lbs. H_2O per Kw Hr" and the entered data may be given as ".506" or "506". In the first example, the heading of the column should read "M lbs H_2O per Kw Hr" where M = 1000. The latter example is correct for the column heading as given. The reader, likewise, needs to be alert when concerned with the final column. The denominator "Lbs. SWF" is in reality "M lbs. SWF".

In actual practice the Operator Data Sheet was filled in one line per each day of operation. Therefore, it has been assumed that the recorded data is the average for the operating period of the indicated day.

The data sheet for each Development Run was reduced to average values for each of the columns noted above. When the column implied a rate, such as "Production, Pounds, Total", the sum for the subject column was divided by the "total operation hours" applicable to that column to obtain the "average hourly rate". This procedure, with respect to the colum "Power, Total Kw Hr", yields a number that is in reality the average "kilowatts" of connected operating load. However, the value thus obtained plays the same role when used with the "averages" as does the "Power, Total, Kw Hr" when used with the appropriate column totals. Those column headings, such as "Extraction Ratio" and "Discharge Brine from XVII (or XII)", are divided by the respective number of entries in the subject column to obtain the desired column average. In this manner, the Operator Data Sheets for each Development Run were averaged and reduced to one line per run. The results of this "averaging" of each Development Run may be found in Table 9.1.

Initially, the Freeport Plant was built and operated as a 12-Effect plant. In the spring of 1967, the plant was enlarged by the addition of a 5-Effect Module for a total of 17 Effects. The nominal capacity of the plant was unchanged although certain of the parameters were impacted. A comparison of these parameters was given in the Sixth Annual Report, Freeport Test Facility, Freeport, Texas, Table 2-3, page 21, which is extracted below.

	12-Effect	17-Effect
Nominal Capacity, 10^6 GPD/Net	1.0	1.0
Gain Ratio, Net (lbs. water/lbs. steam)	10.5	13.8
Extraction Ratio (lbs. water/lbs. SWF)	0.71	0.68
Seawater Supply Rate, lbs/hr	490,000	510,000
Power Demand, kilowatts (including intake pumps)	360	450

From the above table, it is possible to compute the following data:

	12-Effect	17-Effect
Nominal Capacity, lbs./hr. (approximately), Net	350,000	350,000
Steam Rate to Effect I, lbs./hr.	33,000	25,300
Kw Hr/M lbs. SWF	0.735	0.883
Lbs. Water Extracted/Kw Hr.	0.972	0.778

Table 9.1

Selected Averaged Data for Development Runs

1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23
Dev. Run. No.	Year	Dates		Period Total Hours	Time		% Operating Time	SWF LB/HR. (x 1000)	Product Water LB/HR (x 1000)	Steam (Dow) LB/HR	Electricity Kwhr	Concentration Factor		Steam to Effect 1 °F	Brine Out °F	Plant ΔT °F	E.R. $\frac{LB_{H_2O}}{LB_{SWF}}$	Rates P.R. $\frac{LB_{H_2O}}{LB_{Steam}}$	$\frac{\text{MLB } H_2O}{\text{Kwhr}}$	$\frac{\text{Kwhr}}{\text{MLB SWF}}$	E.R. Cols. 20x21	E.R.[2] $\frac{\text{Col. 10}}{\text{Col. 9}}$
		From	To		Operating Hours.	Outage Hours						CF_I In	CF_o Out									
1	1964	2- 6	3- 6	695.0	695.0	0.0	100.0	484.0	343.5	30,900	358	0.991	2.957	240.1	120.5	119.6	0.665	11.08	0.958	--	--	0.710
2	1964	3-14	4-13	717.0	707.4	9.6	98.7	518.0	367.0	31,700	364	0.872	2.943	244.0	120.3	123.7	0.704	11.57	0.999	0.775	0.774	0.708
3	1964	4-23	6-10	1165.0	1151.5	13.5	98.8	471.5	363.0	31,100	386	0.820[3]	2.980[3]	244.6	116.7	127.9	0.705	11.67	0.939	0.819	0.769	0.770
4	1964	6-21	7- 6	359.5	358.5	1.0	99.7	447.0	311.0	27,000	409	0.920	2.560	228.0	117.0	111.0	0.641	11.72	0.815	0.855	0.697	0.696
5	1964	7-13	9-24	1736.0	1111.8	624.2	64.1	468.0	349.0	30,500	406	1.020	3.265	246.0	120.5	125.5	0.678	11.71	0.878	0.948	0.832	0.746
6	1964	10- 1	11- 2	792.0	745.5	46.5	94.2	522.0	403.0	38,200	432	0.913	3.250	270.2	142.1	128.1	0.718	10.59	0.958	0.818	0.784	0.772
7	1964	12- 1	3- 3	2232.0	1765.0	467.0	79.1	482.0	366.5	35,000	402	0.940	3.028	281.7	121.6	160.1	0.688	11.08	0.908	0.860	0.761	0.760
8	1965	3-31	5-24	1282.0	1280.5	1.5	99.9	478.0	347.0	34,000	341	0.830	2.580	273.0	121.0	152.0	0.679	10.44	0.827	0.878	0.726	0.726
9	1965	6-21	11- 1	3178.0	2930.5	247.5	92.2	487.0	365.0	34,500	399	0.950	2.880	272.0	121.0	151.0	0.668	10.62	0.885	0.864	0.765	0.749
10	1965	11-26	2- 7	1746.3	1699.5	46.8	97.3	467.0	341.0	32,800	401	0.950	2.730	291.0	121.0	170.0	0.648	10.41	0.835	0.898	0.750	0.730
10A	1966	2-26	3-28	718.8	664.8	54.0	92.5	416.0	288.0	29,100	461	0.880	2.350	266.0	119.0	147.0	0.621	10.65	0.604	1.116	0.674	0.692
11	1966	5-16	5-30	329.5	321.5	8.0	97.6	462.0	338.0	32,550	437	0.940	2.670	292.0	126.0	166.0	0.650	10.29	0.727	0.995	0.723	0.732
12[1]																						
13	1967	7-10	8- 4	696.0	555.5	140.5	79.8	480.0	278.0	24,700	525	0.952	2.760	269.0	115.0	154.0	0.655	11.85	0.529	1.092	0.578	0.579
14	1967	9-29	1-12	2516.0	2311.8	204.2	91.9	466.0	332.0	29,700	515	0.913	2.775	289.5	93.4	196.1	0.667	12.69	0.644	1.112	0.716	0.712
15	1968	4-14	7- 5	1898.8	1588.8	310.0	83.7	467.5	344.5	24,650	515	0.856	2.565	276.5	101.0	175.5	0.661	13.62	0.666	1.095	0.729	0.737
15	1968	7- 6	9-20	1848.0	1539.4	308.6	83.3	471.5	334.0	23,550	521	0.938	2.535	279.0	109.0	170.0	0.637	14.20	0.630	1.110	0.699	0.708
16	1968	12-18	5- 9	3288.0	3091.1	196.9	94.0	474.5	368.5	25,100	548	0.745	2.351	281.5	86.9	194.6	0.683	13.65	0.673	1.152	0.775	0.777

(1) Data for Run Number 12 was not available.
(2) This ratio is obtained by dividing Product Water (column 10) by SWF (column 9) - includes condensate.

(3) Using tabulated data, $ER = 1 - \frac{CF_I}{CF_o} = 0.725$

The gross gain ratio is obtained by adding "one" to the net gain ratio; the gross gain ratio for the 12-effect plant is 11.5. The weight of one gallon of water at 68°F is 8.31 pounds. The terms "gain ratio", "economy ratio" and "performance ratio" are synonymous. It should be noted that the product of the last two parameters, (Kw Hr/M lbs. SWF) x (Lbs. Water Extracted/Kw Hr), equates to the extraction ratio.

The operation of the Freeport Plant is very logically divided into two parts: i.e., operation as a 12-effect plant, and operation as a 17-effect plant. In the former configuration there were 12 Development Runs, and in the latter, four Development Runs. In this discussion, Run 15 is divided and presented in the bar graphs as "two" runs only because the run was started in one fiscal year and finished in the next. The report of the run is thus divided (about equally) between two annual reports.

The following discussion will be divided in accordance with the 12- and 17-effect sequence of operation.

In Figure 9.1 through 9.5, various data from Table 9.1 have been plotted versus Development Runs to provide a quick appraisal of the plant operations. The 12-effect average rates for seawater feed (475,200 lbs/hr), steam to Effect I (32,280 lbs/hr), and production (348,500 lbs/hr) may be seen in Figure 9.1. The respective design values are given above. As expected, these rates generally show identical variations; e.g., increased production corresponds to increases in steam and SWF.

The overall average production rate of 343,700 pounds per hour, which includes the steam condensate, approimated the design parameter of 350,000 pound per hour. This rate was achieved with the SWF and steam rates below their respective design values.

As experience was acquired in the operation of the Freeport Plant, the plant temperatures were changed to determine their impact upon the overall plant characteristics. Beginning with Development Run Number 6 (see Figure 9.2), the temperature of the steam to Effect I was increased from an average of 240.5°F (Runs 1 through 5) to an average of 278°F (Runs 6 through 11), while holding the temperature of the brine from Effect XII essentially steady. Effectively, the plant temperature difference was increased from an average of 121.5°F (Runs 1 through 6) to 153.5°F (Runs 7 through 11) — the temperature of steam in, minus the temperature of brine out.

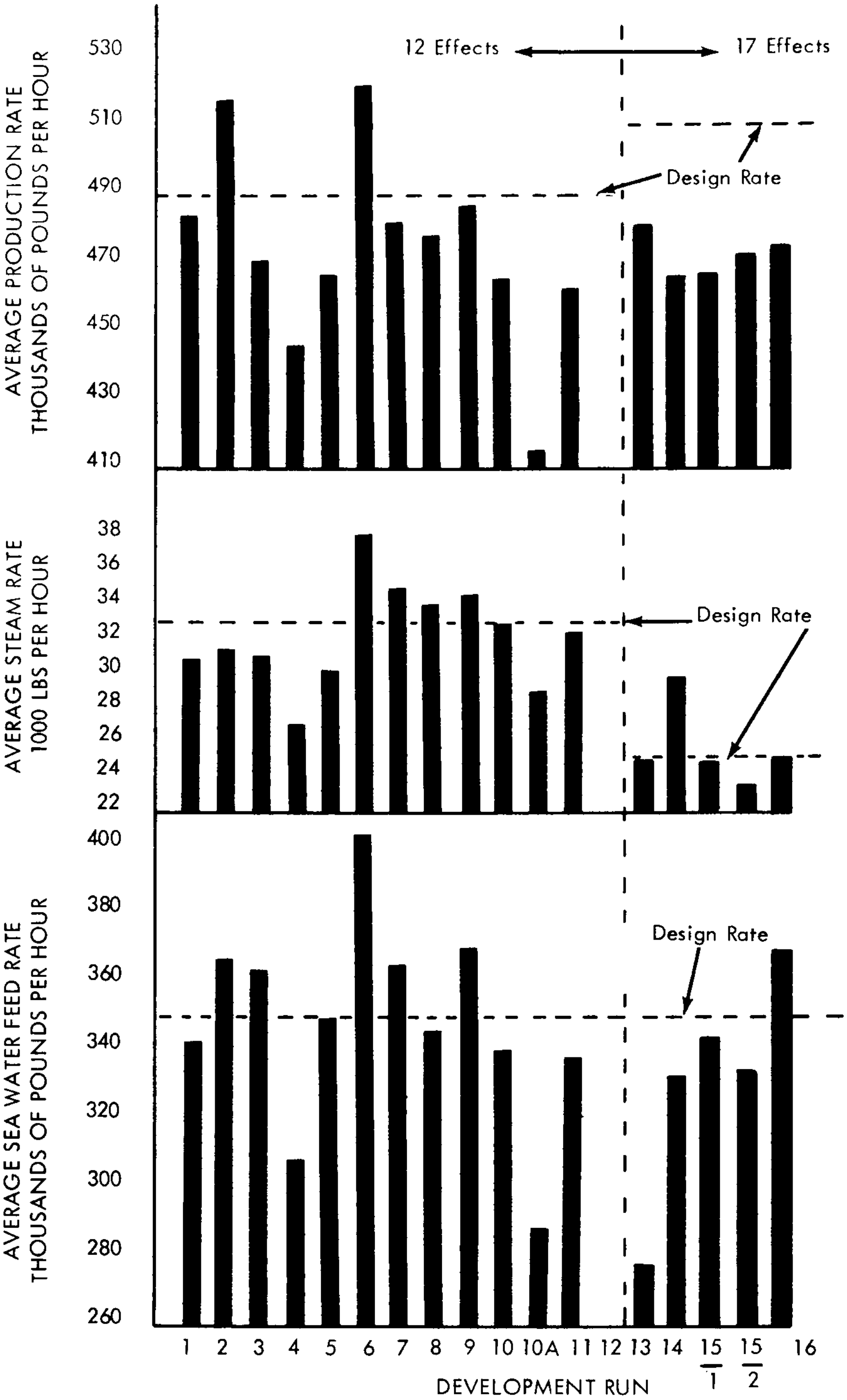


Figure 9.1. Production Parameters

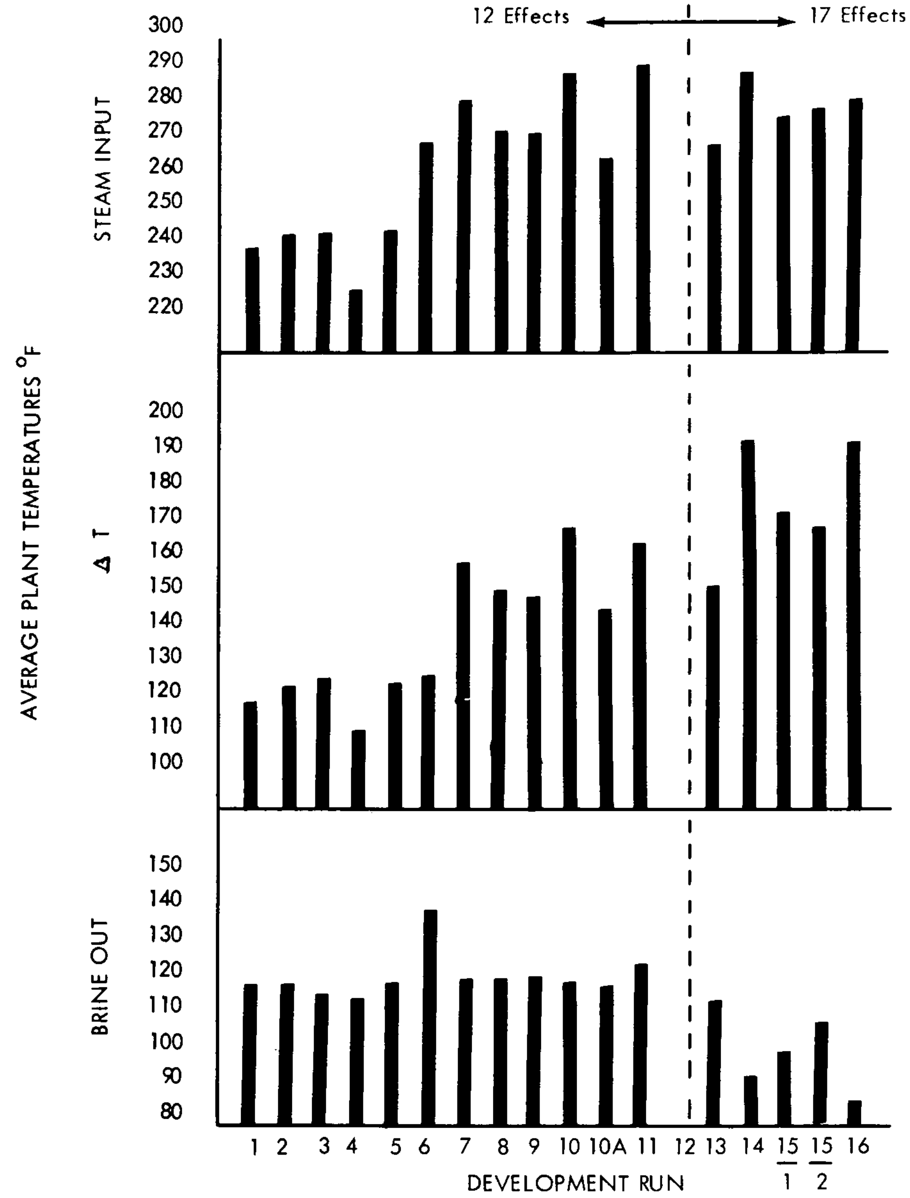


Figure 9.2. Plant Temperatures

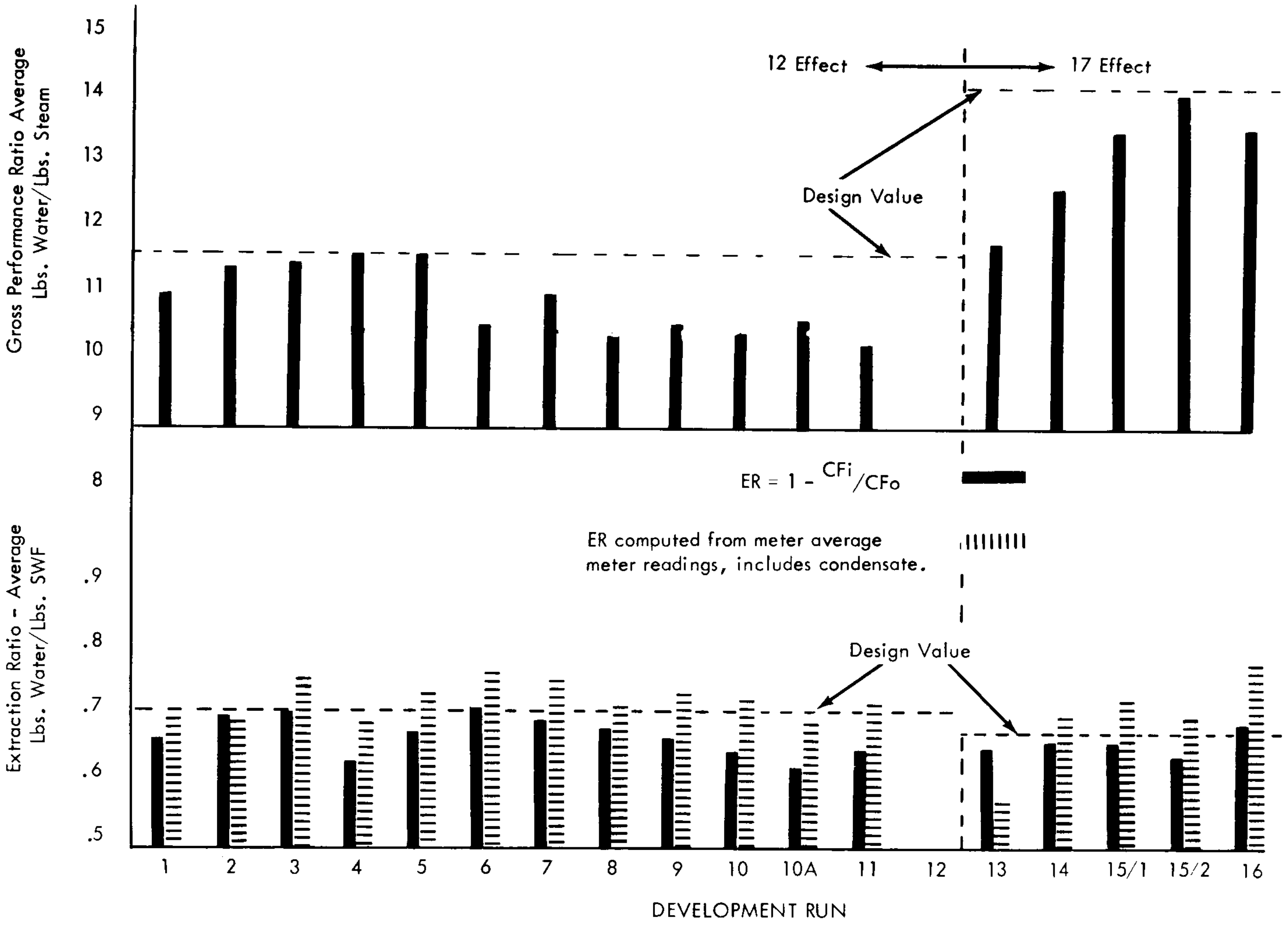


Figure 9.3. Plant Ratios

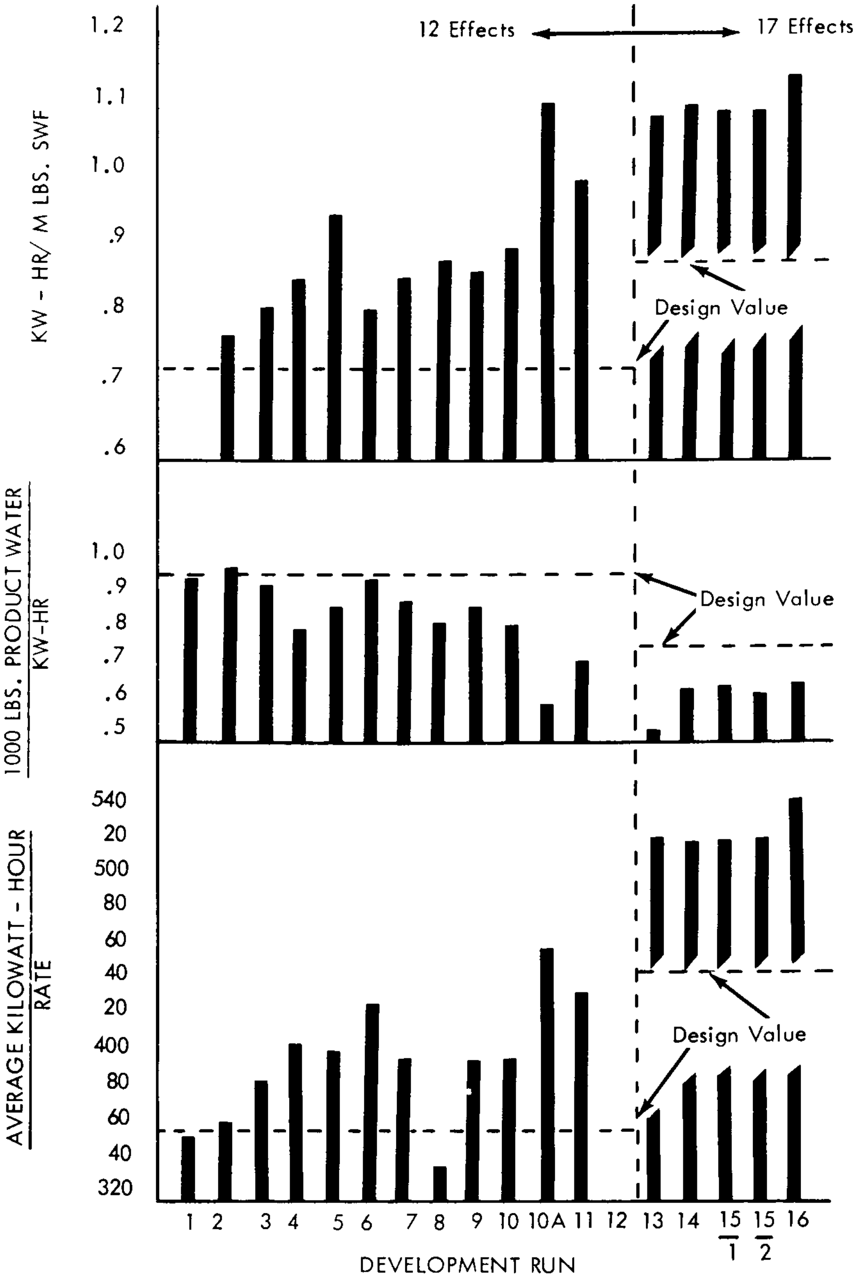


Figure 9.4. Electrical Consumption

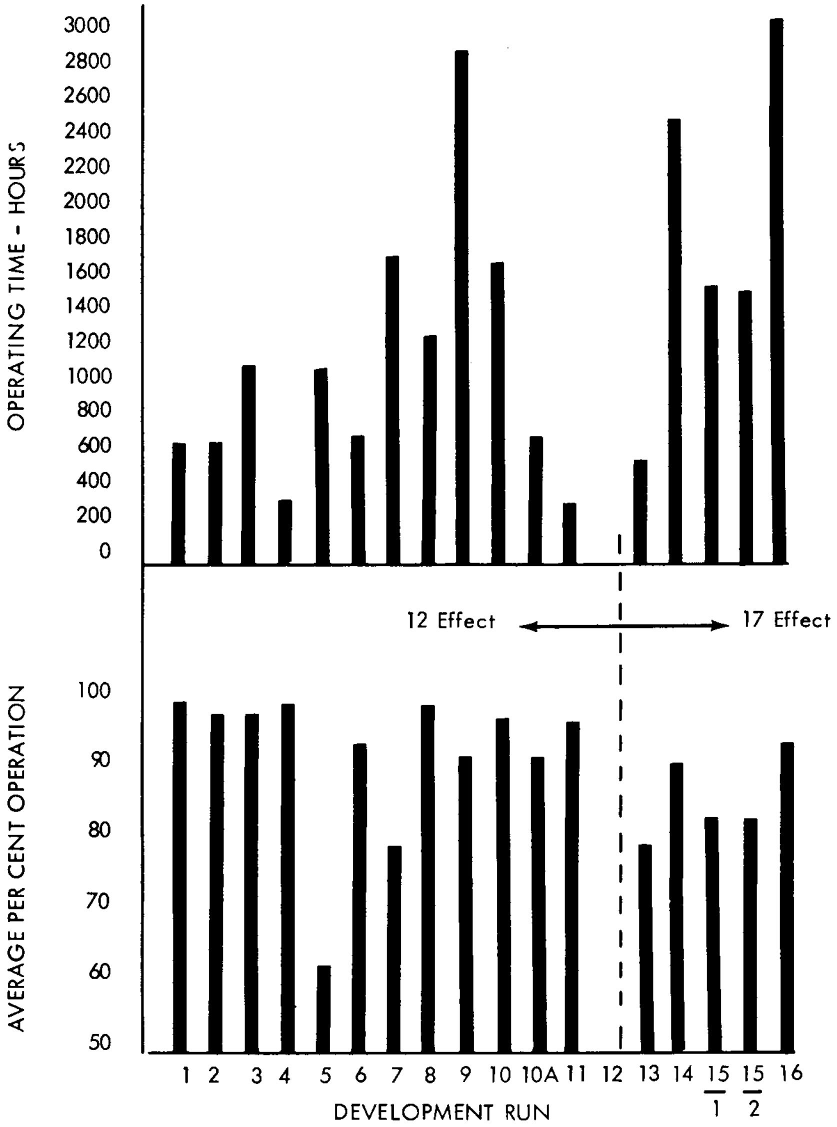


Figure 9.5. Operating Time

In general, the gross performance ratio (Figure 9.3) was reasonably steady for the operation as a 12-effect plant, the average ratio being 10.97 as compared to the design point of 11.5, or 4.7% below the design point. Development Runs 2, 3, 4 and 5 achieved design performance for this parameter.

The extraction ratio exhibited the steadiest operating value throughout the Development Runs. The data on this parameter are graphed first as the average value obtained by computation using the relative concentration factors, and second, as representing the gross ratio (pounds of product water including condensate divided by pounds of SWF) as reported from (presumably) metered readings. The reported computed value of this ratio, with one exception, did not reach the design value of 0.71. Run Number 6, in which the ratio was 0.720, is the exception. Runs Number 2 and 3 were very close, with reported values of 0.704 and 0.705 respectively. The latter value may be a misprint since the stated values of CF_i and CF_o (Run 3), when averaged, will compute to 0.725.

The graphs of Figure 9.4 depict the variation of Kw Hrs per M pounds SWF; pound product water (gross) per Kw Hrs; and the averaged Kw Hrs rate plotted for each Development Run. The 12-effect plant anticipated a connected load of 360 kilowatts. The actual connected load for the 12-effect plant was approximately 400 kilowatts, or 11% over the design point. This deviation from the design values shows up directly in the Kw Hrs per M pounds SWF ratio and inversely in the pounds product water per Kw Hrs ratio.

A graph of the actual periods of operation and the average percent operation are given in Figure 9.5. The graph of actual operating time is included as a matter of information. The duration of each run was essentially independent of the other runs and was governed principally by the objectives of the run. The average percent operation — the ratio of the actual production hours to the total hours for the operating period — gives an indication of the relative amount of unscheduled downtime in each operating period.

The overall operating results of the 12-effect plant were very gratifying. The averaged gross production rate (348,500 lbs/hr) was nearly equal to the design parameter (350,000 lbs/hr) while the corresponding seawater feed (SWF) and steam to Effect I were below their respective design values. This picture is seen upon examination of Figure 9.1. As a result of operating experience and a better capability to control the scale problem, the steam temperature to Effect I was increased to approximately 278°F from the original operating value of 240°F. This temperature change tended to increase the production of potable water. Concurrently, the performance ratio and extraction ratio were close to the predicted design values and indicated a capability of achieving these ratios, were the plant to be seriously operated as a production plant for a reasonable period of time.

It was noted during the preparation of Figure 9.1 that Development Runs 2, 4, 6 and 10A seemed to depart from the pattern of the bar graphs. The following paragraphs will address these Development Runs.

Development Runs 2 and 4 are noted principally due to the fact that they represent the first real "high" and "low" in the data reported. The overall pattern of SWF — steam input — production is followed; both runs evidenced a performance ratio approximating the anticipated design value.

The objectives of Development Run Number 6 included the evaluation of production at the maximum obtainable seawater feed rate. Certain target operating conditions, as given by Table 5-2, page 79, Fourth Annual Report, are cited below and compared with the average values obtained.

	Target	Average
Temperatures:		
Steam to Effect I, °F	260 to 270	270.2
Brine from Effect XII, °F	145 maximum	142.1
Concentration Factors:		
Seawater in, CF_i	1.1 ± 0.1	0.913
Brine Out, CF_o	3.3 ± 0.1	3.250
Active Heat Transfer		
Surface	66,247 ft^2	---

Run Number 6 achieved the above cited objective and operating conditions. The operating period was 33 days, including the day of start-up and a day for shutdown. In this time period there were 4 days in which the operating (producing) time was less than 24 hours. The daily production exceeded the nominal capacity of 1,000,000 GPD on each of the 27 full operating days, reaching a peak of 1,211,000 GPD. The SWF rate likewise was high — averaging 522,000 pounds per hour for this operating period. The plant ΔT of 128°F was not unusual, as the brine outlet temperature was purposely held high, a fact which tended to limit the achievement of other objectives.

The "pumping capability [of the plant] was identified as the limiting factor at a maximum production rate of 1,211,000 GPM metered over a 24-hour period, and higher production rates over shorter periods." (Paragraph 5-22, page 78, Fourth Annual Report, R&D Progress Report Number 171.) The report also notes in this paragraph "...that no mineral scale was deposited, even at the highest temperatures reached [263°F for the brine blowdown from Effect I]."

In general, Run 6 conformed to the overall operational pattern that can be seen in Figures 9.1 through 9.5.

The downward spike of Run 10A in Figure 9.1 merits consideration. A line-by-line perusal of the Operator Data Sheets disclosed a possible discrepancy in the column "Steam (Dow, Total lbs." Sixteen of the 30 lines have entries that were equal to or less than the product of "Steam, Maximum Rate, lbs/hr" and the indicated "Operating Hours"; the remaining 14 entries are greater than the maximum possible total steam. (March 24: 24 operating hours multiplied by 19,800 lbs/hr yields approximately 475,000 lbs. The reported value is 637,000 lbs.) The column "H_2O/lb Dow Steam Feed with Condensate" includes ratios of 3.40 (8.75 operating hours on February 26), 2.74 (8 hours operation on March 9), and 7.14 (24 hours operation on March 24), as well as 32.64 (11 hours operation on March 7) and 19.14 (14 hours operation on March 28, the day of shutdown). Since the steam consumption directly impacts the Economy Ratio (lbs H_2O/lbs steam), it is believed essential to record complete accurate information of the Operation Data Sheets, noting the reason(s) for deviations from normal.

The Fifth Annual Report, Saline Water Conversion Demonstration Plant Number 1, Freeport, Texas, has pertinent comments on the data obtained in Run 10A. The objectives for this run include (page 108): "(3) Evaluate performance of overall process and of Effect 1 with reduced venting rate"; and, "(6) Evaluated the variable rate performance of the overall plant." It was to be expected, and so stated on pages 109 and 135 (Fifth Annual Report) that the production rate for this run would be low. The heat losses during this run were unusually high due to poor plant insulation. In addition, the following comments on operating conditions during the run were extracted from the cited Annual Report (page 135):

> "The ability to transfer heat in Effect 1 was impaired during Development Run No. 10A. This condition was carefully evaluated from data and from visual inspections, all of which were reported in a Letter Report dated May 16, 1966. The conclusions were:

> "(1) that a porous mineral scale, <u>not</u> $CaSO_4$ of any form, was deposited on the inside of the evaporator tubes during the outage between Development Run Nos. 10 and 10A; and
>
> "(2) that a liquid level of heating steam condensate may have existed in the shell of the evaporator tube bundle, because of a restriction in the drain line resulting from an accumulation of magnetic iron oxide flakes."

Thus, it really should not be surprising that the operating data for Run 10A indicated less production of potable water than did the data of other previous runs.

The various data from Table 9.1, applicable to the 17-effect plant, have been plotted versus Development Runs to provide a quick appraisal of plant operations (Figure 9.1 through 9.5). It should be remembered that as a 17-effect plant, there were only four development runs, as compared to 12 runs in the 12-effect configuration. Although the knowledge gained in the preceding operations was beneficial, the results obtained in 4 runs must be viewed with some reservation, if only for an apparent lack of a comparable quantity of data. In fact, the total operating time was 13,432 hours (Runs 1 through 11) and 9,086 hours (Runs 13 through 16), or the 17-effect plant accrued approximately 67.5% of the operating time of the 12-effect plant.

The production parameters given in Figure 9.1 are the seawater feed rate (SWF), the steam input rate to the plant, and the production rate, including condensate. These rates are averaged quantities for each development run. The SWF rate was fairly constant at 471,900 lbs/hr as compared to the design value of 510,000 lbs/hr. The steam rate (25,540 lbs/hr) was slightly greater than the design point. The gross production rate fluctuated in that Run 13 output was unusually low at 278,000 lbs/hr. The averaged production rate of the 5 runs in the 17-effect configuration was 331,400 lbs/hr. However, omitting data from Run 13, the averaged rate was 344,750 lbs/hr, which is more nearly the design value of 350,000 lbs/hr.

The average plant temperatures are shown in Figure 9.2. The high inlet steam temperature to Effect I, used in the latter runs while in the 12-effect configuration, was continued. The temperature of the brine from Effect XII was lowered from the corresponding outlet temperature in the previous plant by approximately 20°F (12-effect plant averaged 122°F and the 17-effect plant 101°F). The net result was the continued large plant ΔT which could be expected to maintain a good production rate. Record ΔT's of 196°F and 195°F were achieved in Runs 14 and 16. The brine temperature out in Run 16 was 87°F and the lowest of any run.

The gross performance ratio and extraction ratio are given in Figure 9.3. The performance ratio shows a steady improvement from 11.85 for Run 13 to a maximum of 14.2 which was recorded for that portion of Run 15 that occurred in FY-1969. The overall average for Runs 13 through 16 inclusive was 13.20, as compared to a design value of 14.8.

The extraction ratio obtained from the formula involving the concentration factors shows a very good degree of consistency. Similarly, the gross ratio, which is based upon metered input SWF and output production, is reasonably steady, if the result of Run 13 is omitted. Only Run 16 had a computed extraction ratio that exceeded the design point.

Figure 9.4 shows the impact of the power demand of the 17-effect configuration. The plant design anticipated an increased load to an estimated 450 kilowatts, whereas the actual demand turned out to be in the order of 525 kilowatts. The result is a rather sharp departure from the design values and previous experiences for the ratios Kw Hr per 1000 pounds SWF and pounds product water per Kw Hr. For the former, the change is directly with the power consumed, and for the latter, the change is inverse.

The operating time per run in the 17-effect configuration, as shown in Figure 9.5, indicates a comparable time period with the 12-effect plant experience for the collection of data. In other words, the duration of each run was sufficient to assure the acquisition of meaningful data. The percentage of operating time per period was approximately 87%, varying from 79.8% in Run 13 to 94.0% for Run 16.

An overview of Figures 9.1 through 9.5, 17-effect operation, quickly points to Run 13 as one that seems to be at variance with the others. The production quantity, the gross extraction ratio, and operating time seem to be low, although the other parameters appear to be comparable. It should be noted that Run 13 was the initial operation of the full 17-effect plant. It could be anticipated — certainly, it would not be unusual — that the initial operation would reflect other than design or near design point performance (see Figures 9.1, 9.3, and 9.5).

The Seventh Annual Report (FY-1968), Freeport Test Facility and Vertical-Tube-Evaporator Test-Bed Plant, Freeport Texas, states the objectives of Run 13 (page 8), which included:

> "A. To obtain heat and material balance data to establish the overall process performance of 17-Effects with allflash product heat recovery."

Continuing the discussion of Run 13, the Seventh Annual Report, page 10, notes:

> "The design capacity was not obtained for several reasons. Maximum steam condensing temperature in Effect 1 was purposely limited to a value lower than design, resulting in an upper limitation to the brine flashing range. The steam jet ejectors failed to function, which necessitated use of the reciprocating vacuum pump. This, coupled with the fact that the cooling water temperature was 78^{o} versus the design value of 70^{o}F limited the lower temperature of the flashing range to a value approximately 14^{o} greater than design. Thus, the overall brine flashing range (and consequently overall plant production) was significantly less than design value."

The Seventh Annual Report continues (paragraph 2-16, page 10):

> "There was a significant difference between the measured and the calculated production rates. The calculated value is based upon the measured extraction ratio (calculated from the Plant inlet and outlet brine concentration factors) and the seawater feed rate, less the estimated vents. The calculated value is greater than the measured value by 110,000 lbs per hour. Since the measured rate is the sum of several measuring devices which had not been calibrated, the calculated rate is probably a more valid value. Several factors adversely affected the performance ratio in Development Run No. 13. First, and most important, was the absence of the preheater in Effect 12 (HX-312). Early in the Development Run this exchanger experienced tubing failure, and the unit had not been replaced at the time the data were taken. A second factor contributing to reduced economy was that most of the new vessels and pipe lines were not insulated until after that data were taken; a third was unrestricted venting from the modular Effects and an uncontrolled flow of steam to the Deaerator-Decarbonator."

The Annual Report also noted that at the end of the Development Run, two problems — both a matter of pipe sizes — were identified. One was the size of the condensate transfer line from the original 12 effects into the new 5-effect module, and the other was the diameter of the internal flash tank condensate transfer lines in the module were not big enough. As a result, there was a blockage in the condensate flow, and consequent loss of heat transfer surface.

In regard to the "calculated" production rate cited in the 7th Annual Report, paragraph 2-16, if one applies this to the "one-line average" of Run 13, the difference is approximately 36,400 pounds of product water per hour. A performance ratio, based upon a computed production quantity divided by metered steam, would be 12.73, a value that is a better indicator of the capability of the 17-effect plant than the one reported (11.85).

The above quotation from the Annual Report indicates the low production of water during this run to be due to equipment failure (steam ejector), a reduction in plant operating temperatures, and the fact that operation occurred before all meters and instruments were properly calibrated (which resulted in uncertain data). In view of this latter admission and the incorrect pipe size problems involving condensate transfer, it seems reasonable to only "note" the data for Run Number 12 and not place too much reliance on the results when assessing the operational capability of the 17-effect design.

The overall averaged parametric values for both plant configurations and the respective design values are shown in Table 9.2.

Although the Freeport Demonstration Plant was operated for relatively short periods of time (as contrasted to a pure production plant), and although each operating period had a specific set of objectives, including research and development as well as production, it is concluded that the plant was capable of approaching or even exceeding the original design parameters. For example, as a 12-effect plant (there were more runs in this configuration), the overall gross average production rate (approximately 348,500 pounds per hour) and the steam rate (32,280 pounds per hour) were very nearly at the design level, whereas the SWF rate (475,200 pounds per hour) was well below the design point. The performance ratio (gross) was 10.97 as compared with a design value of 11.5, or 4.45% low. The computer ER was 0.671 versus a design value of 0.71, or 5.5% low. Also, as a result of operating experiences, it was found that the plant design temperatures could be exceeded safely without scaling forming.

Table 9.2

Summary of Plant Operations

Plant Parameter	Design Value		Averaged Operating Achievements	
	12-Effect	17-Effect	12-Effect	17-Effect
Number of Development Runs[1]	—	—	12	4
Plant Capacity, Lb/Hr				
Net	350,000[2]	350,000[2]		
Produced (gross)			348,500[3]	331,400[3,4]
SWF, Lb/Hr	490,000	510,000	475,200	471,900
Steam Rate to Effect 1 Lb/Hr	33,000	25,000	32,280	25,540
Temperatures °F				
Steam to Effect 1			262	279
Brine Out			122	101
Delta-T			140	178
Performance Ratio: Lb. water/Lb. steam				
Gross	11.5	14.8	10.97	13.2
Extraction Ratio: Lb. water/Lb. SWF				
Gross	—	—	0.732	0.701[4]
Computed	0.710	0.680	0.673	0.661
Kw-Hr/M Lb SWF	0.735	0.883	0.818	1.112
Lb Water (gross/Kw-Hr)	0.972	0.778	0.861	0.628
Total Operating Hours	7920	7920	13,431.5	9,086.6
Per Cent Operating Time	90.5%	90.5%	92.83%	86.54%

Notes:
1. Formally designated Development Runs.
2. Capacity given as "Net" or "Nominal" in design value tabulations.
3. Operator Data Sheet only provides "Total Pounds, including condensate". Figure is gross value.
4. Includes data from Run No. 13 which had a very low production rate.

There are similar indications to be noted for Runs 13 through 16, when the plant was operated in the 17-effect configuration. The overall gross production, while not as good as for the 12-effect plant, was approaching the design point (331,400 lbs/hr, including Run 13; 344,750 lbs/hr, excluding Run 13). Conceivably, the design point could have been reached had the plant been operated more. Likewise, the performance ratio was indicating improvement with each run.

The performance ratio, according to Table 5-1, page 76, Fifth Annual Report, is a function of the number of effects. This is also indicated by Table 2-3, page 27, Sixth Annual Report, which listed this ratio for the 12 effects as 11.5 and for the 17 effects as 14.2. While the 17-effect plant did not achieve this design point (14.2), it would appear, from the trends that may be seen in the figures, that in additional runs this design parameter might have been reached.

The largest change that occurred with the advent of the 17-effect plant operation was the change in electrical power demand. The average connected load for the 12-effect plant was 400 kilowatts, whereas the average load for the 17-effects was 525 kilowatts. The power demanded by both configurations exceeded the respective design values of 360 and 450 kilowatts; hence the anticipated load was underestimated more for the 17-effect plant.

When evaluating the reliability of the Freeport Plant, it is necessary to define the "operating period" and the period when the plant is down solely for the preparation of the next Development Run. By definition, the operating period was taken to be that period of time from the beginning of a Development Run to the time the plant shut down following the completion of the objectives for the run. Admittedly, this period does not include <u>all</u> maintenance that was accomplished; for that reason, the decision could be expected to be controversial. The resultant reliability may well be on the optimistic side.

Using the above definition of operating period, the Freeport Plant exhibited an overall reliability of approximately 90% for the operating periods from February 6, 1964 through May 9, 1969. The initial Development Run was the best operating period since the plant did not falter once. The run with the poorest operating record was Run Number 5, in which the plant was down 624.25 hours in the period of 1736 hours, resulting in an operating percentage of 64.1%. The longest sustained runs were Numbers 9, 15 and 16 at 3178, 3746.75, and 3288 hours respectively. The percentages of operation were 92.2%, 83.5%, and 94%. The 12-effect plant had an overall operating percentage of 92.83%; for the 17-effect plant the overall average was 86.54%. If this type plant were expected to operate 330 days per year (a factor considered when estimating the cost of a "normalized" plant), or at 90.5% reliability, the 12-effect plant can be said to have achieved this objective, while the 17-effect was off by 4%.

Considering the fact that this plant was expected to accomplish engineering (research and development), operational and economic objectives, this operating record is belived to be exceptional.

An item of significance to the operation of the Freeport Plant but which occurred prior to the Development Runs was the impact of Hurricane Carla. This event took place on September 9-13, 1961, and left the plant inundated with six feet of seawater.

In a complex plant such as the Freeport Plant, with its critical data sensing and control instrumentation, flooding by seawater, with accompanying silt and debris, is almost as serious a situation as can be imagined. There is the immediate and direct water damage; the time-dependent damage such as electrical insulation deterioration, salt corrosion, and erosion and scarring of bearings and surfaces due to silt; and the general havoc that is wrought upon delicate instruments.

Under a "crash program," the contractor was able to restart the plant on September 18, 1961, just 5 days later. Operations under emergency procedures continued until September 22, when the P-4 pump motor shorted out and caused the plant to be shut down, as the alternate motor was not available.

In retrospect, the recovery from Hurricane Carla (5 days to the first restart and 19 days to the restart for resumption of the operational program) is complimentary to the original process design and plant construction. The operating contractor deserves a share of the plaudits for his basic knowledge of the systems embodied in the Freeport Plant, his maintenance knowhow in a situation demanding extreme rehabilitation, and his capability to apply and administer his managerial talents efficiently.

The Office of Saline Water, in this situation, held the key to the successful recovery operation. OSW wisely directed the recovery be accomplished and then provided the cooperation and assistance required to fulfill this objective — and example of a successful industry-government relationship to achieve a common goal.

The review of any operation or project invariably triggers ideas and/or concepts that seem to offer improvements to the method of conducting the operation or a similar undertaking. Such has been the case of the review of the Freeport operations. These ideas, thoughts and/or concepts are briefly outlined in the following paragraphs. It is hoped, if these expressions are not deemed practical, that useful ideas in turn will be suggested.

Since the plant was, in fact, operated to gather research and development type data (engineering), the Operator Data Sheets might well have included data as to the accumulation of scale, in what location, under what conditions, and amount. A bar graph showing amount of scale deposited versus Development Run would complement Figures 9.1 and 9.2.

The Operator Data Sheets include information as to the amount of chemicals that were used per day. These chemicals were used for scale control purposes. The results of this addition, included as a comment on the Data Sheets, would have been informative to the reader.

Often times, the pressure to "finish the work" results in errors. Such may or may not have been the case. Certain columns of the Operator Data Sheet, Run 10A, suggests the data may be questionable (see the discussion earlier in this report). Also, it is obvious that more than one person worked in the preparation of the annual reports. For example, the Fourth Annual Report, Table 5-6, page 90b, notes 4707.16 pounds of scale accumulated in Run Number 6. Yet, the writer of the Annual Report (page 78, paragraph 5-22 and page 92, paragraph 5-58) expressly stated the opposite.

The Operator Data Sheets appear to have been prepared for a truly "production plant," whereas the Freeport Plant was operated more nearly as an engineering tool. Since this was the modus operandi, it would have been appropriate to include in or with the Operator Data Sheets, data on the significant critical parameters on which the plant design was based. For example, the "installed heat transfer surface" is a parameter listed in Table 2-3 Design Comparison, 12 and 17-Effect Plants, page 27, Sixth Annual Report, Freeport Test Facility, Freeport, Texas. If these parameters are sufficiently important and necessary that a tabulation is prepared for comparison purposes, it seems that each parameter would be sufficiently important to include these data with the other pertinent operating information.

With respect to the extraction ratio, the value obtained by the use of the formula is necessary in the design of the plant and informative during operations. After the plant has been constructed, the question really becomes one relating to metered values — metered net production divided by metered SWF. The value thus obtained compared to the theoretical value used in the design, and the computed value obtained from the actual concentration factors would be one facet of data that would tell how well the plant is doing: how well the design was achieved. If the input steam condensate from Effect I were metered, the NET production could be determined by a subtraction of this datum from "Total Production including Condensate."

From the viewpoint of demonstrating a production capability, it might be appropraite to operate the plant as a production facility for a sufficient time period, with emphasis on costs following a particularly favorable engineering operating period; for example, following Run Number 6.

Throughout the analysis of operations, it was noted that this component or that component was the "limiting" factor on the operation. Since this analysis was looking at the operations as reported, a search was not made to be certain these bottlenecks were eliminated. It was noted that in one instance, a bottleneck was identified but not corrected until after the next Development Run. Obviously, the problem found impacted the data taken to satisfy the objectives of the succeeding run.

One objective of the operation of the Freeport Plant was the determination of the cost of water produced. This information was computed, but for a "typical month" and/or on a "normalized" basis. Neither data was related to the Operator Data Sheets. This information might well have been prepared for the duration of each development run, with data on all pertinent parameters, and included in a packet with the Operator Data Sheets.

It is realized that such a packet is very apt to become voluminous if an overall plan is not prepared. It is envisioned that appropriate inclusions would be important design parameters, their measured/metered values as these parameters are changed or occur during operations, usual data for a production-type operation if not already included, and production cost information over the operational period. These data would provide the basis for an evaluation of the facility design, the areas of strength and weakness, and costs over the full operating period.

B. AUXILIARY TEST UNIT OPERATIONS

The Auxiliary Test Unit (ATU) was conceived, designed and installed at the Freeport Test Facility to be a research tool in which the threshold conditions for scaling in Effect I could be determined without exposing the main and plant. This unit was erected during the second quarter of fiscal year 1967, and initial operation, which began on January 17, 1967, was completed on February 5, 1967 (Sixth Annual Report, Freeport Test Facility, Freeport, Texas, page 9). This reference further states, on page 28, with respect to the purpose and objectives of the ATU:

> "The ATU is provided with a control and continuous recording system for the purposes of studying high temperature evaporation of seawater. The primary objective of this unit is to identify the calcium sulfate (anhydrite) scaling conditions without exposing the main Plant to these conditions. A secondary objective is to study high temperature and enhanced-surface heat transfer."

On page 96, this annual report further elaborates on the utility of the ATU by stating:

> "The ATU can be used to:
>
> "A. Perform scaling tests at different temperatures and operating conditions:
>   1. Test on normal seawater.
>   2. Test on scale inhibitors.
>   3. Test on feeds with reduced ion concentration.
>
> "B. The ATU can be used to perform heat transfer tests:
>   1. Tests different tube materials.
>   2. Test tube configuration and also enhanced or extended surface tubes.
>   3. Test miscellaneous items such as vent configuration, feed brine distribution, steam side baffling, etc."

The ATU was physically related to the main plant in such a way as to permit its operation to be in parallel or in series with Effect I, and was fed from the Dow steam line after the station meter position.

Operational data, per se, on the ATU was not found in the annual reports or as separate reports. Therefore, a review of the ATU operations in a manner similar to that provided on the main plant is not possible. However, the annual reports for fiscal years 1967, 1968 and 1969, as well as the report entitled "Auxiliary Test Unit," were carefully examined for operating comments.

As indicated above, the ATU initial operation occurred during the period January 7 to February 5, 1967. The Sixth Annual Report indicates that operation was hindered by "erratic control." In the operation of the ATU, "stabilized control will be essential to obtain useful data relative to the temperature and concentration conditions where $CaSO_4$ (anhydrite) scale deposited on falling-film heat transfer surfaces" (paragraph B.2., page 58, Sixth Annual Report). The ATU operating condition became stabilized near the end of Development Run 12; however, sump temperatures as high as $290^{O}F$ had already been reached with concentrations as much as 1.2 times normal seawater. These conditions exceeded the threshold for scaling; therefore, it was not surprising that evidence of scale was found in the tubes. The Sixth Annual Report also stated (page 10, paragraph 2-28): "The ATU proved itself capable and valuable as a research and development instrument, fully justifying extensive future testing."

The main plant was shut down upon completion of Development Run 12A in February 1967, for the purpose of installing the 5-effect module. Operation of the ATU was deferred until Development Run Number 14 which began in September 1967, and the results were reported in the Seventh Annual Report (FY-1968), Freeport Test Facility and Vertical-Tube-Evaporator Test Bed Plant, Freeport, Texas.

The ATU was operating during Development Run Number 14 with the specific objectives as outlined in the report "Auxiliary Test Unit" (K.S. Campbell, July 1968, page 1), and briefed as follows:

> "Operate the ATU at sump brine temperatures of $272^{O}F$, $277^{O}F$ and $282^{O}F$ for 500 hours at each temperature. Obtain overall heat and material balance data for the ATU at each temperature."

The operation of the ATU was to establish reliable scale information versus temperature data.

The initial operating period was to be at $277^{O}F$. The sump temperature varied from a low of $274^{O}F$ to a high of $281^{O}F$ with the average at $276^{O}F$. This run was terminated after 219 hours when it was necessary to shut down the main plant. The ATU was inspected and found to be scaled.

The ATU was retubed with the same type tube and operated at $272^{O}F$ for 368 hours before shutdown. The ATU operation was continuous and smooth. Again, scaled tubes were found. In both cases, the scale was calcium sulfate anhydrite.

Both operational runs experienced unit upsets of a short period duration. The probable cause was sporadic instrumentation failure. The specific relationship of scale versus temperature was not established.

Several mechanical and instrumentation changes were made after completion of Development Run Number 14. The ATU was not started until June 26, 1968, and the results of this operation were reported in the FY-1969 Eighth Annual Report.

After the ATU operation during the Development Run Number 14, aluminum-brass tubes were installed. The operation of the ATU during Development Run Number 15 closely approximated the conditions of Effect I, with sump temperatures of 268°F and 273°F, both with aluminum brass tubes, and 273°F with stainless steel tubes. Each of the three ATU runs were close to 500 hours duration (500, 476 and 458 hours respectively). There was no scale formed during the operation at 268°F. After an unusually smooth operation at 273°F the tubes (aluminum brass) were found to be heavily fouled with calcium sulfate scale.

The ATU was retubed with 316L stainless steel tubes and then operated for 458 hours during which one shutdown for an interim inspection and two serious units upsets occurred. At the interim inspection, scale was noticed, but the amount found did not affect the heat transfer rate. A 1.84 inch gage ball easily passed through all tubes of nominal 1.93 inch I.D. At the end of the run, the 1.84 inch gage ball still passed through the tubes; however, a 1.90 inch ball penetrated each tube about two feet. The scale had the same appearance as before and "some bigger patches of continuous scale" were found. An estimate was made that approximately 5% of the surface was covered with scale, but it was not enough to impede the heat transfer.

Data for a heat and material balance during each run was taken.

The conclusions from these operations were: "that at these conditions of temperature, pressure, and concentration factor the incipient scaling point is in the neighborhood of 273°F."

The Conclusions and Recommendations, Seventh Annual Report (R&D Progress Report Number 479), included the following comments:

Paragraph 5-10, 3, page 200, (on Development Run Number 14):

> "The Auxiliary Test Unit (ATU) operated successfully for long periods of time. Significant progress was made toward defining the anhydrite scaling point."

Paragraph 5-11, 7, page 201 (on Development Run Number 15):

> "The Auxiliary Test Unit (ATU) was significantly improved by several mechanical changes. The operation of this unit was very stable."

Paragraph 5-15, 1 & 4, page 202 (on Recommendations):

> "1. Test scale inhibitors in the Auxiliary Test Unit (ATU).
> 4. Study heat transfer performance of special Tube modifications and nozzle configurations in the ATU."

For Development Run Number 16, the ATU was retubed with Phelps Dodge spirally enhanced tubes — 14 feet long, 90-10 cupro-nickel. Approximately six runs were made with these tubes. Data was taken on heat transfer and the effectiveness of the type of weirs feeding these tubes.

As previously noted, operational data for the ATU was unavailable. However, Development Report Number 9, reporting on Development Runs 15 and 16, contained selected operational data in the form of graphs and abbreviated ATU heat and material balance applicable to Development Run 15. These data are reproduced to illustrate the operation of the ATU.

Figures 9.6 and 9.7 (reproduced from Development Report 9, Figures G.9.a and G.9.b) were obtained by plotting 4 points per day for each parameter as taken from the operator's log. This information is, therefore, raw data and must be viewed as showing a trend, rather than as reduced data. The following comments apply to these plots.

- Steam Flow to Steam Chest and Product Steam Flow — On August 5, 1968, an adjustment was made to the meter and sensing device to obtain readings in the more sensitive part of the meter range. Therefore, the apparent change in the graph does not represent a change in parameter. The plot is interpreted to mean the parameters were under control within the preset operating range.

- Chest Pressure and Sump Pressure — These parameters are roughly consistent with their respective temperatures. The slight change in chest pressure (34.6 to 36.5 psig) corresponds to the increase in chest temperature (280 - 281°F to 282 - 283°F).

- Brine Feed Inlet Concentration Factor — Blending of the plant brine output with the SWF was practiced to maintain an input concentration

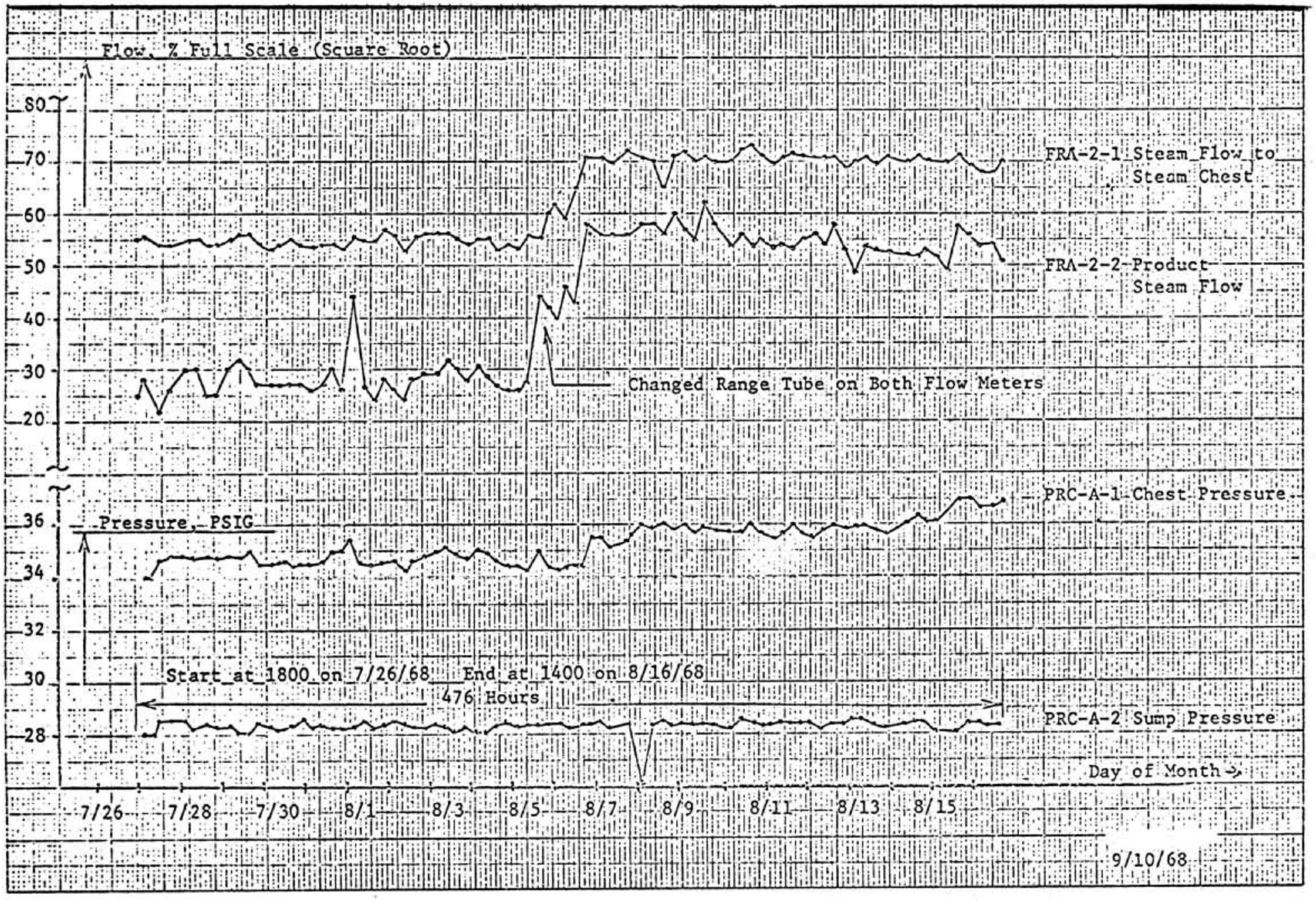


Figure 9.6. ATU Process Variables Plotted Against Time, Run 15-2, 273°F Sump Temperature with Aluminum Brass Tubes

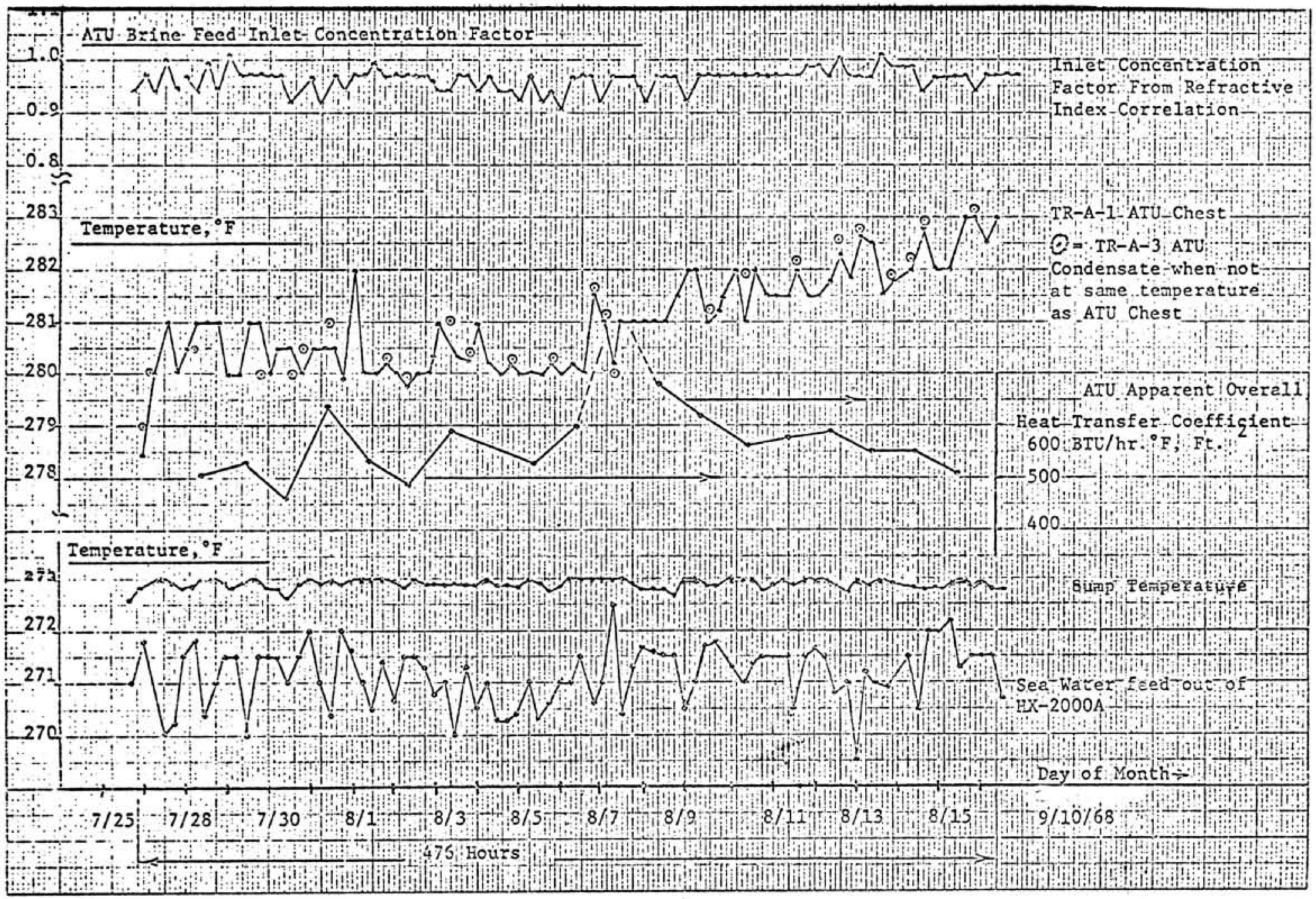


Figure 9.7. ATU Process Variables Plotted Against Time, Run 15-2, 273°F Sump Temperature with Aluminum Brass Tubes

factor (CF_i) approximating that of normal seawater. For Run 15-2, the desired range was 0.94 to 0.99 inclusive. The plotted data shows reasonable success in maintaining the CF_i within the desired limits.

- Steam Chest and Condensate Temperatures — These temperatures will normally be close together, indicating satisfactory operation of the ATU desuperheater and that the ATU chest is operating isothermally. The increase in temperature during the last half of the operating period is consistent with and indicative of scale formation.

- Overall Heat Transfer Coefficient — These data were calculated from the operator's log sheet, and are probably good to 15%. The overall heat transfer coefficient average was about 550 Btu/hr, ^oF, ft^2 over the run. The initial value was about 500 Btu/hr, ^oF, ft^2, raising to a value in excess of 600 Btu/hr, ^oF, ft^2, for the period range at the end of the run. The initial rise in overall heat transfer coefficient correlates with the rise of temperature in the chest and the formation of scale.

- Sump Temperature — The plot of this parameter indicates the ease of control within the prescribed operating range.

- Seawater Feed Temperature out of HX-2000a — This temperature averaged about 271^oF. The fluctuations of +1^oF about this average may be, in part, due to fluctuations of SWF temperature to HX-2000a.

Table 9.3 is a summary of heat and material balance data for Development Run 15.

From the above brief recapitulation on the operation of the ATU, it would appear that, within reasonable limits, the condition for the onset of scaling in Effect I is in the vicinity of 273^oF brine sump temperature. This was one of the objectives of the ATU. Three types of tubes — 316L stainless steel, aluminum brass and Phelps Dodge 90-10 cupro-nickel — were tested, as were two types of weirs — the SVL and spray nozzles. It would appear that the flexibility of the ATU was indicated.

Although truly operational remarks in the three annual reports discussing the ATU operation were very sparse, it would appear the usual operating problems were encountered, these problems being largely of an instrumentation and control nature. These annual reports did comment upon the "smooth operation" for "long periods of time."

Table 9.3

Development Run 15
Summary of ATU Heat & Material Balance Data

I. BACKGROUND INFORMATION												
A. ATU Run Number →	15-1			15-2						15-3		
B. Inclusive Dates →	6/26/68 to 7/17/68			7/26/68 to 8/16/68						8/31/68 to 9/20/		
C. Intent of Run →	No.1 Eff.Conditions			272°F Sump Temperature						Repeat of Run 15		
D. Type of Tubes →	7 x 2"OD x 22' Long x 16 BWG Arsenical Aluminum Brass			Same as Run 15-1						7 x 2"OD x 22'L x 20 BWG 316L Stainless Steel		
II. BASIC DATA												
A. Temperatures, °F												
1. Steam Chest Temperature (1)	273.7	275.6	275.4	278.9	279.2	280.8	280.7	281.0	280.2	283.4	281.8	284
2. Sea Water Feed In (2)	264.3	265.4	265.0	272.0	270.8	271.3	270.2	270.2	270.9	270.6	270.0	270
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
B. Chlorinity, Wt.% Halide												
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	2.0
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	2.1
C. Steam Production Rates, lbs/hr												
1. By Salt Balance	360	570	610	480	480	480	530	760	420	450	360	39
2. By Flow Meter	290	400	390	330	350	470	480	440	470	470	360	42
III. MATERIAL BALANCE, LBS./HR.												
IN												
Sea Water Feed (Metered)	11400	11000	10600	11100	11000	11000	10900	11000	10900	11100	10900	10
Steam to Chest (Metered)	440	520	520	480	500	630	650	570	550	550	560	5

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

Table 9.3
(cont'd)

III. Cont'd.												
OUT												
Brine Out Sump (By Difference)	11110	10600	10210	10770	10650	10530	10420	10560	10430	10630	10540	104
Condensate (Steam to Chest Minus Vent Out)	400	480	480	440	460	590	610	530	510	500	510	50
Vent Out (Metered)	40	40	40	40	40	40	40	40	40	50	50	5
Steam Produced	290	400	390	330	350	470	480	440	470	470	360	42
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	8.8	11.
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	0.33	0.3
D. Overall Evaporative Heat Transfer Coefficient; Btu/hr, °F, ft.^{2}	520	630	610	640	640	700	710	620	730	520	470	42

The net result is: the reviewer of these three annual reports and the report "Auxiliary Test Unit Report" can easily gain the impression the ATU did serve a useful purpose, was capable of achieving its objectives, was a flexible research tool that had not been completely exploited, and did have its ups and downs just as most mechanical/electrical devices do.

Then, on page 45 of the FY-1969 Annual Report, the following statement appears:

> "It is recommended that future work with the ATU be at Freeport discontinued (sic). The original intention was to be able to run a small 'package' evaporator in parallel with various effects in the Freeport Plant. (The actual installation of the ATU does not substantiate this statement.) Experience has shown that a little evaporator such as this required 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."

Continuing, the cited annual report states in paragraph 2.A (3), page 89:

> "The Auxiliary Test Unit objectives have been met with limited success. This unit required considerable operating and engineering attention. It should be either scrapped or shipped to Wrightsville Beach where it would receive the necessary engineering attention."

It appears from the short operating periods (approximately 587 hours during Run Number 14) and the relatively long time between these operating periods (the next serious period of operation was during Run 16), that indeed the objectives for the ATU were achieved "with limited success." The data reported is in keeping with the objectives, and has every indication of being valid.

From the reports and data available a cost analysis of the ATU is not possible — specific related costs were not found.

It can be assumed the operating contractor was closer to the operation of the ATU and its attendant problems than one reviewing reports at a later data. Hence, the contractor should be in the best position to make recommendations for future utilization of a specific item of equipment. It is appropriate and incumbent, upon a review of the subject reports, to raise pertinent questions, such as:

- Will the ATU, in any other location, require less sophisticated instrumentation?
- Have the reports given any statement or indication of an unusually high personnel requirement?
- Have the reports substantiated or given any indication of more operational problems than are normal?
- With respect to the data obtained and reported, have the reports evaluated the ATU versus the design objectives and found the operation of the ATU to be inferior to the design specifications?

Questions such as these should be evaluated, along with the worth of the ATU to the overall program, before arriving at a decision regarding the Auxiliary Test Unit.

From the information contained in the referenced documents, it has been concluded, with respect to the ATU, that:

- The ATU is a valuable research tool.
- The ATU did demonstrate achievement of its primary and secondary objectives, although the supporting data is limited in quantity. (It indicated threshold scaling conditions in Effect I and the research flexibility of the unit.)
- Full utilization of the ATU has not been demonstrated.

C. RELATIONSHIP BETWEEN HEAT TRANSFER RATES, SCALE CONTROL TECHNIQUES, AND PROCESS VARIABLES

The effect of process variables on heat transfer in an LTV evaporator was discussed in Chapter 4, and the effect on scaling problems was discussed in Chapter 5 of this report. A brief discussion of the interrelationship between these three quantities is given here.

Process variable changes or upsets are unavoidable in any plant. This is so because all factors influencing the process behavior are not under the operator's control. For example, temperature of raw seawater feed or its salinity cannot be controlled by the plant operator. Even though problems of this nature can be minimized by proper design and automation (e.g., raw seawater salinity could be controlled by proportional blending with concentrated brine blowdown which comes out of the last effect) it is still important to know the nature of the changes produced when plant parameters fluctuate from their steady state values. On the other hand, the change of parameters may be introduced purposely to meet certain requirements. For example, to increase plant capacity it will be necessary to raise the first effect steam condensing temperature as well as the seawater feed rate to avoid scaling from overconcentration.

The important process variables to be discussed in this subsection are indicated on the simplified block diagram of the Freeport LTV process depicted in Figure 9.8.

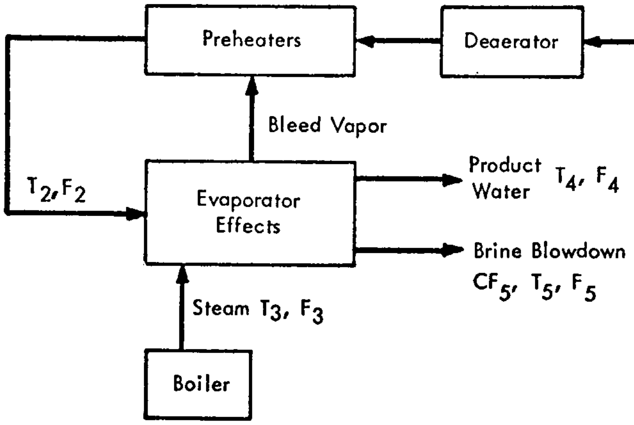


Figure 9.8. Simplified Block Diagram of the Freeport LTV Process

Essentially, the Freeport process consists of first heating the raw seawater feed with product water to facilitiate easy removal of O_2, CO_2, etc. in the deaerator, for corrosion prevention and alkaline scale control. Vapor bled from the evaporator then preheats the feed to approximately the required saturation temperature (T_2) of Effect I. Steam is supplied to this effect at a temperature of T_3 and flow rate of F_3. Product water, collected from all effects, is at a temperature of T_4 and has a flow rate of F_4. Rejected brine has a temperature of T_5, a flow rate of F_5 and a concentration factor of CF_5. T_1, F_1, and CF_1 represent similar quantities for the raw seawater feed.

For the purpose of this discussion, it is assumed that only one variable is changed at a time and that the process is operating in a normal stable manner free of scaling problems. Consider, first, an increase in the steam condensing temperature T_3. Depending on the extent of this increase, the heat transfer rates go up because of the increase in the temperature driving force ΔT as well as an increase in the overall coefficient of heat transfer U. An increase in the rate of heat transfer means greater removal (and production) of water from a fixed input rate of feed. However, greater water extraction also means greater concentration of salts, thus leading to the possibility of scaling. If scaling does occur in any effect then it drastically reduces the heat flux and hence the total plant production.

Consider now an increase in temperature T_1 of the seawater feed. Since all preheating of this feed is done by bleed vapor and/or condensate from the evaporator, its maximum temperature cannot exceed the maximum temperature of the evaporator bleed vapor or condensate — and this maximum temperature is that of the first effect vapor or condensate. Since changes in feed temperature are small, there would be only slight effect on heat transfer rates or on scaling. However, two side effects of this temperature increase are noteworthy — first, the last effect would experience an increase in pressure (reduction of vacuum), and unless the mechanical vacuum system would take up the extra load, the total plant driving force ΔT would reduce and the plant capacity would reduce; secondly, there would be a gain in the performance ratio since an increase in feed temperature represents an increase in enthalpy input to the system.

Now consider an increase in the seawater feed rate F_1 while other parameters are held constant. An increased bleed vapor demand in the preheaters will cause the pressures and temperatures to drop in all the effects, which will result in reduced heat transfer rates and reduction in scaling tendencies. Both extraction and economy ratios will tend to decrease.

Finally, let the concentration factor CF_1 of the feed brine increase. Heat transfer rates will be essentially unaffected until the concentration factor is increased so high that the solubility limit of calcium sulfate anhydrite is exceeded corresponding to the first effect temperature. Scaling will then ensue. The anhydrite scaling threshold for normal seawater ($CF = 1.0$) has been found to be around 268°F. Gypsum scaling in low temperature effects can be controlled up to a concentration factor of 3.2 with the use of polyphosphates. The compounds are effective only up to 170°F. Since higher heat transfer rates are obtained at higher temperatures, further research is indicated in this area to raise the scaling thresholds.

Chapter 10

ECONOMIC HISTORY

A. ORIGINAL PLANT COST

The cost of the original plant at Freeport, Texas, has been found in four different documents. The cost differs in these four sources and reconciliation is not immediately apparent. These documents are:

- "Final Report, Saline Water Conversion Plant at Freeport, Texas," prepared (presumably) by Chicago Bridge and Iron Company. The stated total cost was $1,387,700.00.
- "First Annual Report, Saline Water Conversion Plant, Freeport, Texas," R&D Report Number 71 prepared by Stearns-Roger Corporation, page III-1, B-Capital Investment: "The total capital investment cost for the facility... $1,265,387.50."
- "Summary of Performance Sea Water Desalting Demonstration Plant, Number 1, Freeport, Texas." From Table 7-1 Actual and Adjusted Investments, including A-E Services, Construction Costs, Land and Spare Parts... $1,371,449.00.
- "A Study of Large Size Saline Water Conversion Plant," March 1963, prepared by Bechtel Corporation. From Table IV, page 30: Total Capital Costs, including Start-Up — $1,569,000.00.

The first document listed above is a tabulation of costs of each of 44 bid items of Specification #195 plus start-up costs. The "listed costs are our actual costs covering full overhead and administrative expenses but no profit." This document was submitted in compliance with Addendum Number 1, Specifications #195, dated March 29, 1960.

The First Annual Report, the second document listed above, quotes the "Amount of contract this date" as taken from "Order for Change Number 3, Contract Number 14-01-001-195, Specification #195, dated 8-28-61." This report also notes equipment deletions of $16,290 for clarifier equipment and $54,000 for 6 of 14 heat exchangers removed, and equipment additions of $34,622. The additions and deletions occurred in FY-1962.

The "Summary of Performance" report, the third listed document, reviews the operation in mid-FY-1965, and the economics of the plant. In that report, Table 7-1, Acutual and Adjusted Investments, tabulates construction costs and capital equipment additions and deletions through the first half of FY-1965. The construction cost stated above assumes (as does Table 7-1) that the contractual amount, Order for Change Number 3, is $1,265,388 for plant construction, and adds $66,950 for A-E Services, $27,290 for land, and $11,821 for spare parts. The total cost of construction was listed as $1,371,449.

The last listed document, "A Study of Large Size Saline Water Conversion Plants, March 1963" included the following tabulation in Table IV, Capital Cost of the Demonstration Plant, on page 30:

"Cost	$1,380,000
"Engineering	167,000
"Start-Up	22,000
	$1,569,000"

The amount, "Cost — $1,380,000", seems to relate to the cost shown by CB&I in the first listed report, less the CB&I start-up cost. The engineering cost differs from the value used by Stearns-Roger Corporation in the report "Summary of Performance...." Likewise, the start-up cost differs from the amount used by CB&I; however, CB&I admitted their period of operation was very short and that therefore, the CB&I cost figures do "not reflect a true breakdown of plant operation over a longer period of time."

The total cost given by the CB&I document may well be the actual cost to construct the Freeport Demonstration Plant. A letter from the Director, Office of Saline Water, to the Chief, Fiscal Section, Division of Administrative Services, on the subject of "Final Payment — Chicago Bridge and Iron Company Contract Number 195," dated January 17, 1963, quotes the construction cost as $1,265,387.50. This amount represents the contract cost of Freeport, which is slightly less than the amount stated by CB&I. It is reasonable to accept the actual capital construction investment of the Freeport Plant, including start-up costs as $1,387,700. If one desires the total investment at the time of initial start-up, additional costs, such as architect/engineering services, land, etc., must be added to the capital construction cost. This figure might be $1,569,000.00, the amount shown above.

B. COST OF DESIGN MODIFICATIONS

The review of available documents, principally the annual reports, yielded very little information as to the cost of plant modifications. These reports did review the technical aspects of the changes. Annual Report Number 6 listed $838,102 for maintenance and modification, noting that the bulk of R&D funds went into the major modifications — installation of the 5-effect module and the ATU. On page 4 of the report for FY-1968, the maintenance and modification costs were lumped together as $151,445.

The costs of additions and deletions mentioned in the reviewed documents are tabulated below:

Fiscal Year	Source	Addition	Deletion
1962	1st Annual Report	$34,622.00	$16,290.00 54,000.00
1963	2nd Annual Report	83,413.88 34,320.00	74,024.00
	Summary of Performance		
1962	19 Feb., 1965	34,622.00	65,455.00
1963	"	83,414.00	24,859.00
1964	"	16,052.00	9,350.00
1965*	"	9,578.00	

---

*First half only.

There is good agreement among the above reports. The additions listed in the annual reports are easily identified in the Summary of Performance. The $34,320 found in the 2nd Annual Report was for spare parts, supplies, materials, etc. The $54,000 deletion in the First Annual Report was for the six heat exchangers that were deleted. This cost, plus the cost for spare parts lost in Hurricane Carla, is included in the 2nd Annual Report. The sum of deletions listed in the annual reports ($16,290 plus $74,024) equals the sum of deletions for FYs-1962 and -1963, listed in the Summary of Performance.

Subsequent annual reports omitted the dollar cost reference to plant changes and modifications although they did include pertinent narrative discussions. The Summary of Performance report mentions modification costs for FY-1964 and for the first half of FY-1965.

C. COST OF ROUTINE MAINTENANCE

The maintenance costs, compiled from Annual Reports Number 1 through 7 inclusive, are given in Table 10.1. These costs are listed as "Labor" and "Other." The costs are totalled by annual report. The Eighth Annual Report did not include a maintenance breakdown.

The variation of these costs from year to year, in both the totals and in individual cost items, indicates the possibility that some modification costs may have been included in the maintenance category. On the other hand, it is reasonable to expect the maintenance costs for FY-1968 to be greater than in previous years, if only because the plant during this period was operating as a 17-effect plant instead of as a 12-effect plant. In other words, more mechanical equipment generates a larger maintenance cost.

The principal areas of maintenance are pumps, evaporators, general plant maintenance, and piping. These maintenance cost centers are logically the ones requiring the most attention; they involve mechanical moving parts, exposure to hostile environments, and extensive manual effort for installation. All these factors are usually associated with higher maintenance costs.

The various cost centers shown in Table 10.1 were taken from the respective annual reports. For this reason, there are cost centers entitled "General Maintenance" and "General Plant." Inasmuch as the annual reports did not include a definition of these terms, these cost centers are shown separately in Table 10.1. The Seventh Annual Report lends some credence to this viewpoint in that both cost centers are shown as "active" centers in FY-1968.

There does not appear to be a correlation between Table 10.1, "Total Maintenance Cost," and Table 10.2, "Operational Cost Summation, FY-1962 through FY-1968." The respective data were taken from different sources and there was no specific explanation of the items included in each total.

Table 10.1

Maintenance Costs, Freeport Demonstration Plant, FY 1962-1968

Anual Reports	1		2		3		4		5		6		7		8*	
Fiscal Year	1962		1963		1964		1965		1966		1967		1968		1969	
	Labor	Other	Labor	Other	Labor	Other	Labor	Other	Labor	Other	Labor	Other	Labor	Other	Labor	Other
Supervision & Engineering			7668**													
Instrumentation	4762	2657	4433	797	3379	1754	4499	1102	3809	1000	5293	2705	8863	3943		
Evaporators	3933	1250	4637	24526	5414	3656	10554	1730	5565	2953	1090	1401	5471	2646		
Heat Exchangers	3909	1261	621	75	1801	1437	5338	2194	3823	59	5836	1863	6574	18336		
Other Vessels & Tanks	535	99	1461	1004	1428	383	1206	317	684	137	717	673	1489	1454		
Miscellareous Equipment	516	276		4	703	1260	422	905	138	86						
Pumps	4917	3173	3110	3229	3332	2730	6690	7629	6932	4637	6011	9237	7707	17810		
Electrical	217	2677	25	474	538	978	7	633	1175	815	406	2144	1496	1297		
Piping	1404	2874	1670	4793	3899	6200	6653	8529	8184	9544	6137	12950	5329	4007		
General Maintenance	2713	1027	7076	6196	8854	165	19512	13460					8157	38537		
General Plant	6936	485							3807	810	3037	1401	3337	996		
Buildings	2191	2136	7360	1389					4603	936	4513	983	6210	5796		
Tools		3013							564	441	230	447	578	871		
Distribution System									764	687	122	740	148	17		
SUBTOTAL	32060	20928	38061	42487	30348	18563	54881	36499	39316	22019	33392	34544	55359	95710		
TOTAL	52988		80584		48911		91380		61335		57936		151069			

* Maintenance expense not given in 8th Annual Report.

** Supervision & Engineering specifically given in 2nd Annual Report.

MAINTENANCE COSTS
FREEPORT DEMONSTRATION PLANT, FY 1962-1968

Table 10.2

Operational Cost Summation, FY 1962-1968

Fiscal Year -	1962[1]	1963[2]	1964[5, 12]	1965[6]	1966[7]	1967[8, 13]	1968[9, 10, 12]	1969[11]
A. Administration	$ 84,780.25	$ 67,292.81						
B. Process Operation	53,191.75	45,933.82						
C. Experimental Process Operation	2,260.82	4,216.51						
D. Building and Grounds	4,327.83	8,749.13						
E. Plant Maintenance	55,098.40	64,432.17						
Direct Expense:								
Operations								
Supervision & Engineering			$ 6,695.47	$ 6,625.32	$ 6,980.58	$ 5,004.87		
Labor			38,457.88	28,497.49	28,684.98	20,882.46		
Materials & Supplies			11,020.32	9,319.79	8,251.85	6,614.41		
Utilities			116,303.26	102,871.86	104,465.15	64,339.08	$ 75,505.38	
Total Operations Cost			172,475.93	147,314.46	148,382.56	96,840.82	121,163.19	
Maintenance								
Supervision & Engineering			4,290.72	7,728.64	8,532.94	10,580.25		
Labor			26,136.59	54,211.54	42,831.82	69,291.75		
Materials & Supplies			22,020.57	42,402.69	34,470.44	70,720.73		
Other Maintenance Expense			2,867.07	757.64		2,135.98		
Total Maintenance Cost			55,314.95	105,100.51	85,835.20	152,728.71	120,737.42	
Extraordinary; Experimental & Research Expense								
Extraordinary Labor			77.23				4,077.98	
Experimental & Research: Supervision & Engineering			5,573.62	7,455.53	6,863.80	8 215.48		
Labor			726.93	921.57	1,834.28	159.51		
Materials			1,244.30	9,008.14	4,243.78	2,033.50		
Total Extraordinary; Experimental & Research Expense			7,622.08	17,385.24	12,941.86	10,408.49	37,577.18	
Total Direct Cost			235,412.96	269,800.21	247,159.62	259,978.02	279,477.79	
Indirect Expense:								
G & A Expense								
Labor, Payroll, etc.					18,675.55	38,914.06		
Materials, Supplies, & Expense					15,208.72	9,317.14		
Miscellaneous General Expense					63,033.75	78,816.72		
Total G & A Expense			87,792.33	96,941.10	96,918.02	127,047.92		
Depreciation			69,571.15	50,682.67	53,450.78	52,329.48	43,317.18	
Total Indirect Cost			157,363.48	147,623.77	150,368.80	179,377.40	124,814.58	
Total Production Cost: Direct & Indirect	238,194.08	190,724.44	392,776.44	417,423.98	397,528.42	439,355.42	404,292.37	
Other Non-Scheduled Expenses					13,295.38	5,986.92		

Table 10.2
(cont'd)

Fiscal Year -	1962	1963	1964	1965	1966	1967	1968	1969
F. Plant Additions & Modifications[3]	$ 47,993.39	$ 51,506.93	$ 2,262.47[3]	$ 2,191.42[3]	$ 3,257.03	$ 2,052.39	$ 86,068.33	
Z Account			2,102.78	2,314.75	16,411.93	3,506.19		
GRAND TOTALS	286,187.47	242,231.37	397,141.69	421,930.15	420,492.76	450,900.92	490,360.70	
TOTAL PRODUCTION, GALLONS, GROSS	256,021,695[14]	250,183,830[15]	289,807,260[16]	234,013,230[17]	225,768,250[17]	108,682,956[17]	171,214,307[18,19]	
COST: DOLLARS PER 1000 GALLONS[20]	$ 0.9304	$ 0.7623	$ 1.3553	$ 1.7838	$ 1.7608	$ 4.0425	$ 2.3613	
TOTAL COST PER 1000 GALLONS[22]	1.4217	1.4052	1.2595					

NOTES

1. Ledger Closing Summary - Fiscal Year 1962, Grand Total includes Hurricane Carla costs of $38,535.03.
2. Ledger Summary for Fiscal Year 1963.
3. Line becomes "Major Maintenance" in FY-1964.
4. Includes such costs as: V¹ᴰ Preparation/Clean-up; Development Materials Testing; Chemicals discarded.
5. Cumulative Totals, FY to Date, Ledger Summary of June 1964.
6. Production and Cost Report for Month of June 1965, Ledger Summary. Cumulative Totals, FY to Date.
7. See No. 6. For Month of June 1966.
8. Table III, Ledger Summary for Month of June 1967, Cumulative Totals, FY to Date.
9. Reconciliation of Reimbursable Expenses and Operating Costs, June 1968, FY to Date.
10. Items that could be positively identified are shown.
11. Not available as Ledger Summary Data.
12. Major format change in FY 1964 & 1968.
13. Minor format change in FY 1967.
14. First Annual Report, Data from.
15. Second Annual Report, Data from.
16. Third Annual Report, Data from.
17. Data from Ledger Summary Report. June 1965, 66, 67, 68, Column FY to Date..
18. Date from Engineering Reports.
19. Data from Reconciliation of Reimbursable Expenses & Operating Costs.
20. Total Production Cost ÷ Total Production in thousands of gallons.
21. Table 3-2, Third Annual Report, without Abnormal Maintenance Costs.
22. Data from Third Annual Report, Table 3-2.

D. COST OF UNSCHEDULED MAINTENANCE

Hurricane Carla hit the Texas coast in September 1962. The damage caused to the Freeport Demonstration Plant certainly must be considered unscheduled maintenance. The total cost of repairs due to Carla was listed in the First Annual Report, page III-11, as $38,535.03. This amount is confirmed in the Ledger Summary for the month of June 1962.

While $38,500 is stated as the cost of hurricane repairs, the annual reports also note that certain repairs were necessary due to Carla for some period of time afterwards. The costs of these repairs were not classified as unscheduled but were included in the normal maintenance cost.

Other than the costs resulting from Hurricane Carla, there were no unscheduled maintenance expenses listed.

E. OPERATING COSTS

The financial aspects of any operation are always examined with great interest. The costs of operating the Freeport Demonstration Plant are no exception. Inasmuch as this particular facility had multiple objectives, these costs are that much more important. For a plant operating as a research tool, one would expect the actual cost of operation to be high and the cost of production to be low. Also, there are costs involved during research operation that would not occur if the plant were truly a production plant. In this situation, one must always face the question: just what are the real costs of operating a production plant?

The information available for review included the annual reports, ledger summary reports, reconciliation of reimbursable expenses and operating reports, engineering reports, and development reports. All these documents originated from the operating contractor. The annual reports and most of the development reports were later released by the Office of Saline Water for publication as Research and Development Progress Reports. Undoubtedly, these publications will become the ones most commonly referred to and quoted as containing authoritative information on the Freeport Demonstration Plant.

The annual reports discuss Freeport economics in terms of a typical month and the theoretical or normalized costs. Several annual reports also include a tabulation of actual average daily operating and production costs. Data for the typical month appears to be a reproduction of a monthly report for a period in which the production was reasonably good.

The data which was found for typical months in the annual reports is summarized in Table 10. 3. This data differs from that shown in Table 10. 2 (i. e., Cost per 1000 Gallons), which, for the most part, is based upon data submitted in ledger summaries. While the data for typical months appears to be more consistent than the data in Table 10. 1, neither suggests much hope for economically achieving water by the LTV process when compared to current costs.

The theoretical or normalized costs, on the other hand, are estimates of the cost of operating a hypothetical plant of approximately the same capacity as a production facility, and for which the initial capital investment was based on the estimate of $1, 839, 500, given in R&D Progress Report Number 72, "A Study of Large Size Saline Water Conversion Plants," Table IV, "Normalized Plant Estimates." This investment was adjusted each year, presumably according to the research changes, but the totality of change was not specifically defined. Likewise, the ground rules for estimating the operating costs are based upon R&D Report Number 72, with logical updating as necessary. The presented normalized operating costs are the best predictions of the costs of operating a production plant using the LTV process in a configuration similar to that for the specific fiscal year under consideration.

During the review of the data on the Freeport Plant, the data found principally in the ledger summaries were tabulated to show direct and indirect operating costs; operating and maintenance expenses; and extraordinary, experimental and research expenses, all of which were summed into a total production cost. Additionally, lines entitled "Other non-scheduled Expense," "Plant Additions and Modifications," and "Z Account" were included because of the interest in modifications (and their resultant impact upon capital investment) and to present grand totals (yearly) consistent with the source documents. This information appears in Table 10. 2. Once Table 10. 2 was assembled, it was apparent that with the addition of "Total Production, Gallons, Gross," the actual cost of producing fresh water from seawater could be obtained. This cost was obtained by dividing total production cost by the total production in thousands of gallons. It was noted that a similar cost was reported in the monthly ledger summaryies in FYs-1966 and -1967. These values are for the gross production, not for that amount of fresh water delivered or available for delivery to the customer.

In the Third Annual Report, Table 3-2, "Comparison of Total Production and Cost Figures for Three Year Period " (FYs-62, -63, and -64), gives the total cost of water produced. Other than for FYs-1966 and -1967, these data were not recorded in either the annual reports or the ledger summaries.

Table 10.3

Cost of Water Produced for Typical Months, as Taken from Annual Reports

Annual Report	2[1]	3	4	5	6	7	8
Fiscal Year	1963	1964	1965	1966	1967	1968	1969
Month		October	December	July	July	June	February
Total Production (in gallons)	250,183,830	34,845,900	32,734,900	28,949,720	19,741,400	25,244,641	29,900,000
Total Cost	$421,236.96	$36,668.00	$37,539.63	$34,718.05	$34,238.28	$39,310.91	$43,550.00
¢/1000 gallons	168.37	95	114.67	119.92	173.54	155.72	145.80

Notes:

Data for FY-1962 not available.

(1)Data for FY-1963 are the total yearly figures.

The reported cost of water ($4.04 per 1000 gallons) in FY-1967 is misleading and clearly illustrates the problem of using "actual cost data" from a research and development facility to determine the production costs for a production plant. It was during this fiscal year that installation of the 5-effect module and the Auxiliary Test Unit occurred. Hence, the plant was down — non-productive — for a large percentage of the year; yet certain charges continued. The ledger summary for June 1967, "Cumulative Totals, FY to Date," was the source of this information. This summary also gave the cumulative unit cost of water.

An examination of Table 10.2 for fiscal years 1964, 1965, 1966 and 1967 will give an indication of the cost centers contributing the most to the operational cost. (These are the years in which the data were sufficiently broken down to permit such an examination.) The high cost centers were:

Utilities — electricity and steam
G&A — indirect operating expenses
Labor — operating and maintenance
Materials & Supplies — maintenance
Depreciation — on capital investment

The cost of maintenance materials and supplies may well include some expenses that normally would not be production-type expenses — such as changing tube materials as frequently as was done at Freeport. Hence, this cost might be considered high, compared to such costs for a production facility. The G&A item in this tabulation includes the costs incurred by the Denver "home" office of the operating contractor. Due to the research nature of the operation, these costs may be greater than would the corresponding costs of a true production plant. The costs of utilities, labor, and depreciation are largely influenced by the process itself, design and engineering. Certainly these factors should be carefully evaluated whenever a production plant is being considered.

F. NORMALIZED PRODUCT WATER COST HISTORY

Each annual report estimated the cost of the product water from a similarly configured plant of like capacity. This normalization was based upon the output of one stream day and 330 operating days per year. A summary of these estimates is given in Table 10.4. Included is a similar estimate found in R&D Report Number 72, "A Study of Large Size Saline Water Conversion Plants," Table VI, which is presented for reference purposes (see Table 10.4). Fiscal year 1962 data was not available.

Table 10.4

Normalized Operating Costs

(Basis: One Stream Day and 330 Operating Days per Year.)

Annual Report	1	2		3		4		5		6		7		8		R & D No. 72	
Fiscal Year	1962	1963		1964		1965		1966		1967		1968		1969		Table I	
Effects	12	12		12		12		12		12		17		17		12	
Capacity, GPD		1,000,000		1,010,000		1,155,000		1,080,000		1,080,000		1,080,000		1,080,000		1,000,000	
		$	¢/1000	$	¢/1000	$	¢/1000	$	¢/1000	$	¢/1000	$	¢/1000	$	¢/1000	$	¢/1000
I. Direct Operating																	
Energy			47.3	523.51	51.84	520.93	45.11	576.48	53.38	554.63	51.35	420.80	38.96	441	40.9		40.3
Chemicals			3.6	30.12	2.98	31.14	2.70	31.94	2.96	31.41	2.91	31.41	2.95	32	3.0		1.6
Operating Labor			20.6	152.16	15.07	84.45	7.32	87.84	8.13	95.05	8.80	97.83	9.06	148	13.7		20.6
Maintenance Labor			7.1	33.78	3.34	96.40	8.34	115.34	10.68	124.75	11.55	128.39	11.89	209	19.4		6.8
Supervision				13.54	1.34	16.90	1.46	15.76	1.46	15.76	1.46	15.76	1.46	16	1.5		
Supplies			13.6	27.50	2.72	25.75	2.22	24.01	2.22	21.13	1.96	22.54	2.09	25	2.3		2.7
Total Direct Operating			92.2	780.61	77.29	775.57	67.61	851.37	78.83	842.73	78.03	717.20	66.41	871	80.8		81.0
II. Indirect Operating																	
Payroll Extras			4.6	29.75	2.94	31.50	2.72	35.03	3.24	35.17	3.26	36.19	3.35	57	5.3		4.4
G & A			8.0	53.92	5.34	59.60	5.17	63.49	5.88	63.74	5.90	65.60	6.07	103	9.5		8.0
Depreciation & Interest			35.6	409.18	40.50	364.20	31.57	325.96	30.18	325.96	30.18	340.98	31.57	341	31.6		40.8
Taxes & Insurance			9.7	110.00	10.89	100.00	8.66	93.04	8.62	84.51	7.82	90.15	8.35	99	9.2		11.0
Interest on Working Capital			0.8	9.55	0.95	10.28	0.88	9.95	0.92	10.25	0.95	9.24	0.86	11	1.0		1.1
Total Indirect Operating			58.7	612.40	60.63	565.58	49.00	527.47	48.84	519.63	48.11	542.16	50.20	611	56.6		65.3
TOTAL OPERATING COST				$ 1393.01		$ 1341.15		$ 1378.84		$ 1362.36		$ 1259.36		$ 1482		$ 1463.03	
COST OF WATER			150.9		137.92		116.16		127.67		126.14		116.61		137.4		146.3

NOTE: The cost of water per 1000 gallons is based upon the plant rated capacity.

It is noted that beginning in FY-1964 and for the following years, the normalized cost of product water was less than the reference estimate.

A check of the items included in the normalized cost of plant operations soon shows the single highest cost was for energy, or utilities (steam and electricity). The cost centers next most expensive were depreciation and labor (maintenance and operating). These three items represent from 73% to 81% of the product water cost depending upon which fiscal year one examines. In other words, to significantly reduce the cost of product water, the cost of these items — energy, depreciation and labor — must be reduced. The amount of energy consumed is basically process dependent. Since this discussion is limited to the LTV process, conservation of energy by engineering design and good operating practices is the probable means of reducing the cost of energy required. Depreciation and interest are a function of capital investment which, in turn, is dependent upon engineering design and construction techniques. The required labor for plant operation is largely determined by engineering design and equipment reliability. It is presumed that plant capacity is approximately the same as for the normalized capital costs shown in Table 10.5.

The normalized capital costs are shown in Table 10.5, which includes similar estimates from R&D Number 72 as a reference. There is, from year to year, a change in the individual items of these estimates, especially in special and standard engineering equipment, and in other process facilities relating to the plant. This change could be expected, as the research and development objectives changed somewhat from run to run. Hence, the plant being estimated each year was slightly different from the preceding plants.

In FY-1965 two estimates are given because the heat transfer area was significantly changed during the year. In FY-1967 two estimates are given — one for each of the two configurations (12 and 17 effects).

The estimate given in OSW R&D Progress Report Number 72, "A Study of Large Size Saline Water Conversion Plant," includes a list of itemized price changes, and a list of new items that adjusts the prices of certain items of the plant and adds certain items to improve operational performance.

Table 10.5

Normalized Costs, Plant Capital Investment

Annual Report	R & D #72[1]	1	2	3	4	4	5	6	6	7	8
Fiscal Year		1962[2]	1963[3]	1964	1965 October	1965 December	1966	1967 12 Effect	1967 17 Effect	1968	1969
Capacity, GPD	1,000,000		1,000,000	1,010,000	1,080,000	1,155,000	1,080,000	1,080,000	1,080,000	1,080,000	1,080,000
Operating Days per Year	330		330		330	330	330	330	330	330	330
Principal Items of Equipment ($)											
1. Special Equipment	1,221,165			1,093,900	1,093,900	855,905	909,347[5]	909,347	955,730	955,730	955,730
2. Standard Engineering Equipment	30,560			30,560	30,560	30,560	36,560	36,560	41,200	41,200	41,200
Total Principal Items of Equipment ($)	1,151,725			1,124,460	1,124,460	886,465	945,907	945,907	996,930	996,930	996,930
Process Facilities:											
Total Process Facilities ($)	517,775			517,775	449,075	449,075	449,075	449,075	470,600	470,600	470,600
Other Plant Costs:	170,000			170,000	170,000	170,000	170,000	170,000	170,000	170,000	170,000
Interest on Investment ($)				32,845	32,845	32,845	30,426	30,426	31,400	31,400	31,400
Total Other Plant Costs ($)	170,000			202,845	202,845	202,845	200,426	200,426	201,400	201,400	201,400
Total Plant Costs ($)	1,839,500		1,604,039	1,845,000	1,776,380	1,538,385[6]	1,595,408	1,595,408	1,668,930	1,668,930	1,668,930
Capital Cost per Gallon of Daily Capacity ($)	1.84		1.60	1.827	1.529	1.424	1.483	1.483	1.56	1.564	1.55
					(1.538)[4]	(1.33)[4]	(1.477)[4]	(1.477)[4]	(1.545)[4]	(1.545)[4]	(1.545)[4]

NOTES:

1. The capital costs for a 1 MGPD plant published in R & D Progress Report No. 72 forms the basis of determining the capital cost of a normalized plant.

2. Normalized costs are not given. A written, generalized account of capital investment was given in the First Annual Report.

3. Normalized costs were determined by adjusting the construction cost of the Demonstration Plant as given in R & D No. 72.

4. These values obtained by dividing "Total Plant Costs" by "Capacity, GPD". "Capital Cost per Gallon of Daily Capacity" (above) taken from respective reports.

5. Incorrect addition for "Special Equipment, Principal Items of Equipment". Hence, "Total Plant Costs" and "Capital Cost per Gallon of Daily Capacity" were in error.

6. Capital investment for December adjusted for deactivated equipment.

Chapter 11

CONCLUSIONS AND RECOMMENDATIONS

Over its eight-year period of operation, the Freeport Saline Water Conversion Plant has demonstrated not only the economic feasibility and operational reliability of the multiple effect falling-film LTV distillation process but has also contributed significantly to advancing the state of the art of evaporative desalination. It was the first evaporator ever built utilizing twelve (and later seventeen) effects in series, the first to use falling-film type operation for seawater, and the first such plant designed to operate at 250°F initially and later at 268°F without scale deposition. The Freeport Plant also achieved the rated capacity of 1,000,000 gallons per day with only 63 percent of the evaporator tubes in service, an equivalent production of 1.8 million gallons per day while operating at 265°F.

An important milestone in the Freeport Plant operation came with the addition of a 5-effect modular evaporator to the original 12-effect plant. A low operating temperature of 100°F was achieved while increasing the efficiency and economy of the plant. A number of other important plant modifications were introduced at the time of this changeover to 17-effect operation. Heat recovery from the condensate (product water) was achieved via an "all-flash" operation. This change simplified plant operation without significant loss in performance ratio. The inefficient liquid-to-liquid cross exchangers (Series 200) and its associated pumps and piping were removed so that power and maintenance expenses were reduced.

The high heat transfer capabilities of enhanced surfaces (double fluted and spirally indented) was demonstrated both in the evaporator effect heating element and the feed water preheater. The falling-film transfer rates in smooth tubes were shown to be generally correlatable by Dukler's Theory (OSW R&D Report Number 74).

A number of other engineering advances were made during the eight-year operation of the Freeport Plant. The following categorized list of conclusions is based mainly on the conclusion reported by Stearns-Roger Corporation (the plant operators) in the various annual reports, development reports, and other documents submitted by them to the Office of Saline Water.

A. PLANT DESIGN

- The original Freeport Plant design by W. L. Badger Associates, adopted without significant changes by the constructors Chicago Bridge and Iron Company, proved in general to be an efficient and reliable method of desalting seawater to obtain one million gallons per day of potable water.

- The original magnesium hydroxide sludge recycle scale prevention technique was abandoned within two days of the plant operation due to silt accumulation and calcium carbonate scale of the feed water control valve. This method, proven successful in pilot plant operation by W. L. Badger Associates, was never revived again during the Freeport Plant life.

- Because of the use of old calcium sulfate solubility data, the final concentration factor of 4.0 could not be reached in actual operation. The extraction ratio (pound of water removed/pound of seawater feed) had to be cut down from 0.75 to about 0.67. This meant increased pump capacities and power requirements. Despite these difficulties, a record high production rate of 1.21 million gallons per day was attained during the FY-1965 operations.

- Series 200 heat exchangers for preheating the seawater feed with hot condensate were removed from the plant since the heat transfer rates in these units were extremely low and power and maintenance costs were high. The eleven flash tanks, originally installed only to divert contaminated product water, were pressed into routine service with slight modification (increase) of condensate piping sizes. Vapor produced by flashing of the condensate served as the heating medium in Series 300 preheat exchangers. This additional vapor increased the performance of some of the Series 300 exchangers.

- The original liquid distributor plate in the top waterbox of all the effects proved inadequate for even distribution of seawater (brine) to all the tubes of the heating element. Triangular or V-notch weir significantly improved this situation. Swirl-vane (SVL) weir and porcelain spray nozzles were later tried and the latter appeared to be superior to the SVL weir.

- Sump-to-sump mode of brine transfer is practical and easily controlled compared to the feed forward mode. Flash-induced turbulence in the top waterbox is reduced significantly by the sump-to-sump brine transfer mode. Recirculation of the sump brine to the top waterbox (necessary in the sump-to-sump mode) is helpful in ensuring even distribution of feed brine to the tubes and thus reducing starvation-induced scaling of the tubes.

- The deaerator-decarbonator packed with Maspac FN-90 polystyrene packing (3 - 3/4 inch size) efficiently reduced the dissolved oxygen and carbon dioxide to less than 100 parts per billion and 10 ppm respectively so that corrosion was reduced and alkaline scales were entirely eliminated. The optimum conditions were found to 0.5 to 1.0°F feed flashdown at top and one pound of stripping steam per thousand pounds of brine feed.

- The monel wire-mesh entrainment separators installed in the low temperature effects were quite effective in cutting down the salt spray carryover in product water. However, if water is required for potable purposes only, it may not be necessary to use these separators.

- The multiple effect falling-film LTV process is essentially simple to control for maintaining the quality and quantity of production. For best steam economy, its rate of flow to the first effect should be on ratio control with the seawater feed rate. To avoid scaling in the first effect a temperature override should be provided on the stream flow. The whole plant can be easily automated to meet fluctuating water demand. The plant can work efficiently at capacities ranging from 100% to 25%.

B. START-UP

- Detailed operating instructions are available for Freeport-type plants. Procedures for both hot, cold, and emergency start-up and shutdown are given.

- Cold start-up takes about two days to reach steady-state operation while an already heated plant takes only 30 minutes to stabilize.

C. HEAT TRANSFER

- The expected high heat transfer rates in the evaporator effects were realized except in the carbon steel tubed effects where corrosion product build-up on the tube surface acted as a large thermal resistance.

- Temperature was found to be the single most important variable affecting the rates of heat transfer in the evaporator effects. The higher values of temperature gave, almost linearly, higher values of the overall coefficient of heat transfer U. For a temperature range of 100°F to 270°F, these coefficients were observed to vary from about 300 to 800 Btu/hr ft^2 °F.

- For optimum heat transfer rates in the effects, small diameter but long tubes are indicated for high temperature effects while short but wide tubes would reduce excessive pressure drop in the low temperature effects.

- Liquid-to-liquid transfer rates in Series 200 heat exchangers were so low ($U = 130$ to 220 Btu/hr ft^2 °F) that the heat recovery from the product water was best done by flashing it and using the flashed vapor in Series 300 condensing vapor preheaters. The overall coefficients in the latter preheaters were in the range of 150 to 550 Btu/hr ft^2 °F.

- Performance of both the heat rejection condensers HX-312 and -318 was below expectations. There was considerable subcooling of the condensate and steam approach to cooling brine was quite high. The overall coefficient averaged between 300 to 400 Btu/hr ft^2 °F. It is important to note here that the plant capacity and economy depend on proper operation of this unit.

- Maximum steam economy is achieved by venting the non-condensibles out at the lowest possible temperature. It is better to prevent the inleakage of air into the subatmospheric units (effects and heat exchangers) by proper design and sealing than to remove the air by higher venting rates, etc. A small amount of air can drastically reduce the heat fluxes and hence the plant capacity.

- Heat losses were very high from the evaporator effects and other equipment. In FY-66, the rate of heat loss was so high that an equivalent 100,000 lb/hr product water loss was incurred indirectly.

- The double-fluted and spirally enhanced surfaces of the heating element tubes in effects and preheater HX-306x showed more than double the heat fluxes achieved with smooth tubes. These enhanced surfaces had U values ranging from 900 (in HX-306x) to 1400 Btu/hr ft^2 ^oF (in Effect XIII).

D. SCALE DEPOSITION AND CONTROL

- The alkaline scales, due to bicarbonates of calcium and magnesium in the seawater feed, can be avoided by acidifying the seawater feed to a pH of 4 to 4.5. The acid addition rate is best controlled, however, in relation to the pH of the feed as it leaves the deaerator-decarbonator, which should be in the range 5.5 to 6.0.

- The acidic scales, due to calcium sulfate salts, can be avoided by designing the process such that the solubility limits of these scale-forming salts are not exceeded. However, the calcium sulfate anhydrite solubility limits may be slightly exceeded if the temperature does not exceed 268^oF and the contact times are short (a few seconds).

- Gypsum ($CaSO_4$, $2H_2O$) scales at high concentration factors and low temperatures can be prevented by addition of polyphosphate compounds (e.g., Nalco-918) at temperatures less than about 170^oF and at a concentration of 3 to 4 parts per million.

- With these scale control techniques, the scaling thresholds are set at 268^oF and a brine concentration factor (final) of 3.2.

- Even distribution of brine to the heating element tubes in evaporator effects is essential since a starved tube will concentrate the brine to a much higher concentration than the other tubes, resulting in scale deposition.

- If scale deposition does occur, the gypsum scales can be washed out of the low temperature effects by immediately increasing the seawater feed to lower the final brine concentration factor to the 2.6 to 2.8 range while increasing the polyphosphate addition rate to about 6 parts per million. Cold seawater can be used for off-stream scale removal.

- Anhydrite scale deposit (usually in high temperature effects) is hard to dissolve in hot or cold seawater. High pressure water jets, drilling, or some similar mechanical means has to be used for removing this type of scale on tube surfaces.

E. MATERIALS OF CONSTRUCTION AND CORROSION

- Carbon steel is a satisfactory material of construction for evaporator cones, vapor bodies, steam chests, heat exchanger shells, brine pump casings, deaerator shells, flash tanks, steam and condensate pipe lines, and chemical and product water tanks. This material serves satisfactorily in deaerated water circuits up to 200°F. At higher temperatures a protective coating is required. Baked-on phenolic coating on carbon steel is satisfactory up to 250°F.

  In high velocity-high turbulence areas such as waterboxes, tubesheets, heating element tubes, elbows, pump suction and discharge piping, carbon steel service life is very limited.

- Cupro-nickel (70-30, 90-10) cladding on steel waterboxes, tubesheets and pass partitions performed excellently even in the non-deaerated feed water circuit. Erosion resistance of this alloy exceeds that of carbon steel or aluminum brass. An economic evaluation, for some applications, would be required to determine the relative merits of cupro-nickel alloys versus carbon steel coated with baked-on phenolic coating.

- Aluminum brass, as a tube material in heat transfer service, has performed excellently when the fluid velocities were less than 7 feet/second. The material can withstand shock chlorination and inhibited hydrochloric acid cleaning action. Its heat transfer capabilities are superior to those of cupro-nickel but inferior to those of admiralty brass. However, admiralty brass has shown less resistance to corrosion than that of aluminum brass.

- Series 300 stainless steels show pitting and stress corrosion cracking in chloride environments at temperatures higher than 122°F. Pump shafts and impellers can be made of this material when the flow of seawater is continuous, and during outages provision is made for flushing with fresh water.

- Ni-Resist iron has exhibited good corrosion resistance when used for non-rotating castings in vertical turbine pumps. However, it corrodes in stagnant seawater. Also, magnetic forms of this material are subject to graphitization. It has somewhat lower resistance to cavitation damage as compared to stainless steel, but is superior to either carbon steel or cast iron.

- Monel, Hastelloy C and Titanium showed excellent corrosion resistance as brine pump shaft material, as lining for heat exchanger waterboxes, and as tubes.

- A considerable scope exists for the use of non-metallic pipings and coatings. Dow-Smith "Chemline" pipe comprised of continuous glass fibers embedded in thermoset epoxy resin withstood temperatures to 265°F, and is recommended for temperatures to 300°F.

  Test spools of epoxy-lined asbestos cement, PVC, ABS, fiberglass reinforced epoxy and polyester in non-deaerated feed up to 150°F showed no failures in two years' operation. Saran-lined pipe was satisfactory for transporting acidified seawater feed to the deaerator. Bis-phenol-type FRP plastic "Corite" as a distributor plate in the evaporator waterboxes was not successful since it cracked frequently in handling.

  Napko latex-base paint was acceptable as protective coating for the exposed plant exterior surfaces. Pinhole-free coating of phenolic, shop-applied by spray techniques at 400°F to carbon steel brine piping and other interior surfaces, gave satisfactory service. A hot-air-dried, catalyzed epoxy coating with a heat reflective layer provided good protection to carbon steel waterboxes at temperatures to 240°F. The coatings must be holiday-free. Removal of silt from seawater feed assists in prolonging the life of these plastic coatings.

- The cathodic protection system was not satisfactory. The magnesium sacrificial anodes were consumed at an excessive rate when used for non-deaerated waterbox protection. The impressed-current method afforded protection but only at a high current consumption of up to 1 ampere/ft^2.

F. MAINTENANCE AND EQUIPMENT PERFORMANCE

- A total of 86 problem areas were identified during the eight years of the Freeport Plant operation. Seven of these problem areas accounted for 79% of the maintenance effort. These areas were: pumps, evaporators (included scaling), heat exchangers, corrosion, air and hydro-static tests, pipes and fittings, and leaks. Repair and replacement of seawater and brine piping was the greatest single maintenance item in the plant. Corrosion and leaks accounted for considerable maintenance effort. Cavitation damage to pumps required frequent maintenance. In general, it can be said that maintenance problems at the desalting plant were not unusual or unexpected.

- Since plant instruments are the process performance indicators, recorders and controllers, their frequent clean-up and calibration is essential. Also, the instrument air supply must be clean (oil- and dirt-free) and dry. Capacity of the air-compressor must be adequate to meet the demand of all air-operated devices. At Freeport, air supply was inadequate and frequently the machine-shop air-compressor had to be used for instrument air supply.

G. PLANT OPERATIONS

- A "Standard Operating Procedures" manual was submitted by Stearns-Roger Corporation, the plant operators, in January 1969. Certain parts of this manual were revised in April 1969. Instructions and procedures are given in this manual on aspects of plant start-up, operation and shutdown, except for emergency procedures.

- The number of personnel on the plant crew varied from a low of 13 in FY-1964 to a maximum of 20 in FYs-1968 and 1969.

- By automation of plant control activities, a 50% reduction in plant operators (from a total of 8 to 4) was achieved in FY-1964. Maintenance personnel were trained to act as relief operators.

- A fairly comprehensive product water corrosivity testing and stabilization program was carried out. The results indicated that untreated product water at Freeport was very corrosive but it could be stabilized by treatment with either polyphosphate or sodium silicate.

H. OVERALL PERFORMANCE

- The Freeport Saline Water Conversion Plant demonstrated the economic and technical feasibility of long-term operation of the multiple-effect falling-film LTV distillation process for production of potable water from seawater.

- In general, the design objectives of both the original 12-effect and later the 17-effect plant were met with regard to the technical expectations. For the 12-effect plant, the rated capacity was exceeded by 20% while using 2/3 of the evaporator surface and an extraction ratio of only 0.67, instead of the design value of 0.75. This capacity was not always maintained because of the developmental nature of work that was being performed. The 17-effect plant also operated close to its rated capacity of 1 million gallons of fresh water per day.

- The average gross performance ratio (pounds of water distilled per pound of prime steam) for the 12-effect plant was 10.97 as compared to the design value of 11.5. The 17-effect plant never achieved the gross design performance ratio of 14.8. The actual values ranged from 11.85 to 14.2. Excessive heat losses from the evaporators and exchangers were stated to be partially responsible for this low performance.

- Electrical power demand for the 17-effect plant was 525 kilowatts compared to the design values of 450 kilowatts, an increase of approximately 16.7%. The 12-effect plant also exceeded the design value of the power demand (360 kilowatts), but to a smaller extent (11.2%) — 400 kilowatts was the actual average consumption.

- Defining the operating period as the period of time from the beginning of a development run to the time the plant was shut down following the completion of test objectives, the average on-stream factor was found to be 90% covering the 16-development-run period (February 1964 to May 1969). Since maintenance performed during the plant shutdown period is not included in this calculation, the figure is optimistic.

I. ECONOMIC ASPECTS

Since the production of water at a reasonable cost was stated to be one of the main objectives of the Freeport Plant operations, one would expect a careful and consistent economic analysis of water costs from year to year. Unfortunately, the reports submitted by Stearns-Roger Corporation to OSW do not quite come up to this expectation. Granted, this analysis would have involved some effort but it would have contributed a considerable degree of certainty in projecting Freeport-type desalting plant capital and water costs. The following is a brief list of conclusions drawn from the reported data, which was scanty and sometimes contradictory.

- There were at least four different figures reported as the original cost of the Freeport Plant. These ranged from $1,265,387.50 to $1,569,000.00. Some of this variation arose, probably, due to changes in equipment such as removal of the clarifier-thickener and six Series 200 heat exchangers. If one desires the total investment at the time of initial start-up, the cost of construction plus the architect-engineering services, and land and start-up costs, then the figure might well be $1,569,000.00.

- Very little information was available regarding the cost of design modifications. Annual Report Number 6 by Stearns-Roger listed $838,102 for maintenance and modification — the bulk of this going into the installation of the 5-effect module and the Auxiliary Test Unit in fiscal year 1967.

- Cost of "routine" maintenance varied over the eight-year period from a low of $48,911 in fiscal year 1964 to a maximum of $151,069 in fiscal year 1968. Cost of the modifications to the original 12-effect plant to convert it to 17-effect configuration can probably account for some of the extraordinary maintenance cost in FY-1968.

  The principal areas of maintenance were observed to be the pumps, the evaporators, general plant maintenance, and the piping. Because of the developmental nature of operations at Freeport, the maintenance costs were stated to be higher than would be encountered in a normal continuous production plant.

- Only one item was clearly available as unscheduled maintenance cost — the cost of repairs due to the damage caused by the Hurricane Carla. This was stated to be $38,535.03 in the FY-1961 report.

- Desirable as it is to know the actual operating costs, these are quite entagled in a developmental plant such as the Freeport facility. The apportionment of a given expense between normal production cost and research and development costs is difficult. Viewed in this light, it is not too surprising to find the cost per 1000 gallons as varying from $0.76/1000 gallons to $4.04/1000 gallons during fiscal years 1962 through 1968. The plant production was low during R&D operations or none during modification to 17-effect configuration so that less inaccurate water cost figures are the "normalized" figures arrived at by adjusting developmental and other expenses, etc., as given in the OSW R&D Report Number 72.

- The theoretical or normalized water cost (based, among other things, on 330 operating days per year) varied during the FY-1963 through FY-1969 period from $1.17/1000 gallons to $1.51/1000 gallons. These water costs, either normalized or actual, do not come anywhere close to the projected cost of $0.90/1000 gallons projected by Stearns-Roger Corporation for the 17-effect plant.

- The cost centers contributing the most to water costs were: utilities (steam and electricity); G&A expenses (indirect operating expenses); labor (operating and maintenance); materials and supplies (maintenance); and depreciation on capital investment.

- The normalized capital cost per gallon of daily capacity was estimated as varying from $1.424 to $1.827 over the FY-1962 to FY-1969 period.

J. RECOMMENDATIONS

- The feasibility and smooth operability of the multiple-effect falling-film distillation process has been proven beyond any doubt. What is now required is further work that will lead to a decrease in unit water costs.

- Efforts should be directed towards increasing the capacity of the plant with fixed equipment sizes. The present bottleneck to increase in capacity is the high temperature limit of about 268°F (first effect brine temperature) imposed due to scaling from calcium salts. Both fundamental and applied research must continue in this area, particularly since scaling is a widespread problem. Three approaches to this problem are: the reduction or elimination of either the calcium ion or calcium sulfate; developing scale-loosening compounds; and developing operational (in situ) methods of scale prevention and/or removal.

- Cost reduction study is needed in the two categories contributing the most to water cost — the fixed charges and the energy costs. Some possibilities in this direction are:

  - Reduction in capital costs by use of modular configuration with effects stacked to reduce land area as well as to eliminate pumps and minimize power requirements. Use of cement-concrete shells and pipes is worth investigating.

  - Reduction in energy costs by utilizing off-peak energy or new, cheaper sources of energy — e.g., combination of city refuse burning plants and water production plants.

- Continued studies are required for further improvements in heat transfer coefficients. Double-fluted and spirally indented surfaces showed at Freeport that these coefficients can be at least doubled. Use of additives to induce dropwise condensation of vapors and thus to augment heat flux rates needs further research and testing.

- Finally, since all that remains of a demonstration plant after it is closed down is the written and reported material, this ought to contain maximum useful data pertaining to every stage of plant development for analysis and further improvement and organized in such a way as to be readily available. It is proposed that a format be developed for such reports and for every section of the reports. Throughout this investigation, the incompleteness of reporting of both the technical and economic subjects by the operating contractor was apparent and affected adversely this summarizing effort. The cost normalization procedure developed in 1963 also requires some modifications and updating.

U. S. GOVERNMENT PRINTING OFFICE : 1972 O - 454 997