THIRD ANNUAL REPORT

Saline Water Conversion Demonstration Plant No. 1

FREEPORT, TEXAS

United States Department of the Interior

Office of Saline Water • Research and Development Progress Report No. 123

SALINE WATER CONVERSION PROGRESS REPORT NO. 123 • SEPTEMBER 1964

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FREEPORT, TEXAS

By Stearns-Roger Manufacturing Co., Denver Colo., covering the period, July 1, 1963, through June 30, 1964, for Office of Saline Water, Charles F. MacGowan, director; Raymond H. Jebens, chief, Demonstration Plants Division; Harold D. Singleton, field engineer.

UNITED STATES DEPARTMENT OF THE INTERIOR • Stewart L. Udall, Secretary

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FOREWORD

This is another of a series of reports designed to present accounts of progress on saline water conversion with the expectation that the exchange of such data will contribute to the long range development of economical processes applicable to large scale, low cost demineralization plants of multi-million gallon per day capacity for conversion of sea and other saline waters.

Except for minor editing, the data herein are as contained in a report submitted by Stearns-Roger Corporation under Contract No. 14-01-001-218, which has been submitted as the third annual report on the operation and management of the Freeport, Texas, 1,000,000 gallon per day demonstration sea water conversion plant. Neither the Department of the Interior nor any person acting on behalf of the Department, makes any warranty or presentation with respect to the accuracy, completeness, or usefulness of the information contained in this report, or that the use of any information or process equipment disclosed in this report may not infringe privately owned rights.

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I. INTRODUCTION

1-1 AUTHORIZATION.

1-2 On September 2, 1958, Public Law 85-883 authorizing the Demonstration Plant Program was enacted. The law provided for construction of not less than five desalination plants, each to demonstrate a different water conversion process in actual production tests. Upon completion of site selection and other related studies, a plant designed for 12-effect, Long-Tube, Vertical (LTV) evaporator method was selected to be constructed at Freeport, Texas. The Plant was designated as Demonstration Plant No. 1 by the Office of Saline Water.

1-3 The Stearns-Roger Corporation has operated the Plant since its completion in April, 1961 under Management and Operations Contract No. 14-01-001-218.

1-4 HISTORY.

1-5 The Plant site and process to be demonstrated were selected on March 2, 1959. The design contract was awarded to W. L. Badger Associates and the construction contract to Chicago Bridge and Iron Company. Stearns-Roger Corporation was awarded the Operation and Maintenance contract just prior to completion of the construction contract. During the time period from the inception of the Stearns-Roger contract to Plant construction completion, Stearns-Roger personnel observed performance testing and modifications.

1-6 Beginning with the initial operation in April, 1961, the Plant has demonstrated that the LTV process is economically feasible and possesses a high degree of reliability. Original design intended that the Freeport facility was to be the most thermally efficient, full scale desalination plant in the free world. In the main, this efficiency has been attained, although maintenance of high thermal performance has necessitated some changes from the original process design. The Plant's thermal performance has been approximated by the 36-stage, Multiple-Stage, Flash (MSF) plant at San Diego under short term conditions, and is equalled by a 40-stage, MSF, 500,000 gallon-per-day unit on the Isle of Guernsey[1].

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[1] Dodge, Barnett F., Advances in Chemistry, Series 38, "Review of Distillation Processes for the Recovery of Fresh Water from Saline Waters", American Chemical Society, Washington DC, 1963

1-7 Operations throughout the first year were continually hampered by failures of various experimental features. The second year was largely occupied with correction of deficiencies and failures, but during each year, production exceeded 250,000,000 gallons. Daily production maximums of 1,200,000 gallons were attained. Costs per 1,000 gallons of water have been steadily reduced due to increased reliability, which resulted in reduced operating, labor, and maintenance costs.

1-8 Production was emphasized for approximately the last three months of fiscal 1963 and the first six months of fiscal 1964. The time was devoted to long production runs, rather than development runs. During the 12-month period beginning with April, 1963 and ending with March, 1964, 308,507,100 gallons were produced.

1-9 In January, 1964 it was determined that further production improvements were dependent on process changes, and a schedule of short development-type runs was initiated. Despite the resultant temporary reduced production and increased water costs, these runs have proved most beneficial in providing significant process improvements.

1-10 ABSTRACT.

1-11 Information is presented concerning LTV multiple effect evaporation desalination plants. The technical, logistical, and economical evaluations for the fiscal year 1964 operations are presented, and also the process and mechanical development program results as related to the LTV process in particular and desalination in general. These data are the results of operation, maintenance, and development studies conducted at U. S. Department of the Interior, Office of Saline Water, Demonstration Plant No. 1 located at Freeport, Texas.

1-12 Actual capital and operating costs are presented and compared to theoretical normalized capital and operating costs. Production, economy ratios, progressive cost averages, and maintenance problems are presented. A thorough technical evaluation of the existing process, equipment, corrosion, and materials of construction is included. The process evaluation is extended to include limitations to the temperatures and concentration ratios for scale-free operation of LTV evaporators without pre-softening the feed water. An evaluation of each category of process equipment was made, and is included. Complete information concerning performance of each subsystem is included, with particular emphasis on the deaerator-decarbonator.

1-13 A proposal section is provided to present the modifications recommended, the continuation of the development program, and extraordinary maintenance contemplated for the fiscal year 1965.

II. SUMMARY

2-1 PRODUCTION SUMMARY.

2-2 Total production for fiscal 1964 was 289,807,260 gallons. This figure could have been considerably greater except for numerous and sometimes extended outages that were necessitated during the last half of the year to comply with the development program schedule. Figure 2-1, Daily Water Production, illustrates the ability of this Plant to produce at rated capacity immediately after startup, and to continue for extended periods of time without attenuation of capacity. Table 4-5 (located in Section IV of this Annual Report) tabulates the important production figures for the year. The reduced productivity in June, 1964 was intentional in that the Plant was being operated at a final concertration factor of only 2.5. It is interesting to note that at the re-rated Plant capacity of 1,060,000 gallons-per-day, the production loss was held to 15 percent, and this loss was due to the lack of pumping capacity.

2-3 As of 30 June 1964, this Plant has produced a total of 796,012,785 gallons of water. The limiting production factor for this Plant is its pumping capacity rather than its heat surface. Notable production months are: August, 1963 (31,650,830 gallons at $1.06 per 1000 gallon) ; December, 1963 (31,638,850 gallons at $1.09 per 1000 gallon); and May, 1964 (32,331,500 gallons at $1.13 per 1000 gallon). At the other extreme is October, 1963 (14,953,740 gallons at $2.28 per 1000 gallon) which had extensive downtime necessitated by maintenance and Plant modification. The same reasons hold true for June, 1964 (17,539,500 gallons at $1.91 per 1000 gallon), but with the added restriction imposed on this monthly production by the development program schedule. The months cited above illustrate the strong influence that downtime has on water production costs.

2-4 The Plant was on-stream for 77.3 percent of the time which is equivalent to 282 days of the year. An evaluation of Figure 2-1, Daily Water Production and Table 4-5, Monthly Production and Cost Summary reflects the frequency and extent of downtime encountered during the fiscal year. Much of this downtime was required by the development program. The effect on cost of eliminating excessive downtime can be appraised at just over $1.00 per 1000 gallon of water produced. The development work now in pro-

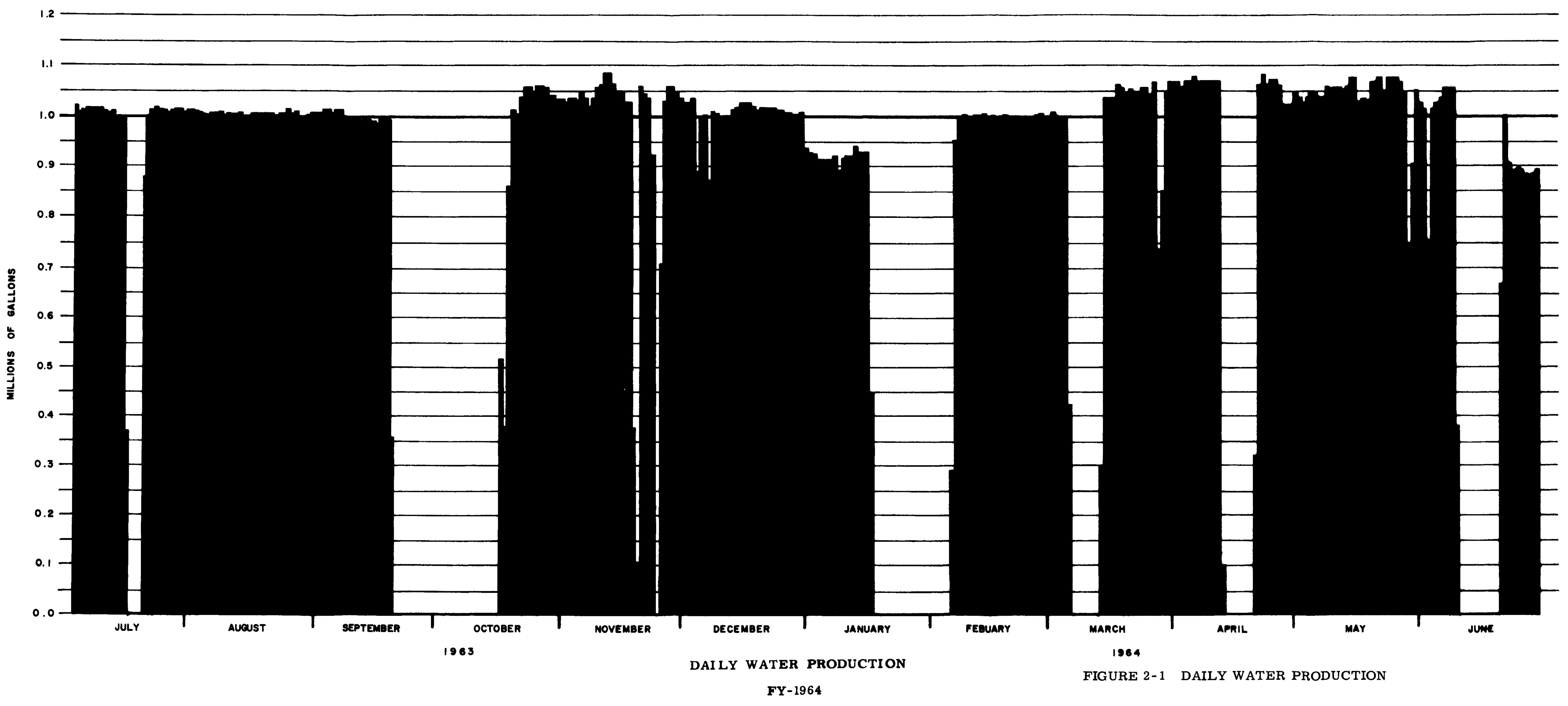


FIGURE 2-1 DAILY WATER PRODUCTION

gress is expected to increase the Freeport Plant productivity, decrease downtime, and result in water costs below $1.00 per 1000 gallon of water produced. Results of development work to date strongly indicate that the productivity can be increased 25 percent with minor improvements to the pumping capabilities.

2-5 MAINTENANCE SUMMARY.

2-6 This maintenance summary not only includes a summation of normal maintenance, but also summarizes the modifications and/or improvements that were performed in conjuction with the maintenance tasks on various items of Plant equipment. The over-all function of the equipment discussed below can be determined by referring to Figure 7-1, Process Flow Diagram, located in the Appendix of this Annual Report.

2-7 At various times of the year, "maximum" impellers were installed in pumps throughout the Plant. At the present time, all brine pumps are equipped with maximum impellers (with respect to drive motor capacity). Other pump maintenance tasks were confined to the replacement of bearings and packings, with the exceptions of pumps P-2, P-4 and P-5. Pump P-2 was replaced with pump P-2a. Pump P-2a has some balance problem on startup. Pump P-4 provided limited service until pump P-5a was removed for repairs in May. Through May and June, pump P-4 required frequent and expensive maintenance due to its low available net positive suction head. Pump P-5a (the vertical turbine canned pump) was removed from service in May due to excessive vibration. This pump had given excellent maintenance-free service for almost five months. During removal, manufacturing defects were discovered and corrections of these defects and other modifications to the pump kept it out of service for the balance of the year. Nevertheless, installation of pump P-5a made very significant improvement in over-all pump maintenance.

2-8 Evaporator maintenance and modification tasks included: replacement and plugging of leaking carbon carbon steel plugs; installation of suction connection to recycle pump on XII effect; and insertion of weirs for better brine distribution. Of all evaporator maintenance performed, the temporary plugging, repair and replacement of carbon steel plugs was the most expensive and time consuming. The recommended replacement of these carbon steel plugs with aluminum brass plugs would be very beneficial in minimizing future maintenance costs.

2-9 Heat exchanger tasks included the complete removal of heat exchangers 202 through 207. In addition, the high temperature 300-series heat exchangers failed on several occasions due to pitting-type corrosion of carbon steel. Two tubes were replaced in heat exchanger 311. Experimental cleaning of heat exchangers with high-pressure (3000-4000 PSI) water jets was attempted, but determined as insufficiently effective to justify its cost.

2-10 An accelerated protective coating program was initiated in February, 1964. The initial protective coat of aluminum paint applied over dirty, corroded surfaces was a failure. This year's effort was the first properly selected and applied protective coating system for the Freeport Plant, and cannot be considered as normal maintenance.

2-11 Other maintenance items of interest are summarized below:

a. Rework of vacuum pump foundation

b. Installation of chemical injection system to stabilize product water

c. Installation of product filter

d. Cleaning of product tank and intake pit;

e. Installation of carbon steel piping that is internally coated with with baked-on phenolic material;

f. Rehabilitation of clarifier-thickener system for silt settling service;

g. Removal of raschig rings from deaerator; and

h. Increasing maintenance problems with control-gear drive connection of level controlling butterfly valve.

2-12 ECONOMIC SUMMARY.

2-13 Great effort has been expended by all concerned to "normalize" costs for comparison purposes. Preliminary attempts to "normalize" the demonstration plants for fiscal 1964 were based on a 350-operating-day year. At the direction of the Office of Saline Water, this Report "normalized" the Plant operation on the basis of a

330-operating-day year. This change results in an approximate five percent increase of "normalized" cost per 1000 gallons over the cost for a plant "normalized" on a 350-operating-day year basis. When making comparisons to other demonstration plant results, the reader is cautioned to first ascertain that the normalizations were all derived from the same basis. Capital costs have also been "normalized" in strict adherence with OSW procedures.

2-14 Section IV, Economic Evaluation, details the actual and theoretical, capital investment and production costs for the "normalized" Plant. Table 2-1 lists these costs for comparison.

	CAPITAL COSTS		OPERATING COSTS		
	TOTAL IN DOLLARS	DOLLARS/GAL DAILY CAPAC	DIRECT CENTS/M GAL	INDIRECT CENTS/M GAL	TOTAL DOLLARS/M GAL
ACTUAL	1,447,362	1.433	64.86	58.04	1.23
THEORETICAL	1,845,080	1.827	77.29	60.63	1.38

TABLE 2-1. COMPARISON OF ACTUAL AND THEORETICAL COSTS OF "NORMAL" PLANT.

2-15 Economic data reporting of a typical month is also included in the Economic Evaluation Section. The month of May, 1964 was selected because it best represents the present operation. May had an unscheduled outage and normal maintenance, with $0.6756 direct costs, $0.4521 indirect costs, for a total production cost of $1.13 per 1000 gallons of water produced. This total cost included the contractor fee and overhead percentage, whereas, the "normalized" costs do not. If the contractor fee and overhead percentage are deducted, the total production cost is decreased to $1.04 per 1000 gallons of water produced. Methods of operation developed during the fiscal year have tentatively proven that the Plant is capable of repeating the performance of May, 1964 over a sustained period of time. Further increases in productivity with only minor increases in indirect costs are strongly indicated for fiscal 1965. It is, therefore, predictable that the Freeport Plant will achieve costs below $1.00 per 1000 gallons of water produced.

2-16 Section VIII, Appendix, contains the ledger summary for the month of June, 1964 which is the source of capital value and cost figures. The ledger summary also presents a detailed accounting of total costs. June, 1964 is selected because the fiscal year (FY)-to-date figures are the yearly totals. The over-all cost figures for the year are $395,038.91, as compared to $421,236.96 for fiscal 1963. Since fiscal 1964 production was up and cost down, the over-all cost per 1000 gallon of water produced was $1.36, a reduction of $0.32 over fiscal 1963.

2-17 TECHNICAL SUMMARY.

2-18 The process flow diagram (figure 7-1) was revised to indicate the flow as it existed at the conclusion of the fiscal year 1964. The diagram is placed in Section VII rather than in this Section as an aid to the reader, inasmuch as this figure is also referenced in other Sections of this Annual Report. This diagram illustrates the corrected flows around effects X, XI and XII; the deaerator; the intake; and the 200-series preheating exchangers. In addition to these corrections, the steam pressure reduction and desuperheating flow, the barometric condenser, and the product water filter have been added to this diagram.

2-19 A new heat and material balance diagram is included in Section V, as Figure 5-1. The heat and material balance diagram is based on actual average operating conditions achieved with a final concentration factor of 3.0, at which condition, scale-free operation was achieved. Additionally, the diagram effectively illustrates the utilization of heat transfer surface, the flows which must be accomodated, and the restricting effect of the limited tube surface in effects IV and X. It also forms the basis for an evaluation of pumping capability and performance of preheating exchangers.

2-20 A new development program was initated in January, 1964 embracing all phases of Plant and process performance. Emphasis for the first four development runs was concentrated in the following:

a. Deaeration and Decarbonation Processes;

b. Concentration Ratio (Extraction Ratio);

c. Brine Distribution in the Evaporators;

d. Scale Formation Control;

e. Evaporator Performance and Design Correlations;

f. Intake Performance;

g. Materials of Construction; and

h. Product Water Stabilization.

2-21 Two development reports were issued which indicated notable progress in brine distribution through the use of notched weirs, and in scale prevention through the use of polyphosphates. Scale-free operation at full capacity and 3.0 final concentration factor was achieved for the 30-day period of Development Run No. 2. Significant progress was made in correlating heat transfer coefficients to the dimensionless parameters characterizing the flow of brine. Studies of scale formation and removal rates indicates on-stream scale removal is not only possible but practical, in that it can be accomplished with very nominal reduction in productivity.

2-22 Deaerator development work conducted resulted in increasing ability to determine the actual results in the form of better pressure, temperature, and flow measurements; and an ability to analyze the vapor streams quantitatively. The deaerator had the capability to remove oxygen to less than 100 PPB consistently, with readings as low as 30 PPB, which are perhaps beyond the accuracy of analytical procedures. Residual CO_2 was measured as bicarbonate alkalinity and reported as equivalent PPM $CaCO_3$. This value has consistently remained less than 10 PPM, and usually less than 8 PPM. No scale has formed in the effect I evaporator.

2-23 Results were submitted to the W.M. Kellogg Company for use in their CO_2 solubility and rate of release studies. Analyses were obtained of the steam from effect XI which indicated negligible CO_2 content and a complete absence of air. Analyses of the vapors from the deaerator indicated a maximum of 0.36 percent CO_2 by volume, and 0.13 percent air by volume. Steam flows were reduced with only a slight rise in residual carbonate, and pH of acidulation was raised with similar results. The net effects were a decrease of both acid and caustic consumption, a decrease in heating of the feed water in the deaerator, and a decrease in steam requirements.

2-24 Deaerator pressure readings were approximately at the proper levels, but the packing was not indicated to be effective. The packing was removed with a resultant net effect of carbonate residuals of 12 PPM maximum, using only the spray nozzle. New plastic packing of different characteristics was ordered, but had not been received at the end of the fiscal year. Better methods of determining residual CO_2 are needed as well as over-all instrumentation for obtaining information. Partial and total pressures have been determined, but lack of basic sea water data hampers the development of design correlations for future deaerators. The Freeport unit obviously has overcapacity with the use of an efficient packing.

2-25 Evaporator performance was studied in effects I, VI, X, XI and XII, for each development run. Heat transfer coeffieients were found to be comparable to those developed at Wrightsville Beach when calculated on the same basis. The coeffieients ranged from a maximum of 731 BTU/HR SQ-FT oF in effect XI to a minimum of 313 BTU/HR SQ-FT oF in effect X. This important design factor is apparently strongly influenced by flow rate as can be determined from the above, in which effect XI had recycle and a Reynolds number of 6866, while effect X had no recycle and a Reynolds number of 2892. Heat flux in all cases was less than the level at which nucleate boiling would occur.

2-26 During the scale-free run (Development Run No. 2), the load of evaporation was observed to shift from effect XII to effect XI as scale was dissolved out of effect XI during the run. This illustrates the non-utility of much surface in effect XII. Carryover was not a problem in any effect at a final concentration factor of 3.0. The temperature differences across each effect varied from as high as $15^{\circ}F$ in effect X to as low as $5.5^{\circ}F$ in effect VI. The "U" values are better at lower Delta T. It is a reasonable conclusion that successful operation can be achieved with a design temperature drop across each effect of $7^{\circ}F$ or less.

2-27 Determination of the available temperature difference for evaporator heat transfer is dependent on the ability to measure temperature and pressure conditions in the top water box. It is certain that much flashing occurs in this section, but the degree of approach to equilibrium is not known. There are indications that some vaporization may take place after the brine leaves the tube. Attempts have been made to measure the appropriate pressures and temperatures, but more instrumentation to obtain adequate measurements is recommended.

2-28 The recirculation in effects XI and XII is beneficial, but the method of brine transfer from effect X to XI and effect XI to XII is considerably less than optimum. Extensive revision in this area is recommended to improve distribution, better utilize the heat transfer surface, and lessen carryover problems resulting from rather violent flashing in the cone of effect XII.

2-29 Over-all Plant performance has been carefully evaluated with the establishment of Plant Operating Paramenters. Operating data have been reduced on a daily basis for determination of many of these parameters. A copy of the Development Run operator data sheet is included in the Appendix (Section VII) to illustrate the depth of analysis of each day's operation for development purposes. The most interesting parameters and the typical values for each are:

Steam Economy Ratio Over-All (net)	10.83 LBS H_2O/ LB Steam
Steam Economy Ratio Corrected	10.12 LBS H_2O/LB Steam
Average Production Rate (gross)	1,060,000 GPD
Average Total Plant ΔT	$126^{\circ}F$
Maximum Steam Demand	32,700 LBS/HR
Extraction Ratio	0.67 LB H_2O/LB SWF

2-30 The requirements of the development program have prevented a run utilizing a complete combination of known optimum conditions. The following is a summary of optimized date presented in Section IV, using parameters identical to those above.

Steam Economy Ratio Over-all (net)	10.94 LBS H_2O/LB Steam
Steam Economy Ratio Corrected	10.56 LBS H_2O/LB Steam
Production Rate, (gross)	1,103,980 GPD
Total Plant ΔT	$131^{\circ}F$

Steam Demand	32,000 LBS/HR
Extraction Ratio	0.70 LB H_2O/LB SWF

For these operating conditions, the Freeport Plant can operate continuously, scale-free with the use of notched weirs and polyphosphate addition at effect VIII and IX.

2-31 The individual evaporator performances are indicated on the heat and material balance diagram (Figure 5-1). The restrictive effect of effect IV evaporator with only 4,500 SQ-FT of heat transfer surface, and of effect X evaporator with only 5,000 SQ-FT is illustrated by the heat and material balance diagram. The addition of 33 percent surface in effect XI and 42 percent in effect XII cannot be properly utilized because of the limited surface in effect X. The situation in effect XII is further degraded by the withdrawl of steam for the deaerator. Adjustments in the surface available in the evaporators is recommended to balance the load and prevent these restrictions.

2-32 The heat exchangers and flash tanks are only partially utilizied at present. Heat exchangers 208, 209, 210, 211 and 212 could possibly be bypassed, but condensate pumping capabilities would be insufficient under those conditions. Heat exchanger 201 could readily be eliminated and should be removed from service when significant maintenance is required. Heat exchanger 215 is improperly utilized by branching the flow to the Dow product tank prior to the point at which the condensate enters this exchanger.

2-33 A scheme for quantitative measurement of scale has been devised, and was used to measure scale during each outage. The results were estimates of scale formation rates under various operating conditions. This method also served to establish scale removal rates during periods in which cold sea water is being circulated.

2-34 Pumping requirements and capabilities have been the subjects of extensive study. Sea water feed rates are limited to approximately 500,000 LB/HR by existing pumps. The present limitations are imposed by pumps P-1, P-17, P-18, and the condensate pumps P-38 through P-43. Pump modifications and/or replacements are recommended to achieve higher rates required for planned temperature and production increases.

2-35 Studies being conducted at the end of the fiscal year included silt removal in the clarifier-thickener, intake recirculation and operation, and final condenser performance. The final condenser 312 appears to be considerably below desired performance due to lack of proper steam distribution. It is recommended that steam-side baffle improvements be considered as a minimum measure to correct this deficiency.

2-36 A program is in progress to develop water stabilization treatments applicable to all desalination plants. It is designed to provide corrosivity data which will enable stabilization of the very pure converted water to prevent damage to any normal municipal water distribution system.

2-37 Materials of construction have been evaluated and this program will continue. The materials evaluation resulted in recommendations (for specifications) of carbon steel, cupro-nickel, aluminum brass, Ni-Resist iron, Admiralty brass, and the 300-series stainless steels. It is also recommended that the non-metals be further investigated to substantiate expanded usage of these materials as piping, vessels, and applied coatings.

2-38 Process limitations have been established which require no anti-scale treatments other than the pH method of decarbonation and a 4 PPM polyphosphate injection in the latter effects. The temperature limitation is a first effect steam temperature of 270°F and 100°F for brine overboard. This will allow use of 20 effects with resulting economy ratios in the range of 20 to 1. The final concentration factor should not exceed 3.5, requiring only 1.4 LB of sea water feed for the extration of a pound of water.

2-39 <u>OPERATIONS, PERSONNEL, SAFETY, AND LABOR.</u>

2-40 OPERATIONS.

2-41 At the beginning of fiscal year 1964, the Plant was operating at its rated capacity. A short outage as taken at the end of July, 1963 during which improvements were made to the pumping capability. High-pressure water jet cleaning of heat exchangers was also investigated, as preparations were made for a longer sustained run. At this time, shock chlorination was implemented every tenth operating day, and continued until mid-January, 1964. August, 1963 was a month of continous production and no downtime. This situation continued until September 23, at which time an extensive outage was taken (partially) because of large silt accumulation in the preheat exchangers. Hurricane Carla threatened but did not materialize. The pumping capacity was again increased by the installation of maximum size impellers in additional pumps. Also, drilling of effect XI tubes was necessary at this time.

2-42 Production resumed on October 16 with a revised deaerator level sensing system which proved to be a major improvement. Continuous operation was achieved, except for a short (37-hour) unscheduled outage and a scheduled 24-hour outage, until the January 17, 1964 outage. During the two interim outages, a venting study corrected a condensate siphoning problem, pump P-5a was installed in place of pump P-5, and pump P-2a was installed to replace pump P-2. In addition, freeze protection was successfully applied to the entire Plant. As a result, the Plant economy and efficiency remained very good during a period when the temperatures dropped to 20°F. The application of chlorinated-rubber protective coating was pursued during the entire outage period when weather permitted.

2-43 During the outage which began on January 17, 1964, brine distribution weirs were installed in effects X and XII as well as the performance of other modifications to prepare the Plant for Demonstration Run No. 1. Quantitative measurement of scale was accomplished for the first time, and accurate starting conditions were established for Development Run No. 1. This Run began on February 6 and was completed on March 6. The brine distribution weirs proved to be successful. Operations during the Run were concentrated to obtain good data at constant conditions. Blending strong brine with raw sea water was necessary to obtain a satisfactory inlet concentration factor for comparison purposes.

2-44 During the outage beginning March 7 and ending March 13, preparations were made for Development Run No. 2. These preparations included the setting up of procedures for the revised desuperheater and the polyphosphate injection at pump P-18. Scale was again quantitatively measured and removal rate studies were performed with cold sea water circulation. Development Run No. 2 began on March 13 and continued to April 13. Emphasis was placed on the evaluation of vacuum systems, deaerator vapor streams, and heat exchanger performance. The brine distribution weirs were omitted from effect XI to properly evaluate the effect of polyphosphate addition; this proved very successful as no scale formed during this Run. Addition of stabilization chemicals and the filtration of product water to City of Freeport were initiated at this time. During the shutdown following Development Run No. 2, brine distribution weirs were installed in effect XI.

2-45 Development Run No. 3 was conducted without polyphosphate addition to evaluate the effect of brine distribution weirs in effect XI. It was necessary to attain frequent and accurate data on the evaporators in order to monitor the formation of scale which was shifting from XI effect to XII effect because of the weirs. Assistance was provided to Dow Chemical Company who were making a process study of the Plant for OSW. Development Run No. 3 was completed and discontinued on June 9, immediately after the visit by the Prime Minister of Israel. Interruptions during the Run were limited to problems associated with pumps P-4 and P-5a.

2-46 The shutdown of mid-June, 1964 was extended to complete the protective coating program and the tie-in of the clarifier-thickener. Pumps were readjusted for maximum performance in anticipation the low-final-concentration-factor run. This Run (Development Run No. 4) began on June 22 and is still in progress at fiscal year's end. The Run has established an upper limit of pumping capability of 505,000 LB/HR SWF.

2-47 PERSONNEL.

2-48 Significant changes in management and technical personnel were made during the fiscal year. T.S. Frost was transferred to the Denver organization in September, 1963 and J.H. Born was promoted from Plant Engineer to Plant Superintendent. Robert Bonvillain was relieved of operating duties to become a full-time Plant Chemist. The vacated position of Plant Engineer was not filled until May, 1964 when R.D. Rhinesmith assumed that position. D.D. Kays was assigned as Project Engineer in the Denver office in February, 1964 to provide technical direction to the Plant and assistance to the Project Manager, C.G. Rogers. Figure 2-2 depicts the resulting organization.

2-49 Changes in operating personnel were limited to three resignations and two new employments. A training program was instituted to upgrade the capabilities of existing Plant personnel and to train new employees. In January, 1964, the operating staff was reduced to one man per shift (see Figure 2-2). This change was a result of gradual Plant automation and has proved to be a successful method of operation.

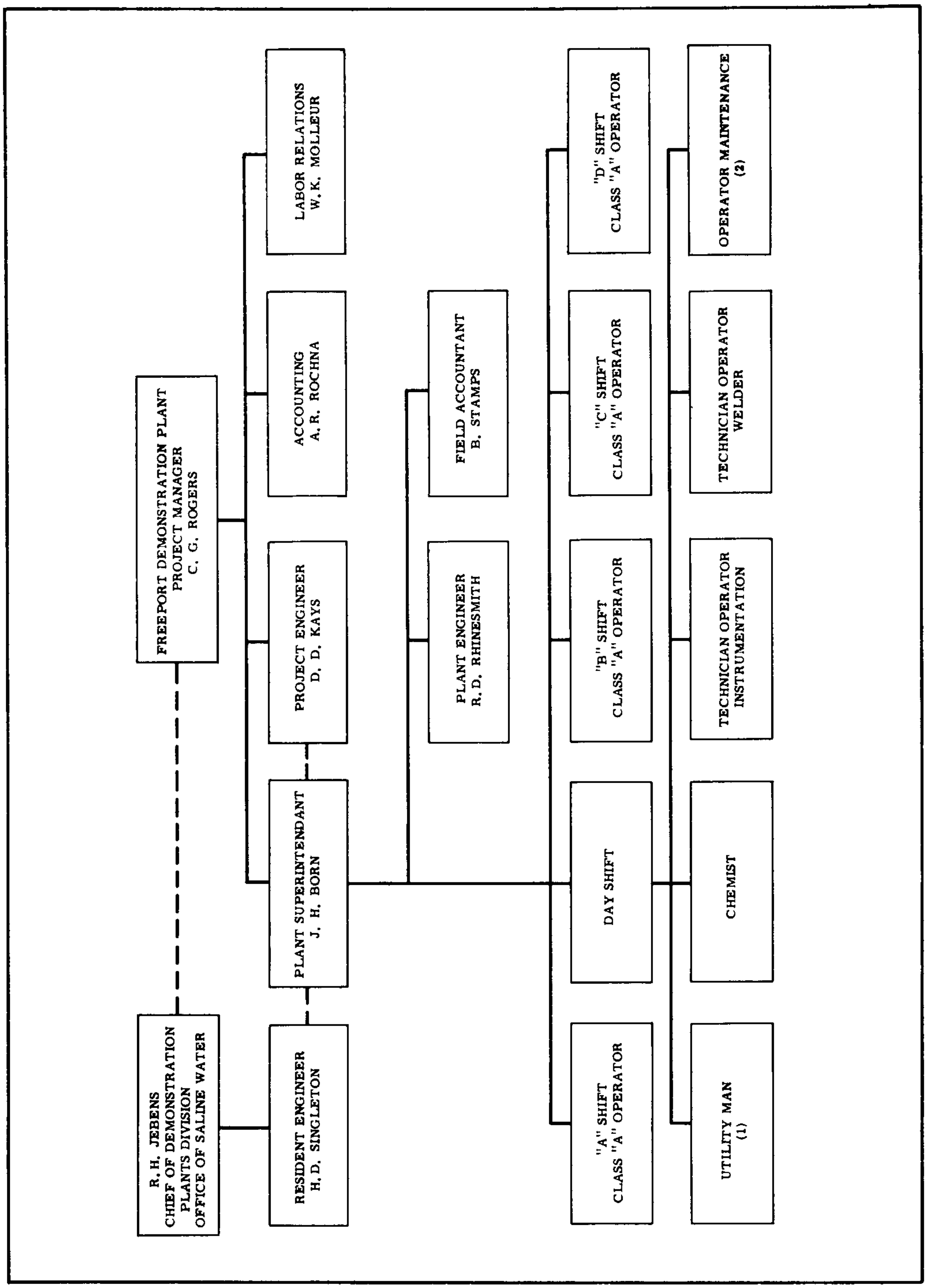


FIGURE 2-2. FREEPORT PLANT ORGANIZATION

2-50 SAFETY.

2-51 Safety meetings were scheduled at least once every month, and were generally held at more frequent intervals. As a result of the Plant's Safety Program, a good safety record was achieved. Nine injury accidents and one lost-time accident of 40 hours were sustained out of a total 68,183 working manhours.

2-52 LABOR.

2-53 In July, 1963, a petition was received for a National Labor Relations Board election to determine whether the hourly-wage employees wanted representation by the International Union of Operating Engineers. This petition was deemed valid and the election was held in September. The results were unanimous for such representation. The first employer-Union contract negotiation meetings were held in October, 1963 and continued (as progress could be made) until March 19, 1964. On this date, a labor contract between Stearns-Roger Corporation and Local No. 564 of the International Union of Operating Engineers was approved.

2-54 <u>CONCLUSIONS AND RECOMMENDATIONS.</u>

2-55 CONCLUSIONS.

2-56 Conclusions relative to the over-all Plant operation can be drawn from the fiscal year's operation as follows:

a. The Plant is capable of producing at design rate under conditions that are significantly adverse from the original design conditions.

b. The Plant is capable of scale-free operation at a final concentration factor of at least 3.0 and probably at or near 3.5.

c. Gradual replacement of inappropriate materials and equipment has improved reliability, and this coupled with scale-free operation, can enable the Plant to remain on-stream in excess of 330 days per year.

d. Water costs will level off at just over $1.00 per 1000 gallons of water produced at present production rates.

e. At the present time, pumping capacity is limiting the productivity of the Freeport Plant and is also inhibiting the efficient use of the available heat transfer surface.

f. The elimination of heat exchangers 202 through 207 has imposed an increased heat exchange load on the corresponding 300-series heat exchangers. These heat exchangers were adequately tubed to carry this increased load, however, the same may not be true for the low-temperature and pressure end of the Plant.

g. The final condenser does not provide an acceptable approach to the condenser water temperature.

h. The deaerator is providing excellent decarbonation and deoxygenation service but the economy of stripping steam is not as good as desired.

i. Effective and reliable metallic materials of construction have been proved to be acceptable for most services, however, their costs are not competitive in some cases.

j. Polyphosphate is an effective anti-scale treatment when applied to sea water at proper concentration factors and temperatures.

k. Notched weirs are effective brine distributors for the LTV evaporator. Proper distribution affects both scale deposition and heat transfer by preventing tube starvation.

l. Canned vertical turbine pumps are effective in handling sea water under low net positive suction head conditions. They are more reliable and require less maintenance than the double suction centrifugal pumps.

m. Gypsum-type scale can be washed out with cold sea water from the evaporator tubes. Rate of washing can exceed 300 LB/HR if the tubes are sufficiently open to allow a reasonable cold sea water flow.

n. When production is increased to approach the capability of the heat transfer surface, the limited surface in effects IV and X will be the restrictive factor.

2-57 RECOMMENDATIONS.

2-58 The recommendations presented in this part of the Annual Report will be confined to items that are not directly required as a part of Development Program. They are grouped for discussion under the following categories: process equipment; mechanical equipment; and instrumentation and control equipment.

2-59 Process Equipment.

2-60 The recommended process equipment changes and/or modifications are listed below:

a. Apply at least two types of coatings in one or more of effects IV through IX. One of these coatings should be heat cured while the other be air dried. It is quite possible that relatively inexpensive coatings may be beneficial in these vessels.

b. Replace all carbon steel plugs in evaporator tube sheets with aluminum brass plugs. Fill the top tube sheet plugs with an epoxy potting compound. This modification will reduce maintenance time that is required for testing and plugging of leaks during each outage as well as improve the economy of the Plant by eliminating steam and condensate leaks through the existing plugs.

c. Modify distributor plates, water boxes, and brine connections for effects XI and XII. Re-install pump P-20 so it can transfer brine from effect X to effect XI. Circulate brine from effect X to effect XI and from effect XI to the top water box of effect XII. Install a new line to the existing opposite connection at the top water box of effects XI and XII to provide a variable recycle flow. This will eliminate violent flashing in the cones of effects XI and XII and introduce a slightly less concentrated brine to the tubes. It will also eliminate bypassing and stratification in these effects.

d. Install a vapor connection with a differential pressure control valve to bypass vapors from the water box of effect XII to the 312 condenser. This installation will reduce carryover from effect XII.

e. If pumping capacity is increased (refer to paragraph 2-62a, below), add (increase) heat exchanger surface in effects IV and

X. This will eliminate a significant restrictive factor to increased productivity.

f. Improve the baffling in the 312 condenser and attempt to improve the approach temperature of this heat exchanger.

g. Repair the coating of the split product tank or replace the tank with a glass reinforced plastic tank..

h. Expand the use of non-metallics in piping and in other applicable equipment throughout the Plant.

i. Modify the deaerator tank to provide for on-stream removal of silt. This modification will aid in correcting low net positive suction head problems.

2-61 <u>Mechanical Equipment.</u>

2-62 The recommended mechanical equipment changes and/or modifications are listed below:

a. Relocate, modify and/or replace existing pumps throughout the Plant to provide a pumping capacity that is consistent with the available heat transfer surface. This will provide better indications of the real capacity of the heat exchange equipment (heart of the process), and furthermore, it will provide valid determinations of process limitations.

b. Replace the existing screen on the traveling screen with a finer mesh (when replacement is necessary).

c. Replace the vacuum system block valves with ball or plug valves. The latter valves will provide a more positive shutoff and thereby permit the vacuum equipment to function at a greater peak capacity. They will also eliminate the possibility of equipment damage resulting from chlorine leakage during shock chlorination treatments.

d. Overhaul the air compressors so they can provide a reliable and clean source of instrument air.

2-63 <u>Instrumentation and Control Equipment.</u>

2-64 The recommended instrumentation and control equipment changes and/or modifications are listed below:

a. Install an automatic pH controller for the deaerator. This equipment will provide complete control of one more variable and will improve the performance and efficiency of the process.

b. Repair or replace the evaporator level control linkage to provide a more positive and proportional response. The mis-match of valve stem and gear drive has reached the limit of wear that is consistent with reliability.

c. Convert the apparent heat transfer coeffieient instrument system to a more valuable service.

d. Initiate new analysis and reporting techiques for residual carbon-dioxide measurement of various process streams.

III. PLANT ACTIVITIES

3-1 GENERAL.

3-2 The Plant activities for the period covered by this Annual Report is prefaced by a short operating history, spanning the first two fiscal years of Plant operation.

3-3 OPERATING HISTORY.

3-4 July 1, 1963 marked the beginning of the third year of operation of the Freeport Saline Water Conversion Demonstration Plant No. 1. Before delving into the operating trends of this year, a review of the operations during the first two years will be helpful.

3-5 Table 3-1 compares the total production and cost figures since the startup of Plant operation. In addition to the fiscal year periods, a continuous period of similar length, wherein only routine and normal maintenance was performed, is included. The current fiscal year shows an increase over prior years in both total production and percentage of on-stream time. Costs per 1000 gallons show a corresponding decrease. These figures are based on actual expenditure and include contractor's fixed fee and G & A expenses. The "Normalized Plant" figures are shown elsewhere in this Report.

3-6 PRODUCTION IMPROVEMENTS.

3-7 Improvements in production were achieved by numerous changes in the Plant process equipment and in operating conditions. Early in the first year, it became apparent that operation at a final concentration factor of 4. 0 was not practical because of heavy scale formation ($CaSO_4 . 2H_2O$) in XI and XII effects. A slower rate of scale formation was achieved by operating at a final concentration factor of 3. 0. Until the scaling problem could be solved, it was decided to operate at a final concentration factor close to 3. 0. This necessitated the installation of larger size impellers in the small brine pumps (P-17 through P-22) to handle the increased flow of brine through the low pressure effects.

3-8 Another improvement that was accomplished during the first year was the replacement of the carbon steel tubes and tube plugs in certain heat exchangers. The tube plugs were the first to fail and they were replaced with plugs of the same material as the tubes in each exchanger. These failures forewarned of the necessity at a later date to replace the tubes in those heat exchangers which were tubed with carbon steel. Late in

	April 1961 to June 30, 1962	July 1, 1962 to June 30, 1963	July 1, 1963 to June 30, 1964	April 1, 1963 to March 31, 1964
Total On-Stream Time for Period (Days)	291	243	282	298
Total On-Stream Time for Period (% of Calendar Days)	73.66	66.58	77.26	81.64
Total Production for Period (Gallons)	256,021,695	250,183,830	289,807,260	308,507,100
Avg. Production Per Stream Day (Gallons)	879,981	1,029,563	1,027,685	1,035,259
Avg. Production Per Calendar Day (Gallons)	648,156	685,435	793,992	845,225
Avg. Production Per Calendar Day (Exclusive of Periods of Abnormal Maintenance) (Gallons)	688,230[1]	817,594[2]	854,403[3]	871,030[3]
Total Cost Per 1000 Gallons (Dollars)	2.0191	1.8223	1.3553	1.2812
Total Cost Per 1000 Gallons (Exclusive of Periods of Abnormal Maintenance) (Dollars)	1.4217[1]	1.4052[2]	1.2595[3]	1.2432[3]
Direct Costs Per 1000 Gallons, (Dollars)	1.2913	1.2391	0.812	0.794
Direct Costs Per 1000 Gallons (Exclusive of Periods of Abnormal Maintenance) (Dollars)	0.845[1]	1.0172[2]	0.755[3]	0.794[3]

Periods of abnormal maintenance are defined as follows:

1. Costs for period of September 9, 1961 to October 2, 1961 not included (Hurricane Carla period)
2. Costs for period of January 1, 1963 to March 1, 1963 not included (retubing 5 evaporators with Aluminum-Brass tubes)
3. Adjustment made for more frequent shutdowns made necessary by increased development program activity

TABLE 3-2. COMPARISON OF TOTAL PRODUCTION AND COST FIGURES FOR THREE YEAR PERIOD the first year, all exchangers with carbon steel tubes (201, 301, 305, 208, 308, and 211) were retubed with aluminum brass tubes.

3-9 During the first winter, freezing weather necessitated Plant shutdowns on several occasions. Freeze protection of instruments, purge water lines, and other critical points eliminated freezing weather as a deterrent to continuous production.

3-10 The inadequacy of pumps P-4 and P-5 for their service was demonstrated early by shaft and impeller failures in both pumps. Installation of ni-resist impellers and monel shafts improved the reliability of these pumps; however, an entirely different design has proven suitable for this service.

3-11 Acid treatment of the sea water feed followed by deaeration to remove carbon dioxide and oxygen was adopted as the means of controlling carbonate scale formation and corrosion. The alternate scale control technique of magnesium hydroxide seeding was abandoned due to excessive silt buildup in the tank and rapid calcium carbonate formation.

3-12 The second year of operation produced additional improvements to the Plant process equipment. Many of these improvements were incorporated to increase Plant production by minimizing downtime and increasing operation efficiency. At this point, however, it became apparent that in order to make further process improvements, a closer look at the technical aspects of the operation would be required. Certain modifications adopted during the second year were accomplished for the sole purpose of obtaining the necessary technical data, and thereby, enable a more comprehensive evaluation of the Plant and process.

3-13 One of the numerous instrumentation changes adopted during the second year was the installation of automatic brine level controllers on each evaporator. Another instrumentation change was the installation of a pH recorder of the sea water feed at the outlet of the deaerator tower. This installation enabled a more precise control of acid and caustic feeds. Still other instrumentation changes consisted of numerous thermowell and pressure tap installations. Also, for observation of evaporator operation, port-hole sight glasses were installed in various evaporators.

3-14 MAINTENANCE HISTORY.

3-15 To minimize maintenance, heat exchangers 202 through 207 were bypassed during the second year; there was little temperature pick up across these exchangers and they required excessive maintenance.

In addition, piping sections coated with baked on phenolic material were installed in the discharges of P-11, P-12 and P-13 pumps.

3-16 The acid injection system was a continuous maintenance problem until the sea water piping at the point of injection was replaced with saran-lined pipe. The mixing orifices, downstream of the injection point, are also saran lined. The one-inch acid feed line is carbon steel except for the portion that ties into the sea water feed line. This portion is teflon and extends through a saran-lined nozzle in the sea water feed line. The addition of an acid feed pump of improved design, provided a reliable acid feed system.

3-17 Approximately halfway through the second year of operation, it became necessary to retube all evaporators which were originally tubed with carbon steel. Pitting type corrosion and an affinity for local scale attachment caused their failure. The replacement tube material was aluminum brass.

3-18 A piping change to eliminate recirculation of concentrated brine to the feed was incorporated to improve Plant operation. The change was accomplished by discharging the effluent brine to a waste canal on the North side of the Plant. In an attempt to improve the operation of XII effect evaporator, piping was installed to bring the brine from XI effect into the vapor body of XII effect and pump brine from the vapor body of XII effect to the top of XII effect evaporator. This modification was primarily accomplished to reduce the amount of vapor being forced through the tubes of the heating element. It also minimized liquid starving, and thereby, decreased scale formation in the tubes.

3-19 OPERATING ACTIVITIES.

3-20 Efforts to improve over-all Plant operation were continued during the third fiscal year. These improvements included: reducing maintenance; evaluating various operating modes; standardizing operating procedures; making additional Plant modifications; and obtaining more detailed knowledge of the technical aspects of the process. The improvements of this year, coupled with the improvements of the first two years, produced record total production and total on-stream time (refer to Table 3-1). Total production for the 1964 fiscal year might have been higher; however, Plant shutdown frequency was increased in the latter part of the year to accelerate the acquisition of developmental information pertinent to all aspects of the process. These development programs have an adverse effect on operating costs but will eventually result in major operating improvements.

3-21 VARIABLES AND PARAMETERS.

3-22 Attempts at holding process variables constant during a given period have been quite successful. Piping was installed to recirculate a controlled amount of concentrated brine to the sea water feed; and thereby, maintain the concentration factor of the incoming sea water between 0.9 to 1.1 when necessary. (Low inlet concentration factors result in erroneous comparisons.) During the winter months, warm sea water (312 return) was intentionally blended with the sea water feed (at the intake pit) in order to control XII effect vacuum and temperatures. The pH control of sea water to the deaerator has been greatly facilitated by a piping modification which utilizes the clarifier-thickener tank as a pre-settling basin. This modification provided a constant suction head for P-3 pump, and thereby, improved its operation. Another change, which contributed to better pH control of the sea water to the deaerator, was the relocation of the lower impulse line to the deaerator level control. Its original location was quite close to a pump suction nozzle and the turbulence in the vicinity led to erratic level control and fluctuating sea water feed rates.

3-23 OPERATING MODES.

3-24 Emphasis has been placed on adopting standard operating procedures which would maintain operating variables at a constant level. Startup, shutdown, chemical addition, and process changes all are performed according to prescribed procedures. Some of these procedures are presented in the Appendix of the Report.

3-25 A major improvement in over-all Plant reliability was achieved when a vertical-turbine-type, canned pump was installed to deliver deaerated sea water feed to the system. This pump is flexible enough in its operation to continue pumping under extremely low suction head conditions. The greater flexibility simplifies trouble shooting during upset conditions and minimizes shutdowns.

3-26 In order to reduce iron pick up by low pH water, piping was installed to vent the non-condensable vapors from II and III effects to the deaerator rather than to the succeeding effects. These vapors contain carbon dioxide gas formed by decomposition of carbonates and bicarbonates in the sea water feed to the first effect evaporator. It is conceivable that a significant portion of carbon dioxide could dissolve in the product water, lowering its pH and causing a measurable quantity of iron to dissolve from the pipe and vessel walls. Late in the year, an additional piping was installed to vent these vapors to the atmosphere as well as to the deaerator.

3-27 One evaluated mode of operation was the use of vapors from II and III effects as the only source of stripping medium in the deaerator. Even though there is some carbon dioxide present in these vapors, the driving force between the CO_2 dissolved in the sea water and that in the vapors is sufficient to remove most of the CO_2. Effluent sea water alkalinities during this evaluation were in the range of 8 to 10 PPM. Oxygen content of this sea water was in the range of 100 PPB. This mode of operation may be desirable; however, further evaluation is being conducted to determine whether it has an over-all process advantage.

3-28 The amount of venting from the heating elements of each evaporator was evaluated. Venting is necessary to remove non-condensable gases. Without venting, these gases will build up and effectively blanket the heat transfer surface; and thereby, create a decided drop in the rate of heat transfer across the tubes. An excessive venting rate bypasses steam and reduces over-all efficiency. The results of this venting study indicated that the evaporators could operate with the lower vents completely closed and the top vent slightly open. The venting rate on effect II and III evaporators was maintained at a higher rate than the others because of the possibility of CO_2 buildup from incomplete decarbonation in the deaerator. As a result of this evaluation, venting rates were adjusted which has effectively improved operation. Measurement of actual venting rate is only possible for the combined flow from effects II and III.

3-29 <u>MAINTENANCE ACTIVITIES.</u>

3-30 Extensive efforts were made throughout the year to correlate the amount of scaling present in an evaporator with the change in temperature drop across the evaporator. It was determined that a relationship exists between the apparent heat transfer coefficient and the amount of scale present in the evaporator heating elements. (Refer to the Appendix for sample calculations and typical data.) As the scale builds up, the apparent heat transfer coefficient falls off. This information could greatly assist a production plant in scheduling maintenance and in measuring the extent of lost efficiency occasioned by plant upsets. It can also be an effective tool to forewarn of any plugging due to scale accumulation.

3-31 SCALE REMOVAL.

3-32 Scale removal has been a persistent maintenance problem that necessitated excessive maintenance manhours. Several techniques for removal were attempted including the use of high pressure water (3000-4000 PSI) jets. Although this method was effective, the high-pressure equipment rental costs were quite expensive. Washing the scale out with cold sea water was attempted as an alternate method. This method proved quite effective and resulted in savings of up to 100 man hours per shutdown. To attain satisfactory results by this method, the washing must be initiated before the scaled tubes become completely plugged. Although the method is effective in removing the gypsum scale from effect X, XI, and XII evaporators, its effectiveness in removing calcium sulfate anhydrite scale from effect I evaporator is yet to be determined.

3-33 SCALE PREVENTION.

3-34 A successful means of preventing gypsum scale in X, XI, and XII effects while operating at a final concentration factor of 3.0 was developed by the addition of a polyphosphate compound to the suction of the brine discharge pump on effect VIII evaporator. Sufficient polyphosphate compound was fed that would result in a concentration of 4 to 5 PPM based on the sea water feed rate. Tests to determine the effectiveness of this material at a final concentration factor of 3.5 are currently being conducted.

3-35 FREEZE PROTECTION.

3-36 Maintenance downtime was further decreased by the added installation of freeze protection on critical equipment. Operations during the first two years defined the areas which needed the most protection; and additional electric heater tapes and insulation were installed in these areas. These protective measures and the implementation of a standard freeze protection procedures permitted operations at temperatures as low as $20^{\circ}F$.

3-37 EVAPORATOR SILT BUILDUP PREVENTION.

3-38 The installation of necessary piping to use the clarifier-thickener as a pre-settling basin was a maintenance-time-saving modification. If the silt in the sea water feed could be precipitated before the water is fed to the evaporators, cleaning of the cones would become unnecessary. This modification was completed near the end of this report period and has not yet been fully evaluated.

3-39 EQUIPMENT PERFORMANCE.

3-40 INSTRUMENTATION.

3-41 pH Cell and Recorder.

3-42 Very little difficulty has been experienced with this piece of equipment.

It is a valuable operating tool that assists in the optimum control of plant variables. In the Freeport Plant, the pH cell and recorder is connected to continuously record the sea water feed pH. An additional recorder or a manifolding system to monitor product water pH would be helpful, particularly for large-scale production plants. The importance of recording product water pH increases with the use of stabilizing treatments.

3-43 Conductivity Cells.

3-44 Conductivity cells are valuable indicators of individual evaporator product water quality. The only maintenance they require is an occasional cleaning; however, they can be damaged by careless handling. They have additional value in that they assist in determining evaporators that have tube or plug leaks. During operation, leaks in the tubes will occur from the steam side to the sea-water side and are difficult to detect. In addition, at the time of shutdown, sea water can leak into the condensate system and an approximate location of these leaks can be determined by means of conductivity cells.

3-45 Apparent Heat Transfer Coefficient Meters.

3-46 These instruments have some value as an operating tool; however, their utility do not justify their cost. The values they register are not actual heat transfer coefficients but are the actual vapor-to-vapor temperature differences across each evaporator, and are recorded as heat transfer coefficients. The same information is duplicated from the dynalog temperature indicator; and therefore, the transfer coefficient meter should be functionally converted to record temperatures elsewhere in the Plant. It could be utilized to a great advantage in determining the brine or vapor temperatures at the top and bottom of the evaporator tubes.

3-47 Automatic Level Controls.

3-48 The incorporation of automatic level controls into the Plant process equipment has been a significant 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.

3-49 The level in the deaerator is controlled by a butterfly valve located in the sea water inlet line to the tower. Initially, this method of control resulted in considerable level variation; however, relocation of lower impulse line and installation of a valve positioner on the control valve have greatly improved the control technique.

3-50 Modifications to standardize the deaerator tower level control have recently been completed. These modifications were incorporated to accommodate the future installation of automatic pH control equipment on the deaerator feed line.

3-51 Compax Level Controls.

3-52 Each evaporator is equipped with a Compax level controller. These controllers require considerable maintenance such as purge cleaning and lubricating of gear operators. However, their utility value far exceeds their maintenance costs.

3-53 The original evaporator level controllers were manually controlled with a butterfly valve. When the butterfly valve was converted to automatic control, the conversion was improvised with existing equipment. Although no serious difficulty has been encountered with this improvised conversion, the valve stem - drive gear connection is excessively worn. Either the valve stem or the drive gear should be replaced.

3-54 Temperature Measurement.

3-55 The existing temperature measuring instruments have been satisfactory from a maintenance standpoint. However, numerous measuring points have been added to acquire more process evaluation information. Several of the original thermowells were too short to sense accurate temperatures. In some cases, the thermowells did not penetrate the liquid stream. All replacements of original thermowells have been of longer length and sized to penetrate the centerline of the pipe.

3-56 Flow Measurement.

3-57 Maintenance of flow measuring instruments has been routine. The product flow meters are checked regularly and calibrated when necessary. The steam and sea water feed flow meters are checked and calibrated whenever such action is indicated from the material balance information. The steam flow recorder was modified to include condensate pots on the recorder impulse lines. These pots function to stabilize the impulses which result in steadier chart records.

3-58 The flow of vapors to and from the deaerator is presently read on dial-type indicators that are located in the Control House. A recording attachment for these flows would be beneficial in deaerator studies.

3-59 Condensate Purity Instrument.

3-60 The condensate purity instrument has required no maintenance other than occasional tube replacement.

3-61 Pressure Gages.

3-62 The pressure gages have been fairly reliable for the service they perform. Pressure gages that provide salt water service maintain their calibration for only a short period of time. Before heat balance data are logged, these gages are checked and calibrated (if required). The vacuum gages and the pressure gages that provide fresh water service are more reliable; but again, they are checked and calibrated (if required) before important readings are logged.

3-63 Alarms.

3-64 The alarm system has been adequate, and has required very little maintenance. Alarm system includes: high and low evaporator level alarm; low pH alarm; and high condensate conductivity alarm.

3-65 EVAPORATORS.

3-66 First Effect.

3-67 The first effect evaporator is exposed to sea water at temperatures above 240°F. The water box and tube sheet have experienced considerable corrosion and erosion damage (see Figure 3-1). The vapor body has corroded to a lesser extent than the water box. Adequate means of protection will be required for full scale production units. Two protective coatings have been tested at Freeport: Dampney Apexior #1 appeared to have some temporary value; and an epoxy coating of Brutem 30 failed in less than 30 days (see Figure 3-2). A baked on phenolic material has provided effective service in brine piping from the first effect, and probably would be suitable for service inside of the evaporator. The rolled-in carbon steel plugs have all failed and have been plugged with wood plugs. Replacement of carbon steel plugs with rolled-in, aluminum brass plugs would provide a more effective repair job.

3-68 Second Effect.

3-69 The second effect evaporator is in much the same condition as the first. Some of the pitting is a little deeper in this water box; however, this is functionally due to the amount of clearance between the vessel wall and the distributor plate. Accelerated corrosion and erosion takes place if salt water is allowed to pass between the distributor plate and

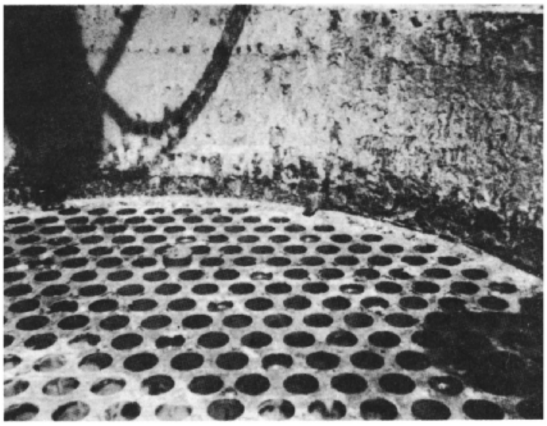


FIGURE 3-1

TYPICAL EVAPORATOR WATER BOX - NOTE THE DEEP PITS JUST ABOVE THE TUBE SHEET.

FIRST EFFECT EVAPORATOR WATER BOX, SHOWING TUBE INSERTS - NOTE THE EPOXY COATING WHICH HAS DETE-RIORATED AND FALLEN.

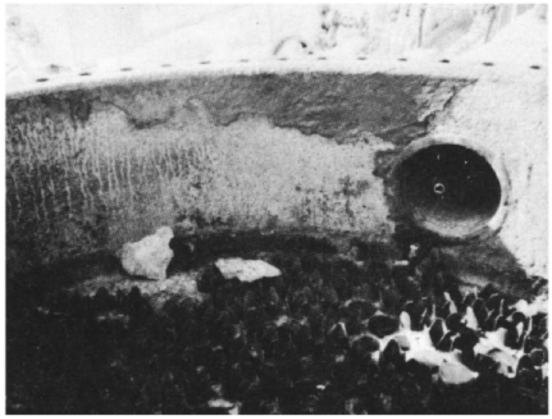


FIGURE 3-2

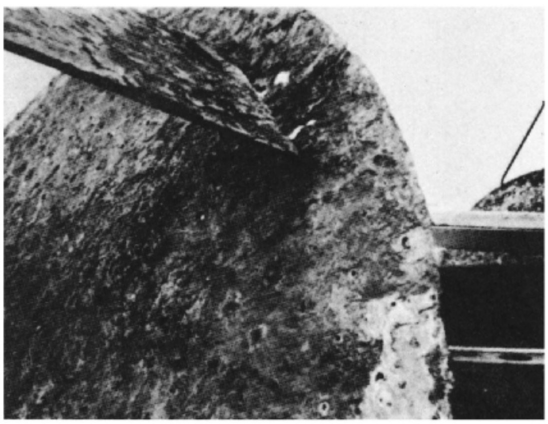


FIGURE 3-3

SECOND EFFECT EVAPORATOR DISTRIBUTOR PLATE - NOTE THE HOLES ADJACENT TO IM-PINGEMENT PLATE CAUSED BY CORROSION AND EROSION.

vessel wall. A coating of Brutem 30 epoxy has been moderately successful in this effect. At the time of this report, it has been in service approximately 90 days and shows some signs of deteriorating. The carbon steel plugs have also failed and been repaired or replaced in this evaporator. The carbon steel distributor plate has failed and has been replaced (see Figure 3-3).

3-70 Third Effect.

3-71 The third effect evaporator is in much the same condition as the first two except there are less tube plug failures and less actual corrosion in this effect. Brutem 30 epoxy coating has been in service approximately 90 days and appears to be in good condition.

3-72 Fourth Through Ninth Effects.

3-73 These evaporators reflect progressively less corrosion in their water boxes and vapor bodies. A less expensive protective coating material could possibly be justified in all evaporators of these effects.

3-74 A few of the carbon steel tube plugs have also failed in these evaporators. Should these plugs be replaced with rolled-in aluminum brass plugs, it is recommended that the upper plugs be filled with an epoxy potting compound to eliminate pocketing and subsequent concentration of brine.

3-75 Tenth Effect.

3-76 The corrosion in this effect is significantly less than that in the warmer effects. Piping was installed in this effect to feed from this effect to the XI effect vapor body in lieu of the water box. Other modifications include: addition of port-hole sight glass to observe liquid level above the distributor plate; and installation of notched wiers to improve brine distribution.

3-77 Eleventh Effect.

3-78 The eleventh effect evaporator is in good condition. The same modifications as listed above for the tenth effect were performed on this effect. In addition, an unsuccessful attempt was made to measure the heater outlet, fluid temperatures by the installation of a thermowell and support trough under the heating element.

3-79 A major maintenance task in the past was the drilling out and removing of scale from the tubes of this evaporator. This task is no longer necessary. An effective maintenance technique has been developed wherein sea water is periodically circulated through the evaporator as the scale begins to form. The present operation, utilizing this maintenance technique and brine distribution wiers, does not form scale in this effect while operating at final concentration factors up to and including 3. 0.

3-80 Twelfth Effect.

3-81 The twelfth effect is also in good condition. The present operation utilizes the sea water washing technique, the brine distribution wiers, and a polyphosphate addition; as a result, no scale is formed at concentration factors up to and including 3. 0. If the polyphosphate addition is discontinued, scale will form slowly in this effect.

3-82 Distributor Plates.

3-83 To supplement the discussion of the distributor plates, Figures 3-4 through 3-7 are included. Figure 3-4 depicts the distribution pattern during a normal flow rate. Note how the high velocity of the brine underneath the impingement baffle causes a lower fluid head on the plate in the immediate vicinity of the baffle. The fluid head at the opposite end from the baffle is considerably greater, resulting in higher liquid flow rates in this area. This explains the observed scaling pattern. Blocking the space under the baffle (see Figure 3-5) makes the fluid head more uniform on the distributor plate. Figure 3-6 again depicts the head distribution at normal flow rate. Figure 3-7 depicts the same evaporator with a very high flow rate to the distributor plate. The high flow rate has a tendency to result in a more uniform fluid head. These figures illustrate the need to adequately redesign the water distribution system. Figure 3-8 illustrates the accumulation of product corrosion chips and silt which results in poor brine circulation. Notched wier inserts adequately combat this problem, however, a uniform fluid head on the distributor plate is still desired.

3-84 HEAT EXCHANGERS.

3-85 Heat Exchanger 201.

3-86 The water box on this heat exchanger displays considerable corrosive damage. A protective coating applied to the metal surface is urgently needed. Investigations are underway to filed-apply a baked-on phenolic coating.

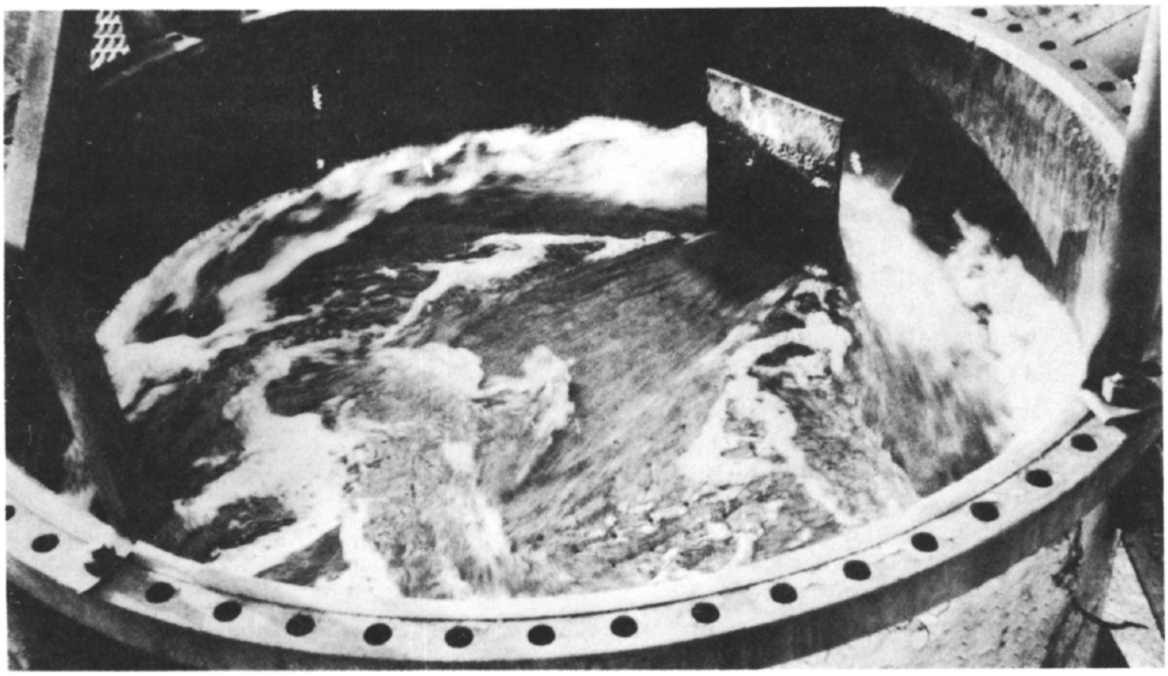


FIGURE 3-4. DISTRIBUTION PATTERN FOR NORMAL FLOW RATES

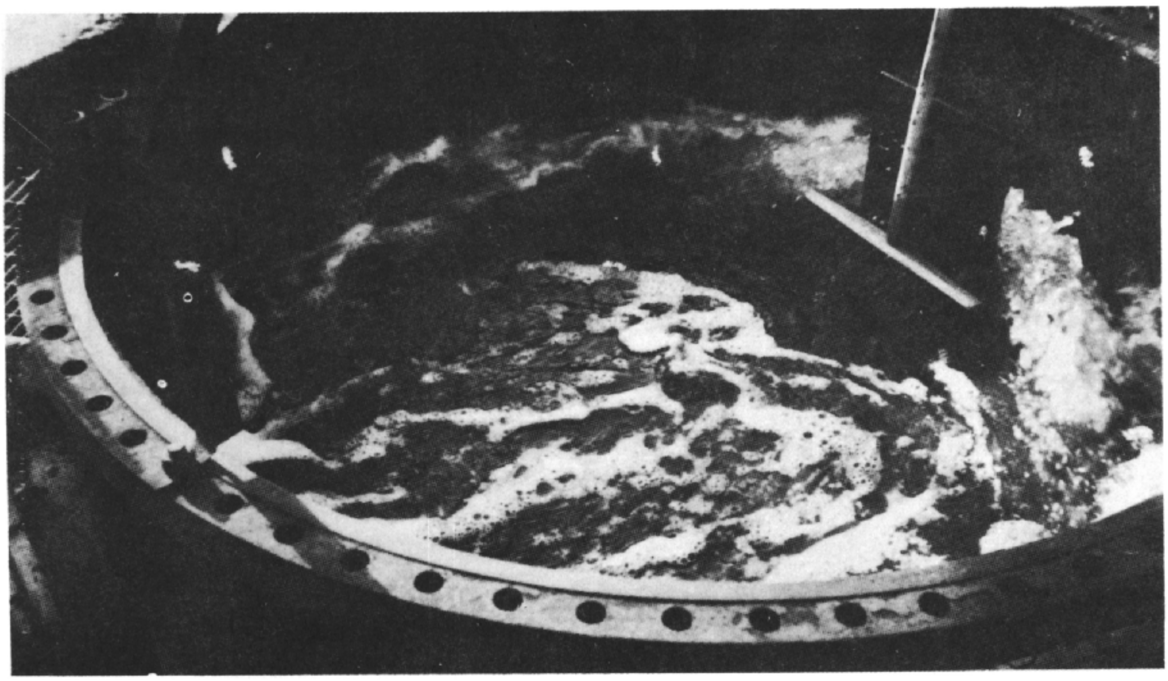


FIGURE 3-5. DISTRIBUTION PATTERN WITH BAFFLE HOLE BLOCKED

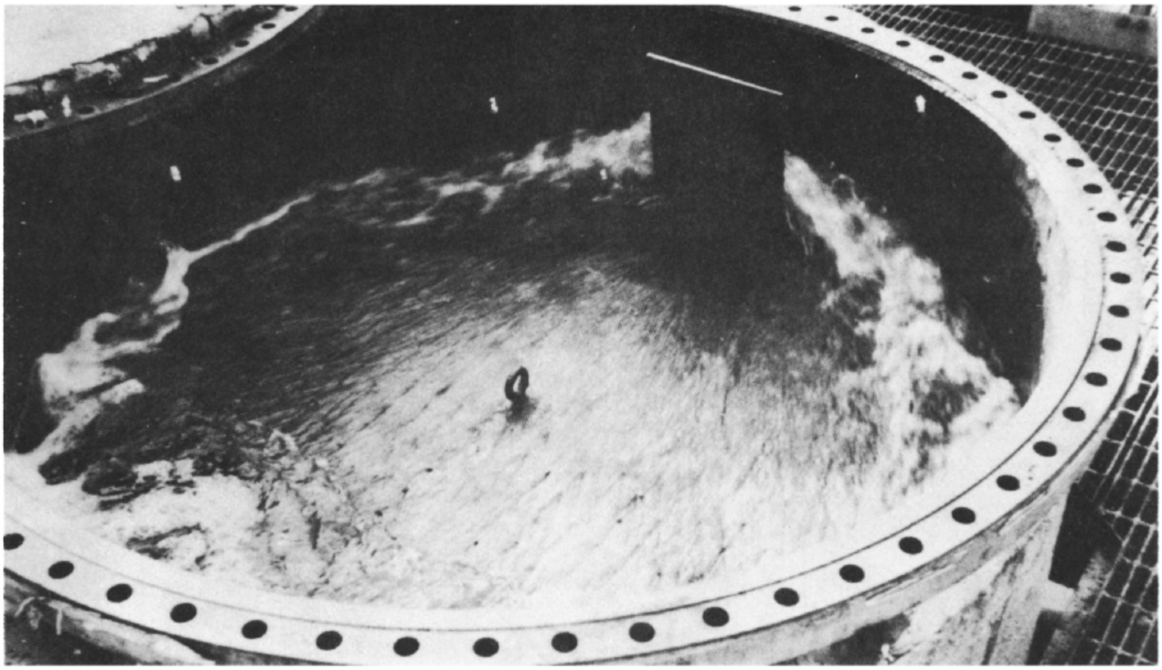


FIGURE 3-6. LIQUID HEAD DISTRIBUTION DURING NORMAL FLOW

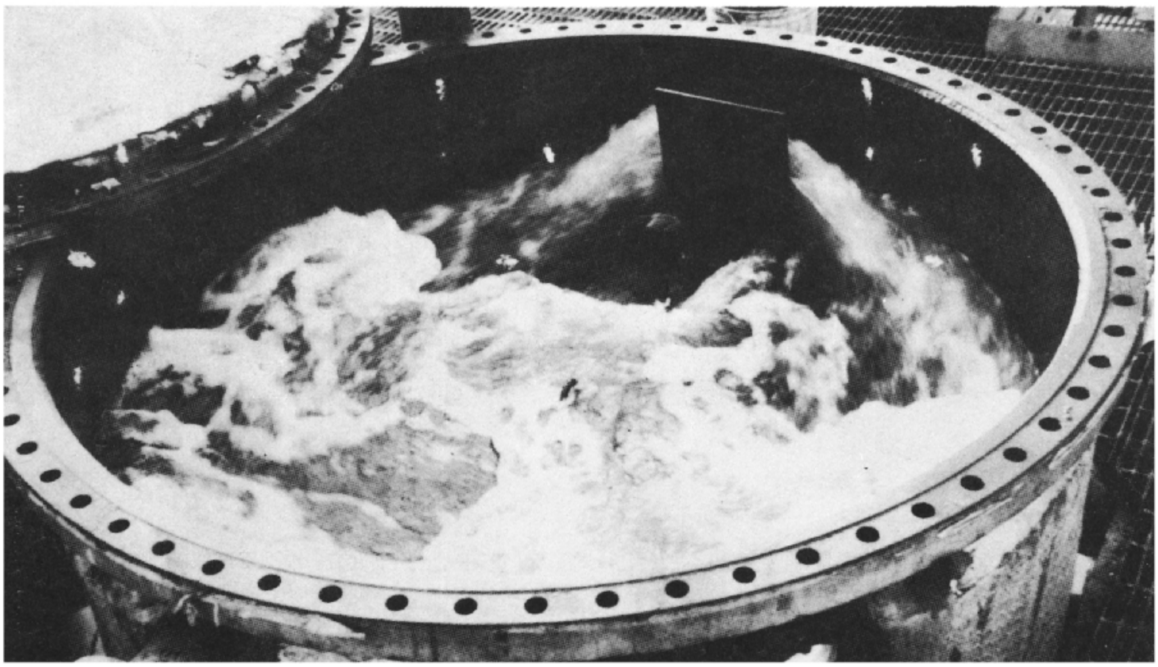


FIGURE 3-7. LIQUID HEAD DISTRIBUTION DURING HIGH FLOW

FIGURE 3-8. CORROSION CHIPS AND SILT DEPOSITS ON DISTRIBUTOR PLATE

3-87 Heat Exchangers 208 Through 211.

3-88 Little maintenance has been required for these exchangers. The water boxes display some signs of corrosion. Heat exchanger 209 developed a leak from the water box to the atmosphere, however, this leak was repaired with a weld patch.

3-89 Heat Exchangers 212 Through 214.

3-90 These heat exchangers function in the nondeaerated sea water circuit, and consequently, they require cleaning at more frequent intervals than the others.

3-91 Inhibited hydrochloric acid has been used as a cleaning solution but it is recommended only as a last resort. There have been indications of some tube metal loss when hydrochloric acid is used. Soaking and flushing the exchanger with water is an adequate cleaning technique.

3-92 Heat Exchanger 215.

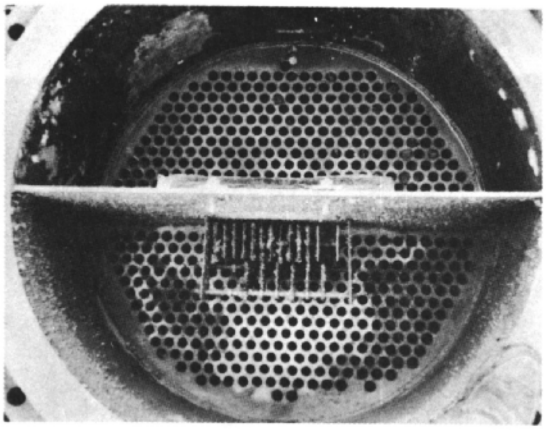
3-93 This heat exchanger is the first one in the nondeaerated sea water circuit. It accumulates shells and marine life during extended runs; however, periodic opening and cleaning operations obviate the necessity of chlorine injection to control marine life (see Figure 3-9). This heat exchanger also functions as the final cooler for fresh water that is delivered to the City of Freeport. For this reason, precautions are taken to flush the exchanger during each shutdown.

3-94 300-Series Heat Exchangers.

3-95 Most of the 300-series heat exchangers have required only routine cleaning and inspection. However, the water boxes and pass baffles are deteriorating, especially at the high-temperature end of the process. Figure 3-10 illustrates a typical pass baffle failure. This type of failure has occurred in heat exchangers 301, 304, and 306. Figure 3-11 illustrates a failure in the shell of heat exchanger 302. A patch was welded over the hole on the outside and a 316 stainless steel liner was welded on the inside of the shell. This method of repair will be evaluated.

3-96 FINAL CONDENSER.

3-97 The only maintenance required by the 312 condenser is periodic opening and inspection to determine the cause of suspected plugging. Occasionally, silt builds up and impedes the flow of cooling water, however, this residue can be readily washed out by reverse flow. Figure 3-12 illustrates the cause of one such plugging. During the run just



HEAT EXCHANGER 215 WATER BOX AFTER 30 - DAY RUN NOTE THE LACK OF SHELLS AND MARINE LIFE

FIGURE 3-9

HEAT EXCHANGER 306 WATER BOX - NOTE PASS BAFFLE DETERIORATION. SIMILAR DETERIORATION HAS OCCURRED IN EXCHANGERS 301 AND 304

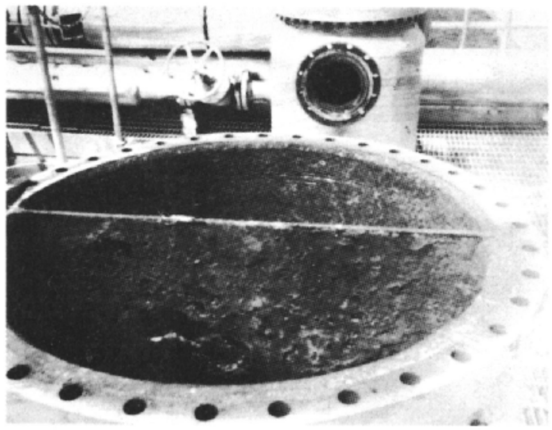
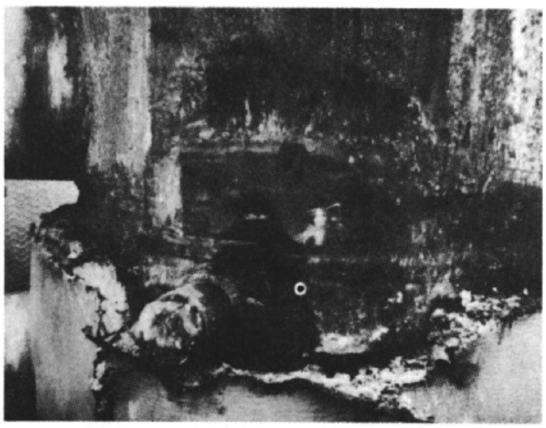


FIGURE 3-10



HEAT EXCHANGER 302 - NOTE THE HOLE IN THE SHELL OF THE EXCHANGER

FIGURE 3-11

prior to the inspection noted by Figure 3-12, the magnesium anodes had completely dissolved and the covers had fallen to the top of the tube sheet.

3-98 DEAERATOR TOWER.

3-99 Maintenance of deaerator tower consists largely of opening, inspecting, and cleaning silt accumulation from the bottom of the tower. The protective, coal-tar-epoxy coating has been quite successful in that portion of the tower that is below the packing. In the packed area, the coating has been only moderately successful. Apparently some relative movement takes place between the packing and the vessel wall. This abrasion breaks the coating and allows the sea water to attack the vessel wall. However, an 18-month life span can be expected from the protective coating.

3-100 Figure 3-13 illustrates the unbroken condition of the packing after 18-months of operation. Apparently, last year's experience of broken raschig rings was caused by the loss of the inlet spray nozzle.

3-101 SPLIT TANK.

3-102 The Freeport-side of this tank is beginning to require an increased amount of maintenance. The original protective coating has failed and the walls display considerable rusting. Some experimenting has been done with a new coating, however, a better solution is required. The rusting walls could be sandblasted and then recoated with a protective coating that is approved for potable water service. A possible replacement of the tank with a filament-wound, glass-reinforced plastic tank should also be considered.

3-103 CHEMICAL STORAGE TANKS.

3-104 These tanks have been adequate for their service. A small savings in chemical cost might be achieved by providing a greater acid storage capacity.

3-105 TRAVELING SCREEN.

3-106 The purpose of this screen is to keep sea weed, debris, and large size marine life from entering the suction of the circulating sea water pump. It performs well and requires little or no maintenance other than lubrication. A finer screen might be used without restricting flow to trap the smaller particles.

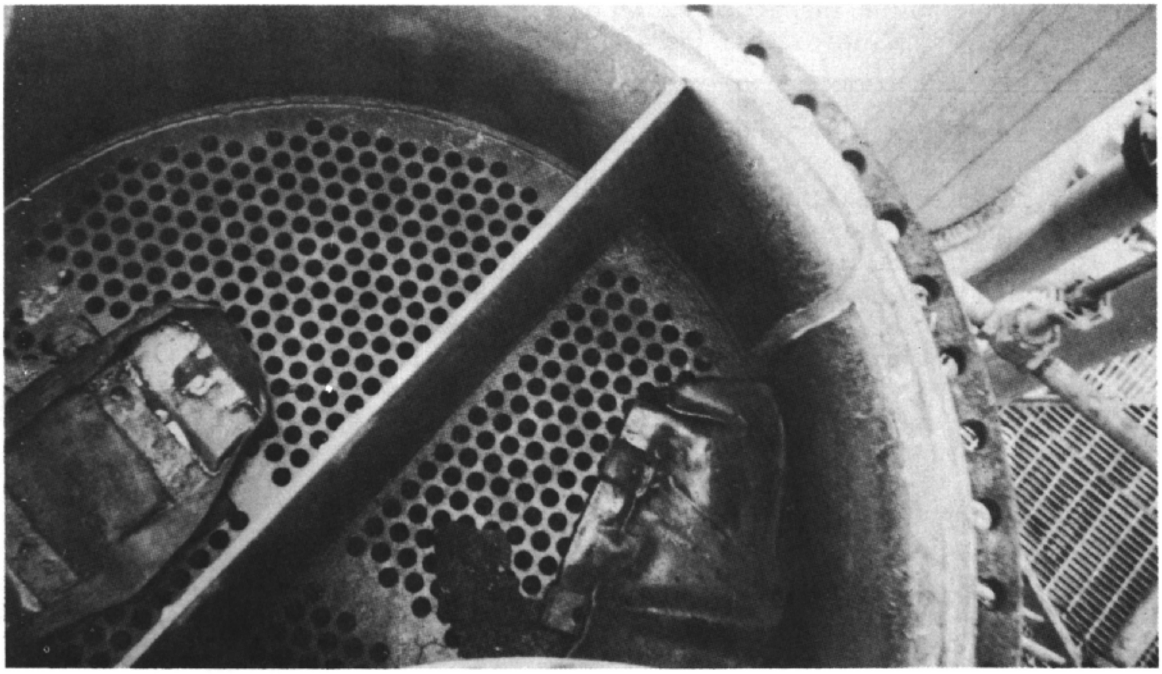


FIGURE 3-12. OPENED VIEW OF CONDENSER 312

FIGURE 3-13. DEAERATOR TOWER PACKING

3-107 SEA WATER INTAKE PIT.

3-108 A considerable amount of fine silt passes through the traveling screen. This material gradually settles in the intake pit and eventually builds up to the level of the pump suction. This silt buildup necessitates a cleanout of the intake pit at least once a year. The most effective method of accomplishing cleanout requires the services of a diver and a dredge pump. This method of cleaning costs approximately $500.00 per year.

3-109 It is important to settle the silt out in the pit, otherwise it accumulates in the cones and on the distributor plates of the evaporators.

3-110 PUMPS AND COMPRESSORS.

3-111 Pump P-1 and Standby P-2.

3-112 These pumps were overhauled in October 1963 and have given excellent service since. The October overhaul consisted of: metallizing and machining the shaft; fabrication and installation of a stainless steel shaft coupling and sleeve; repairing the casing; epoxy coating the bowls; and installing cutlass bearings.

3-113 Pump P-2a.

3-114 In December of 1963, this pump was purchased and installed to replace pump P-2. After three months of operation, a blade on the second stage propeller broke off. The pump was pulled and two new bronze propellers were installed. The pump has performed satisfactorily since that time. Both pumps P-1 and P-2a are scheduled for a complete inspection in October of 1964.

3-115 Pump P-3.

3-116 Maintenance on this pump has been minor. It has consisted of replacing the inboard and outboard bearings and epoxy coating the suction bell. This pump is not spared, and therefore, it must operate continuously whenever the Plant is onstream. It has performed well and has never been the primary cause of a Plant shutdown.

3-117 Pump P-4.

3-118 Little maintenance was required by this pump because it operated very infrequently during the year.

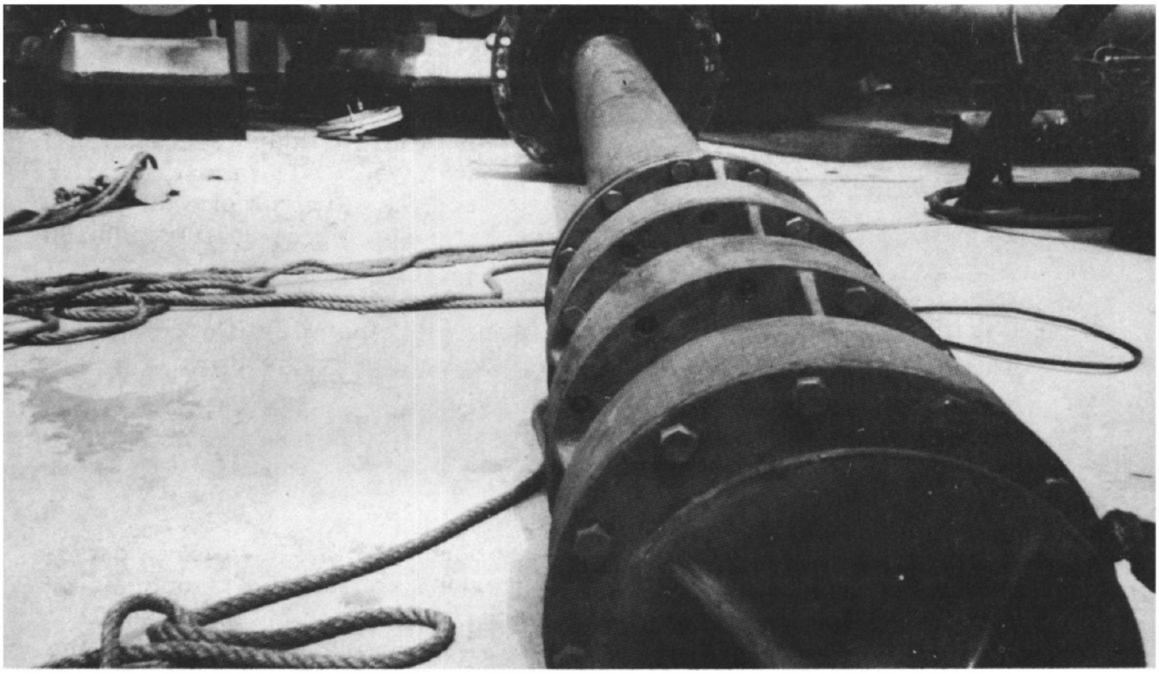


FIGURE 3-14. PUMP P-5a AFTER PULLING

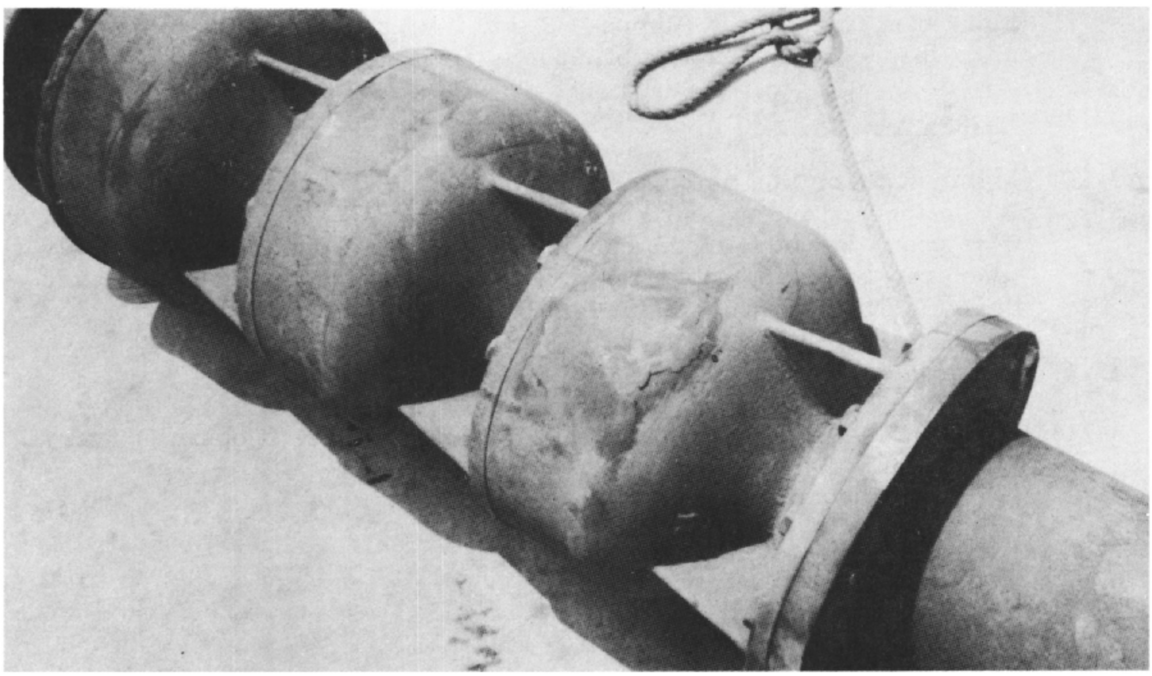


FIGURE 3-15. THIRD STAGE BOWL OF PUMP P-5a

3-119 Pump P-5.

3-120 Maintenance of this pump was held to a minimum in anticipation of replacing it with a vertical, can-type pump in November of 1963.

3-121 Pump P-5a.

3-122 This pump was installed in November of 1963. It was capable of operating under conditions of very low, net positive suction head. Minor difficulties were encountered with the shaft sleeve after five months of near-continuous operation. The snap rings, which hold the sleeve in place on the shaft, failed and allowed the sleeve to ride up on the shaft. This difficulty was solved by securing the sleeve to the shaft with set screws. After six months of operation, a slight vibration in the shaft necessitated a complete shutdown and inspection of the pump. Figure 3-14 displays the pump as it appeared after pulling. Note that some of the bowl cap screw heads are missing. Figure 3-15 displays the third-stage bowl as it appeared after pulling. Note the crack in the top of the casting. A closeup of this crack is shown in Figure 3-16. The middle bowl (second-stage) revealed a spot where the casting had been repaired prior to the pump's original assembly (see Figure 3-17). The suction bell displayed evidence of considerable graphitization and a chipped graphitar bearing(see Figure 3-18). The pump was returned to the manufacturer for modification and replacement of faulty parts. When the pump is re-installed, it should only require periodic bearing replacements.

3-123 Pumps P-11 Through P-16 (Large Brine Pumps).

3-124 Maintenance of the large brine pumps consisted largely of bearing replacements and stuffing box repacking. To minimize bearing replacement, a scheduled lubrication program is strictly enforced.

3-125 Pumps P-17 Through P-22 (Small Brine Pumps).

3-126 Maximum size impellers were installed in the small brine pumps to increase their capacity. This modification has resulted in some difficulties with overloading the electric drive motors, but has improved the over-all operation of the Plant. Maintenance on these pumps has been confined to bearing replacements and stuffing box repackings.

3-127 Brine Pump P-23 (Recirculating Pump for XII Effect).

3-128 This brine recirculating pump operates with the discharge valve approximately half-closed in order to keep cavitation at a minimum.

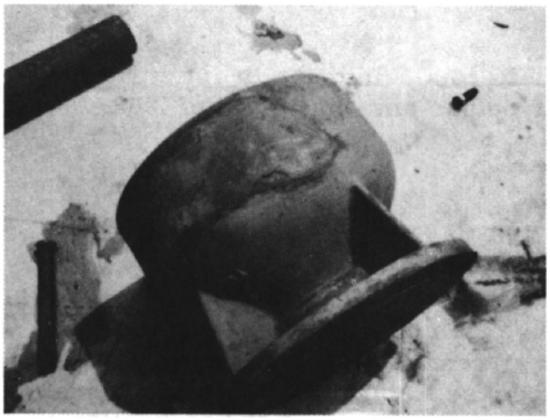


FIGURE 3-16

THIRD STAGE BOWL OF PUMP P-5a - NOTE LARGE CRACK IN THE CASTING

SECOND STAGE BOWL OF PUMP P-5a - NOTE SPOT PUMP WHERE CASTING HAD BEEN REPAIRED PRIOR TO ASSEMBLY

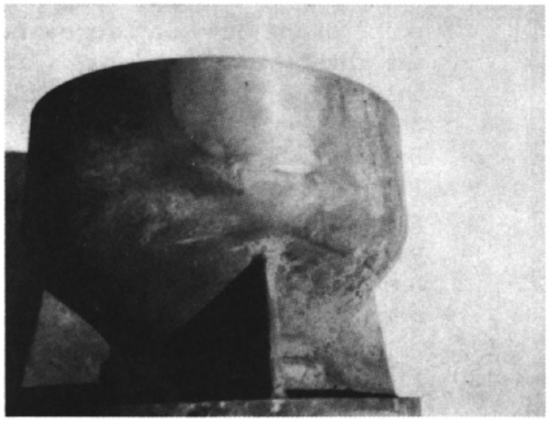


FIGURE 3-17

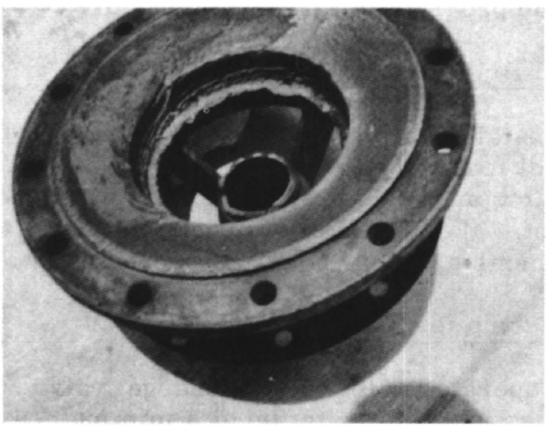


FIGURE 3-18

SUCTION BELL OF PUMP P-5a-NOTE GRAPHITIZATION, CHIPPED GRAPHITAR BEARING, AND EVIDENCE OF EROSION

Cavitation, however, has eroded the suction bell and necessitated the application of protective epoxy coating. Other maintenance performed on this pump included bearing replacement and stuffing box repacking.

3-129 Pumps P-31 Through P-43 (Condensate Pumps).

3-130 These condensate pumps handle a clean, non-abrasive fluid. As a group, they have required relatively little maintenance. The two exceptions are pumps P-40 and P-41. Pump P-40 required a replacement of the inboard and outboard bearings; while a chipped impeller in pump P-41 necessitated a complete replacement of the impeller.

3-131 Pump P-44.

3-132 This pump operates with the discharge valve one-quarter to half open in order to keep cavitation at a minimum. Nevertheless, this pump experienced sufficient cavitation to ruin the impeller (see Figures 3-19, 3-20, and 3-21). The ruined impeller was replaced with a spare.

3-133 Acid Pump P-46 (Lapp Pulsafeeder).

3-134 This acid pump handles 94 per cent sulfuric acid solution which is a corrosive fluid. The maintenance required by this pump has not been unduly excessive considering the service it performs. For the most part, maintenance has consisted of periodic freeing up of the suction and discharge valves and general clean up of the liquid-end of the pump.

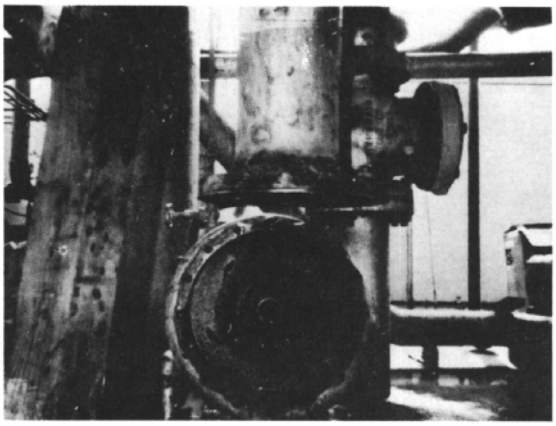
3-135 Product Pumps P-50 and P-51.

3-136 The product pumps have operated very saitsfactorily. They have required no maintenance other than an occasional replacement of packing.

3-137 Instrument Air Compressor P-60.

3-138 The two instrument air compressors require considerable maintenance to maintain their performance at peak efficiency. Up to now, maintenance has primarily consisted of: piston rod and ring replacements; and valve overhauls. In the near future, however, a major overhaul will be required. The overhaul will consist of: reboring the cylinders; and installation of oversize pistons and rings.

3-139 The instrument air requirements of the Plant have greatly increased since the Plant was first designed. For this reason, these compressors operate almost continuously while the Plant is operational.



PUMP P-44 WITH SUCTION BOWL REMOVED - NOTE CAVITATION DAMAGE TO IMPELLER

FIGURE 3-19

PUMP P-44 IMPELLER REMOVED - NOTE THE SHROUD CAVITATION DAMAGE

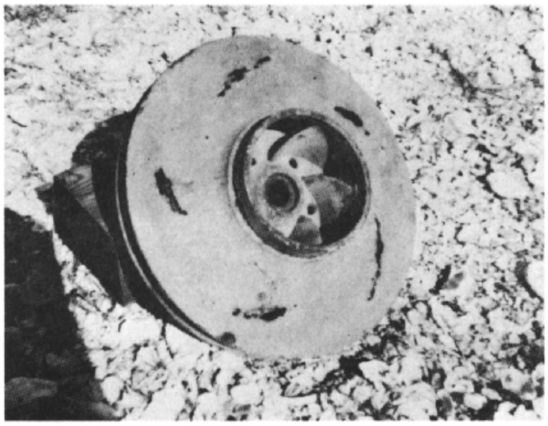
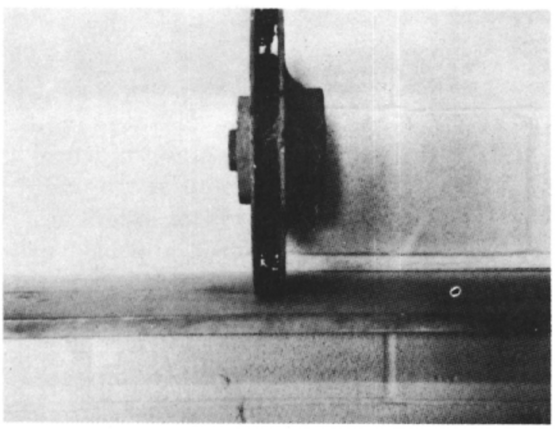


FIGURE 3-20



PUMP P-44 IMPELLER, VERTICAL VIEW - NOTE THE VANE CAVITATION DAMAGE

FIGURE 3-21

3-140 Utility Air Compressor P-61.

3-141 The utility air compressor was installed during the second year of Plant operation. This compressor supplies shop air to operate various air tools, sand blasters, etc. Little maintenance was required on this compressor until June 1964; at this time, the compressor was disassembled, cleaned, and new valves and gaskets were installed.

3-142 Vacuum Pump P-62.

3-143 The vacuum pump operates with much vibration. This vibration apparently sheared off the tie bars between the concrete block foundation and the concrete pad which allowed the block foundation to vibrate with the pump. This condition was corrected by chipping out around the foundation, inserting rebar, and pouring additional concrete to secure the foundation.

3-144 During chlorine treatment of heat exchanger 215, the chlorine solution leaks through the butterfly valve into the internals of the compressor. The chlorine attacks the valve strips and pits the seating surfaces which results in excessive valve replacements. To combat this problem, extreme care is taken to ensure that the block valve is holding while the chlorine is injected. In addition, the frequency of chlorine injection is held to an absolute minimum. Replacement of block valve with a more satisfactory type valve is recommended.

3-145 HEAT TRACING FOR FREEZE PROTECTION.

3-146 During freezing weather, it is necessary to protect small lines and instrumentation that are exposed to freezing temperatures. Presently, this protection is provided by means of electric heater tapes. In future installations of similar function, the installation of steam-traced lines in all critical services is strongly recommended. These steam-traced lines should be installed at the time of original construction.

3-147 SEA WATER AND BRINE PIPING.

3-148 Repair and replacement of sea water and brine piping is the greatest, single maintenance item of the Freeport Plant. The major piping in this category are:

a. Suction piping at brine pumps P-11 through P-17;

b. Discharge piping at brine pumps P-14 through P-19;

c. Sea water feed piping from heat exchanger 301 to first effect evaporator; and

d. Sea water feed piping from heat exchanger 212 to deaerator.

3-149 All piping failures which required maintenance were located adjacent to weld seams or on the bottom of the pipe line.

3-150 Numerous piping modifications or additions have been incorporated into the Freeport Plant. These modifications or additions include:

a. Brine feed line from effect X to vapor body of effect XI; and

b. Cross-over piping at pit area to permit the utilization of pumps P-1 and P-2 simultaneously on a two-pass condenser operation.

3-151 In addition, two stress-relieved sections of pipe have been installed adjacent to the weld seams of the brine pump P-15 discharge. Numerous failures have occurred at this location in the past. Perhaps by stress-relieving, the corrosion rate can be significantly reduced.

3-152 Dissolved oxygen and high temperatures are the biggest factors that contribute to piping failures. Figures 3-22 and 3-23 provide a visual comparison of piping that perform similar service but carry different products. Figure 3-22 illustrates the nondeaerated sea water feed line piping while Figure 3-23 illustrates the deaerated sea water feed line piping. Note the buildup of corrosion products in the nondeaerated sea water line and the lack of corrosion product buildup in the deaerated sea water line. Figure 3-24 illustrates the sea water feed line piping to the deaerator. It also reflects the results of scaling and corrosion on piping that carries nondeaerated sea water.

3-153 STEAM AND VAPOR PIPING.

3-154 No significant maintenance is required by the steam and vapor piping. A new vapor line was installed from the vents of effects II and III to the deaerator. This line disposes the noncondensable vapors from the two effects.

3-155 CONDENSATE PIPING.

3-156 No significant maintenance is required by the condensate piping. The

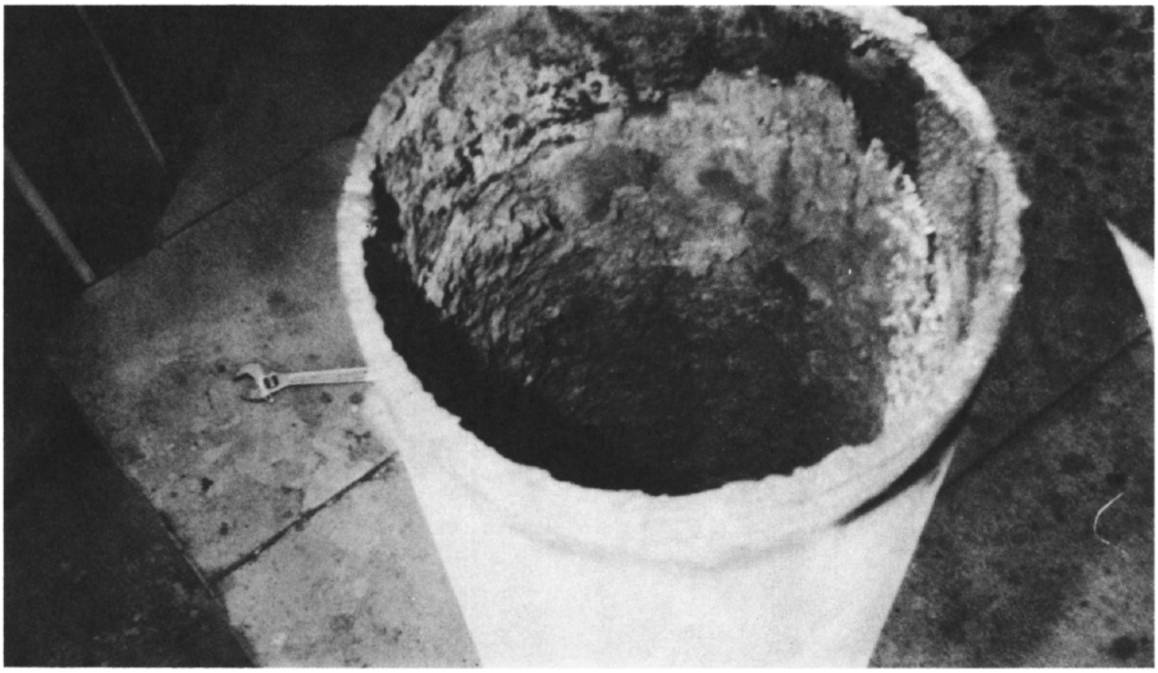


FIGURE 3-22. NONDEAERATED SEA WATER PIPING

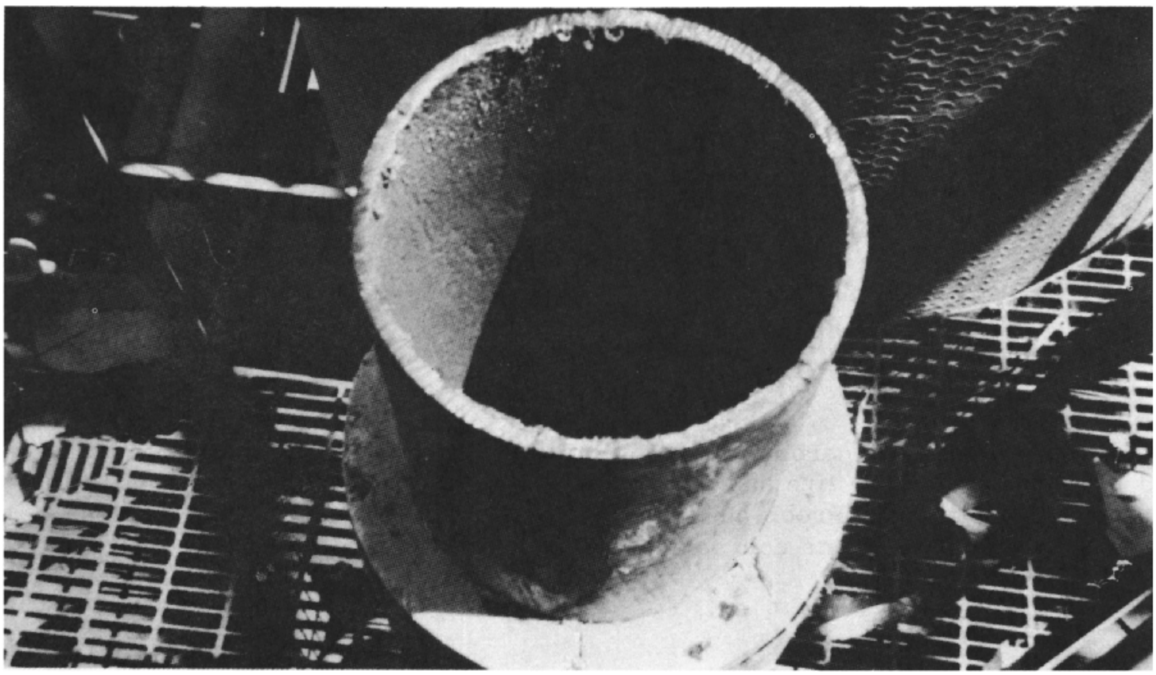


FIGURE 3-23. DEAERATED SEA WATER PIPING

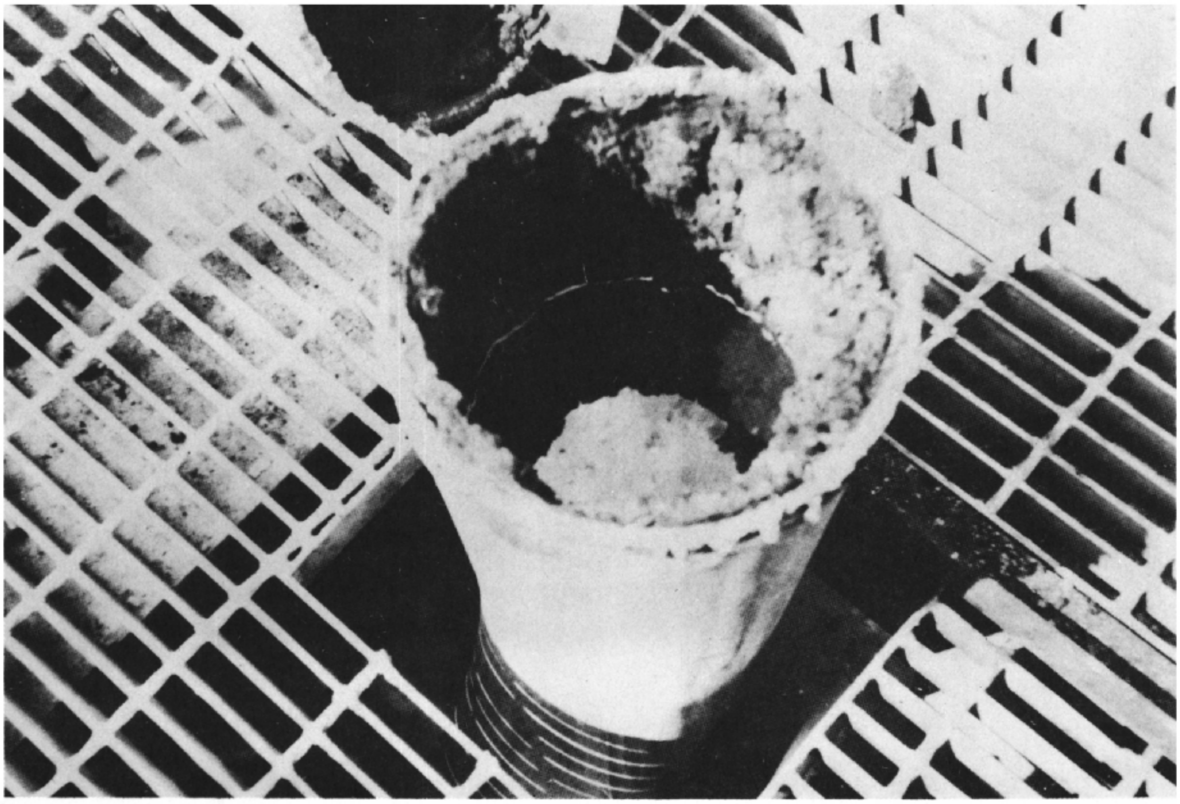


FIGURE 3-24. SEA WATER FEED PIPING

new condensate lines that been installed at the Freeport Plant include:

a. Line to bring fresh water to suction of pump P-3; and

b. Line from pumps P-42 and P-43 discharge to pump P-44 discharge.

3-157 ACID PIPING.

3-158 The one-inch, carbon steel line from the acid pump to the top of the sea water feed line has been replaced in its entirety twice during the current year. Problems with atmospheric moisture dilution caused the acid to become corrosive and resulted in the failure of this line.

IV ECONOMIC EVALUATION

4-1. GENERAL.

4-2. Much confusion can arise in the discussion of economic costs of any process, especially those of a developmental nature. Care must be taken that the bases upon which the costs are developed are thoroughly understood; this is particularly true with respect to the costs of salt water conversion. Because of the many unknown factors which existed at the inception of the Saline Water Conversion Program, the Demonstration Plants were equipped with experimental process equipment and unproven materials of construction. Therefore, it would be possible for this vast test program to unduly inflate the costs of desalination.

4-3. Every attempt has been made to consider all abnormal features of the Freeport Demonstration Plant in order to arrive at realistic cost figures. The Office of Saline Water Research and Development Progress Report Number 72 has been used as a guideline in developing costs for a "Normalized Plant". The basis utilized to develop each figure in the tables appearing in this Section is provided and the actual calculations are presented in the Appendix (Section VII).

4-4. PLANT OPERATING COSTS.

4-5. Table 4-1 presents daily average operating costs based on a 330-day year, and were derived from the costs incurred in May, 1964, which had 31 operating days. The Production and Cost Report for May 1964 is presented in Table 4-4.

4-6. Items 7, 9, 10 and 11 were derived in accordance with the procedures specified in Report No. 72.

4-7. General and administrative expense is one-three-hundred-and thirtieth of the actual G and A expense charged for the year. It includes such items as administrative salaries, travel expense, safety expense, communications, public relations and union materials, and Denver office expenses. Items not included are: experimental and research expenses; OSW supervision; and Management and Operations Contractor fixed fee and overhead percentage.

4-8. The average actual Plant cost computed on this basis is $1.23 per gallon, based on a gross production of 1,100,000 GPD.

DEMONSTRATION PLANT NO. 1

(BASIS: ONE STREAM DAY AND 330 OPERATING DAYS PER YEAR)

I. DIRECT OPERATING COSTS	$/Stream-Day	¢/1000 Gal.
1. Energy		
a. Electric Power ($0.008/KWH)	84.77	7.70
b. Steam ($0.45/1000-LBS)	358.02	32.54
2. Chemicals:		
Sulfuric Acid	39.86	3.62
Caustic Soda		
Soda Ash		
Calcium Chloride		
3. Operating Labor	123.27	11.20
4. Maintenance Labor	49.51	4.50
5. Supervision of Operation and Maintenance	33.23	3.02
6. Supplies and Maintenance Materials	24.86	2.26
Total Direct Costs	713.52	64.86
II. INDIRECT OPERATING COSTS		
7. Payroll Extras	47.17	4.28
8. General and Administrative Expenses	157.59	14.32
9. Depreciation and Interest (0.02242% of Total Capital Cost)	333.11	32.08
10. Taxes and Insurance	91.61	8.32
11. Interest on Working Capital	9.76	.88
Total Indirect Operating Costs	639.24	58.04
TOTAL OPERATING COSTS	1352.76	122.91

TABLE 4-1. AVERAGE DAILY OPERATING AND PRODUCTION COSTS

DEMONSTRATION PLANT NO. 1

I. PRINCIPAL ITEMS OF EQUIPMENT (PIE)	Dollars	% of PIE
1. Special Equipment (Shop and Field Assembled)		
a. Evaporator Effects	609,740	54.22
b. Deaerator Equipment	16,600	1.48
c. Vacuum Equipment	14,700	1.31
d. Heat Exchangers	243,645	21.67
e. Pumps and Drivers	65,545	5.83
f. Desuperheater and Seal Water Pump	31,400	2.79
g. Water Cooling Equipment	11,500	1.02
h. Air Compressor Equipment	7,000	0.62
i. Acid and Caustic System	5,630	0.50
j. Concrete Work	21,800	1.94
k. Miscellaneous	66,340	5.90
SUBTOTAL	1,093,900	97.28
2. Standard Engineering Equipment		
a. Spare Parts	21,200	1.89
b. Overhead Crane	9,360	0.83
SUBTOTAL	30,560	1.89
TOTAL PIE	1,124,460	100.00
II. PROCESS FACILITIES		
1. Raw Feed and Cooling Water Intake Facilities (Includes Pumps and Pipeline Only)	120,000	
2. Site Development	36,800	
3. Insulation	40,200	
4. Painting	18,700	
5. Electrical	70,275	
6. Piping	Included in Installation PIE	
7. Instruments	50,300	
8. Buildings	42,800	
9. Boiler Plant Facilities	138,700	
TOTAL PROCESS FACILITIES	517,775	

TABLE 4-2. CAPITAL COSTS FOR A "NORMALIZED PLANT." (Sheet 1 of 2)

TABLE 2 (CONT'D)

III. OTHER PLANT COSTS	Dollars
1. Engineering	120,000
2. Interest on Investment	32,845
3. Startup Expense	30,000
4. Cost of Site	20,000
TOTAL OTHER CAPITAL COSTS	202,845
TOTAL PLANT COSTS	1,845,080
Capital Cost Per Gallon of Daily Capacity	1.827

TABLE 4-2. CAPITAL COSTS FOR A "NORMALIZED.PLANT." (Sheet 2 of 2)

4-9. CAPITAL COSTS FOR A "NORMALIZED PLANT."

4-10. Table 4-2 presents the capital cost for a "normalized" Long-Tube-Vertical (LTV) Plant. The cost figures were derived from the OSW Report No. 72.[1]

4-11. The capital cost of heat exchangers is reduced by $20,455 to account for the removal of six-200-series heat exchangers.

4-12. The capital cost of pumps and drivers is reduced by $6,810 to account for six condensate pumps that are no longer in service.

4-13. Interest on the investment is computed at 2.0 percent of I and II, in Table 4-2.

4-14. OPERATING COSTS FOR A "NORMALIZED PLANT."

4-15. Table 4-3 presents the operating costs for a "normalized" LTV Plant. The cost figures for most of these data were derived from OSW Report No. 72.

4-16. The electric power cost is computed in accordance with the method cited in OSW Report No. 72.[2]

4-17. The fuel oil is computed on the basis that 0.872 pounds of saturated steam at 261 °F. is required for each gallon of product water. This basis has been established during periods of normal operation which was achieved during 7 of the 12 operating months covered by this report.

4-18. Sulfuric acid consumption was computed at 50 LB/HR of 98 percent acid.

4-19. Caustic soda consumption was computed on the basis of Plant ex-

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[1] Office of Saline Water, Research and Development Progress Report Number 72, page 31a.

[2] Office of Saline Water, Research and Development Progess Report Number 72, pages 43, 44, and 16.

perience and amounts to approximately 124 LB/DAY on an anhydrous basis.

4-20. Soda Ash and Calcium Chloride consumption is also computed on the basis of Plant experience and amounts to 50 LB/DAY each.

4-21. Operating labor is computed on the basis of two men per eight-hour shift. A premium rate of three percent for afternoon shift and five percent for graveyard shift is included ($3.17/HR).

4-22. Maintenance labor is calculated on the basis of 5 maintenance men working 5 days a week with an allowance for one relief operating shift per week.

4-23. Supervision of operation and maintenance is figured on the basis of one supervisor devoting half-time or $4800 per year.

4-24. Supplies and maintenance material costs are 0.5 percent of capital investment excluding startup costs.

4-25. Payroll extras are 16.0 percent of operating and maintenance labor.

4-26. General and Administrative costs are estimated at 25.0 percent of operating and maintenance labor and payroll extras.

4-27. Depreciation and interest is based on a 20-year plant life and 7.4 percent per year rate (or 0.02242 percent per day) of the total capital cost less the value of the land.

4-28. Taxes and insurance costs are based on 2.0 percent of the capital cost less the startup expenses.

4-29. Interest on working capital is determined by totaling items 1 through 10 (Table 4-3) and multiplying this total by 60 days/330 days and then computing 4.0 percent of the product.

4-30. The cost per 1000 gallons is based on a production of 1,010,000 gallons per day, net (1,100,000 gallons per day, gross). It has been demonstrated that a production rate of approximately 1,200,000 gallons per day (gross) can be achieved; however, pumping equipment limitations presently preclude the possibility of operating continuously at this rate. It is reasonable to assume the experienced

DEMONSTRATION PLANT NO. 1

(BASIS: ONE STREAM DAY AND 330 OPERATING DAYS PER YEAR)

	$/Stream-Day	¢/1000 GAL
I. DIRECT OPERATING COSTS		
1. Energy		
a. Electric Power ($0.0085 KWH)	76.73	7.60
b. Fuel (Oil, Gas)	446.78	44.24
2. Chemicals		
a. Sulfuric Acid 98% ($0.012/LB)	14.40	1.43
b. Caustic Soda ($0.03/LB-Anhyd)	3.72	0.37
c. Soda Ash ($0.11/LB)	5.50	0.54
d. Calcium Chloride (0.13/LB)	6.50	0.64
3. Operating Labor (3.17 HR)	152.16	15.07
4. Maintenance Labor (3.17/HR)	33.78	3.34
5. Supervision of Operation and Maintenance	13.54	1.34
6. Supplies and Maintenance Materials	27.50	2.72
TOTAL DIRECT COSTS	780.61	77.29
II. INDIRECT OPERATING COSTS		
7. Payroll Extras	29.75	2.94
8. General and Administrative Expenses	53.92	5.34
9. Depreciation and Interest (0.02242% of Total Capital Cost)	409.18	40.50
10. Taxes and Insurance	110.00	10.89
11. Interest on Working Capital	9.55	0.95
TOTAL INDIRECT OPERATING COSTS	612.40	60.63
TOTAL OPERATING COSTS	1,393.01	137.92

TABLE 4-3. OPERATING COSTS FOR A "NORMALIZED PLANT."

rate of 1,010,000 gallons per day (net) for Plant comparison purposes.

4-31. PRODUCTION AND COST REPORT FOR TYPICAL MONTH.

4-32. A typical Monthly Production and Cost Report of the Freeport Plant operation is provided in Table 4-4. Of the twelve monthly reports that were submitted during the period covered by this Annual Report, the May 1964 Report has been selected as typical, because in this time period abnormal operating and maintenance problems were infrequent.

4-33. MONTHLY PRODUCTION AND COST SUMMARY.

4-34. Table 4-5 presents pertinent production and cost data that have been compiled on a monthly basis for the complete operating period covered by this Annual Report. The Table provides sufficient information to evaluate and summarize the Plant's annual operation.

4-35. Figure 4-1 graphically provides essentially the same information for a visual, economic comparison of the Plant's monthly production and cost figures.

4-36. SIX-MONTH PROGRESSIVE AVERAGE COST.

4-37. Table 4-6 presents the actual and adjusted, six-month progressive average costs per 1000 gallons of water produced. The adjusted figures for the period between October 1962 and June 1963 are presented to compensate for the numerous outages necessitated by Development Programs.

4-38 Figure 4-2 graphically provides the same information for a visual appraisal of the six-month, direct and indirect, progressive average costs.

4-39. MAINTENANCE COST SUMMARY.

4-40. Maintenance costs incurred during the fiscal year 1964 are presented in Table 4-7. The various account numbers listed in the column headings are identical to those used in the monthly reports.

4-41. Total maintenance cost was $55,109.35 or $4,592.26 per month. These amounts would be excessive under normal operating conditions, but in this instance can be justified because of the several experi-

DEMONSTRATION PLANT NO. 1

PRODUCTION AND COST REPORT FOR MAY 1964

GROSS PRODUCTION: 32,331,500 gallons				
		CURRENT MONTH	CENTS PER 1000 GAL	FY TO DATE
DIRECT COSTS				
Operation:				
500	Supervision & Engineering	604.16	1.87	6086.38
501	Labor	3453.25	10.69	36285.84
502	Materials & Supplies			
	502.1 Process Chemicals	619.40	1.91	8149.69
	502.2 Laboratory Supplies	134.89	.42	1343.26
	502.3 Product Treating	414.50	1.28	414.50
503	Utilities			
	503.1 Steam	10499.22	32.47	87925.33
	503.2 Electricity	2486.40	7.69	20929.60
	TOTAL OPERATING COSTS	$ 18211.82	56.33	161134.60
Maintenance:				
510	Supervision & Engineering	326.78	1.01	4020.82
511	Labor			
	511.1 Process Plant	994.73	3.08	17297.39
	511.2 Other Plant	320.62	.99	5031.66
	511.3 Other Than Plant	70.86	.22	70.86
512	Materials & Supplies			
	512.1 Direct-Process Plant	231.84	.72	14735.92
	512.2 Direct-Other Plant	44.83	.14	3149.22
	512.3 Indirect	371.85	1.15	2571.38
513	Other Maintenance Expense	57.54	.17	1853.55
	TOTAL MAINTENANCE COSTS	$ 2419.05	7.48	48730.80
Extraordinary Expense:				
531	Labor			77.23
Experimental & Research:				
540	Supervision & Engineering	787.68	2.44	4637.84
541	Labor	325.74	1.01	726.93
542	Material	99.31	.30	1013.33
	TOTAL EXTRAORDINARY/ EXPERIMENTAL & RESEARCH	$ 1212.73	3.75	6455.33
	TOTAL DIRECT COSTS	$ 21843.60	67.56	216320.73

TABLE 4-4. PRODUCTION AND COST REPORT FOR TYPICAL MONTH (Sheet 1 of 2)

DEMONSTRATION PLANT NO. 1

INDIRECT COSTS:	CURRENT MONTH	CENTS PER 1000 GAL.	FY TO DATE
General & Administrative:			
Labor			
920 Administrative Salaries	830.95	1.18	3731.91
921 Payroll Burden	1317.35	4.07	13293.20
923 Reports and Procedures	517.96	1.60	3660.45
924 Travel			767.27
925 Safety	38.76	.12	396.34
926 Public Relations & Union			1554.55
Material & Expense			
930 Office Supplies	70.88	.22	941.71
931 Safety Supplies	13.50	.04	257.69
932 Communications	235.73	.73	2438.21
933 Public Relations & Union	64.80	.20	111.88
935 Plant Transportation	26.91	.08	363.02
936 Freight-unallocated			794.15
937 Miscellaneous G & A Expense			190.74
Other			
940 Denver Office Expense	2614.20	8.09	20510.33
950 Contractor Fixed Fee	1489.00	4.61	16379.00
951 Contractor Overhead Percentage	1536.47	4.75	13920.68
TOTAL GENERAL & ADMINISTRATIVE COSTS	$ 8756.51	25.69	79311.13
Depreciation			
980 Unallocated	$ 5861.74	18.13	63708.99
TOTAL INDIRECT COSTS	$14618.15	45.21	143020.12
TOTAL COSTS	$36461.85	112.77	359340.85

TABLE 4-4. PRODUCTION AND COST REPORT FOR TYPICAL MONTH (Sheet 2 of 2)

DEMONSTRATION PLANT NO. 1

MONTH	DOWN TIME HOURS	PER CENT ON-STREAM TIME	TOTAL PRODUCTION GALLONS	AVERAGE PRODUCTION GALS/STREAM DAY	THERMAL EFFICIENCY LBS-WATER / LB-STEAM	TOTAL UNIT COST DOLLARS/ 1000 GAL	DIRECT UNIT COST DOLLARS/ 1000 GAL
July	136	81.7	26,085,250	1,029,680	11.69	1.2794	0.8305
Aug.	0	100.0	31,650,830	1,020,995	11.69	1.0574	0.6476
Sept.	231	68.0	20,646,940	1,013,347	11.76	1.4503	0.8043
Oct.	396	46.8	14,953,740	1,031,292	11.41	2.2833	1.4138
Nov.	62	91.4	28,777,990	1,049,650	11.67	1.1638	0.7156
Dec.	0	100.0	31,638,850	1,020,608	11.74	1.0915	0.6881
Jan.	348	53.0	16,608,280	1,016,800	11.90	1.7072	0.9701
Feb.	144	79.3	23,282,970	1,012,300	10.92	1.3634	0.7866
Mar.	213	71.0	23,199,260	1,048,550	11.39	1.4249	0.8620
Apr.	196	72.8	23,092,150	1,057,670	11.99	1.3407	0.7614
May	12	98.4	32,331,500	1,060,000	11.54	1.1138	0.6756
June	252	65.0	17,539,500	899,450	11.24	1.9062	1.0885
TOTAL	1990	77.3	289,807,260	1,027,685	11.58	1.3553	0.8123

TABLE 4-5. MONTHLY PRODUCTION AND COST SUMMARY

DEMONSTRATION PLANT NO. 1

	TOTAL COST/1000 GAL. IN DOLLARS		DIRECT COSTS/1000 GAL.	
Mar.-July 62	1.29		1.03	
Mar.-Aug.	1.26		0.81	
Apr.-Sept.	1.18		0.86	
May-Oct.	1.29		0.89	
June-Nov.	1.29		0.84	
July-Dec.	1.29		0.80	
Aug.-Jan 63	1.34		0.81	
Sept.-Feb.	1.41		0.84	
Oct.-Mar.	1.41	1.34*	0.85	0.82*
Nov.-Apr.	1.31	1.25*	0.78	0.75*
Dec.-May	1.29	1.24*	0.77	0.74*
Jan.-June	1.42	1.24*	0.83	0.74*

* SIX-MONTH PROGRESSIVE AVERAGE ADJUSTED TO COMPENSATE FOR OUTAGES NECESSITATED BY INCREASED DEVELOPMENT PROGRAM ACTIVITY

TABLE 4-6. SIX-MONTH PROGRESSIVE AVERAGE COSTS

MONTH	ENGRG. & SUPV. (Acct. 511)	LABOR (Acct. 511)			MATERIALS (Acct. 512)				Acct. 513	TOTAL COST
		PROCESS PLANT 511.1	OTHER PLANT 511.2 & 511.3	TOTAL Acct. 511	512.1 DIRECT PROCESS PLANT	512.2 DIRECT OTHER PLANT	512.3 IN-DIRECT	TOTAL Acct. 512	OTHER MAINT. EXPENSE	
July '63	$ 553.28	1654.48	401.93	2056.41	2097.71	32.81	286.56	2417.08	153.18	5179.95
Aug	288.64	124.61	436.94	561.55	108.09	104.27	140.98	353.34	-	1203.53
Sept	286.36	2112.62	515.63	2628.25	823.26	41.98	247.27	1112.51	-	4027.12
Oct	525.00	3968.77	830.03	4798.80	4958.99	87.25	597.48	5643.72	97.06	11064.58
Nov	171.82	720.68	557.36	1278.04	674.96	312.50	213.94	1201.41	216.75	2868.02
Dec	85.91	725.77	319.06	1044.83	131.48	86.58	225.85	443.91	705.74	2280.39
Jan '64	568.12	1735.04	340.02	2075.06	1553.12	525.10	463.39	2541.61	7.50	5192.29
Feb	430.28	1299.13	563.85	1862.98	929.69	507.35	375.09	1812.13	168.96	4274.35
Mar	464.17	2092.15	354.38	2446.53	2654.79	73.66	374.63	3103.08	141.58	6155.36
Apr	320.46	1869.41	391.84	2261.25	575.81	37.28	546.62	1159.71	119.14	3860.56
May	326.78	994.73	391.48	1386.21	231.84	44.83	371.85	648.52	57.54	2419.05
June	269.90	2771.51	965.17	3736.68	589.54	87.26	887.25	1564.05	1013.52	6584.15
Totals	$4290.72	20,068.90	6067.69	26,136.59	15,329.28	1940.87	4730.92	22,001.07	2680.97	55,109.35
Monthly Aug.	$ 357.54	1672.34	505.62	2177.96	1277.39	161.73	364.23	1833.35	223.41	4592.26
% of Total Maint.	$ 7.79	36.42	11.01	47.43	27.82	3.52	8.58	39.92	4.86	100.00

TABLE 4-7. MAINTENANCE COST SUMMARY

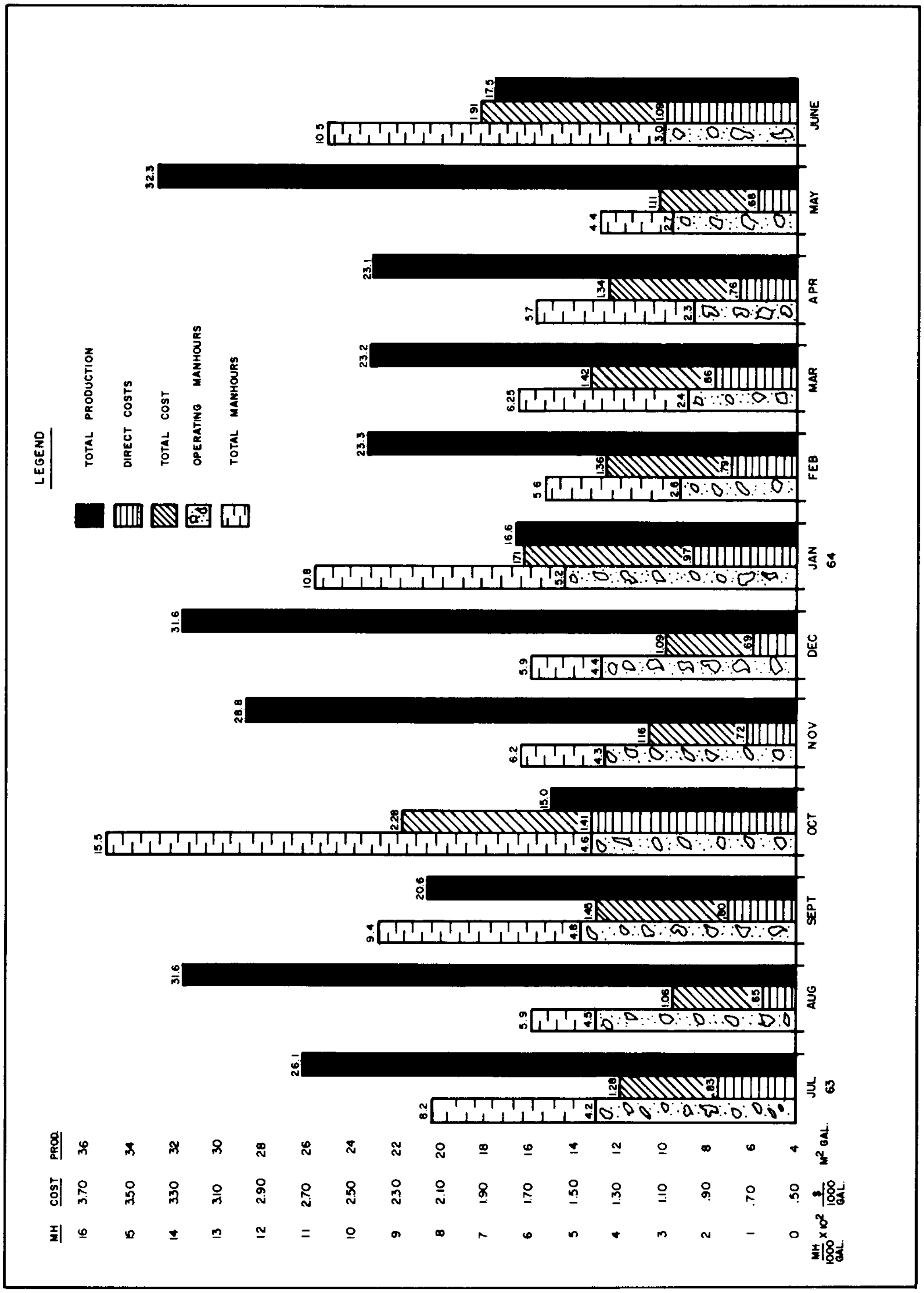

FIGURE 4-1 GRAPHIC MONTHLY PRODUCTION AND COST SUMMARY

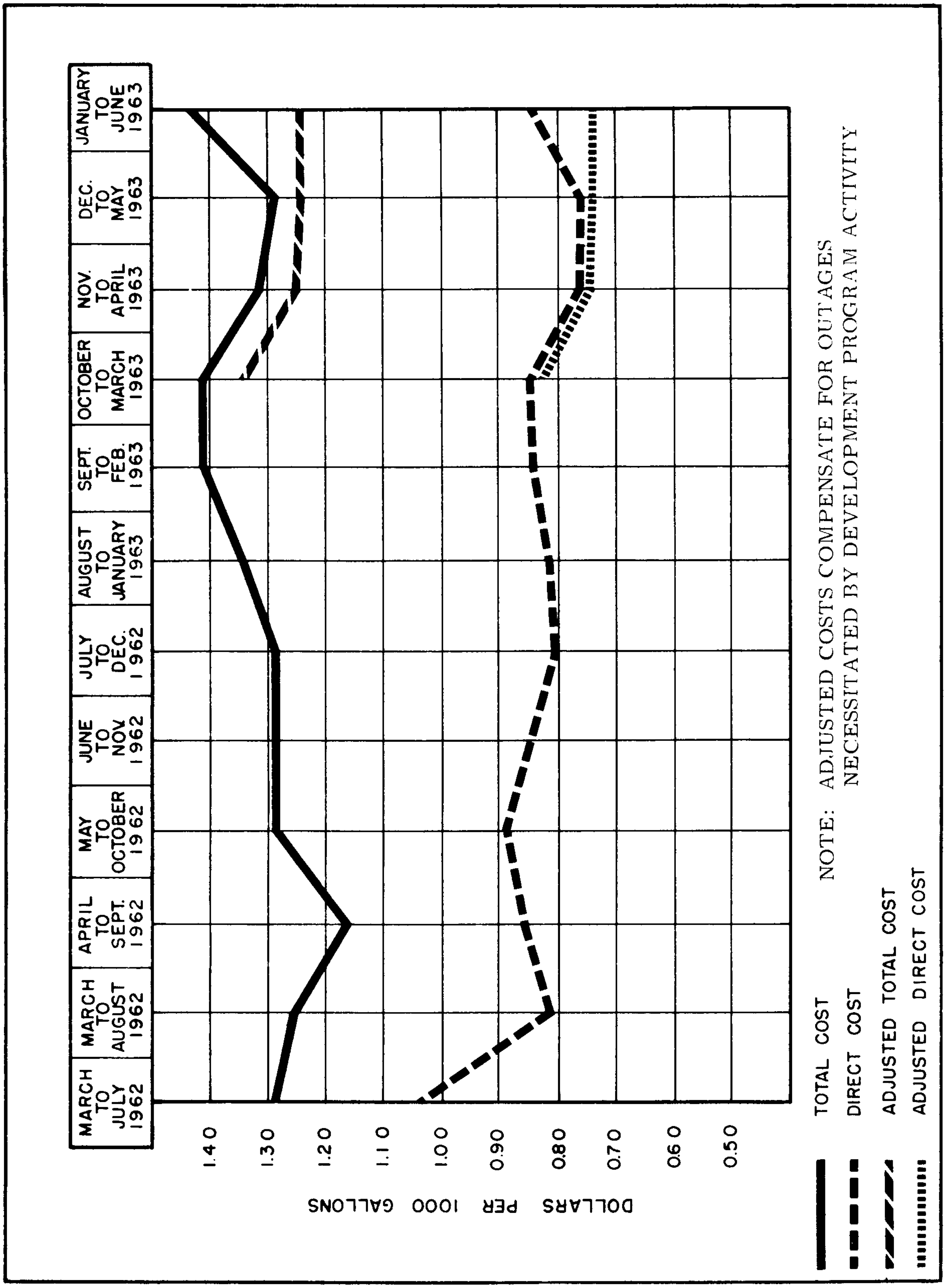

FIGURE 4-2. GRAPHIC SIX-MONTH PROGRESSIVE AVERAGE COSTS

mental features which require abnormal maintenance. An equally importanat factor in maintenance costs are the requirements imposed by the development program. It is anticipated that maintenance costs will decrease in fiscal year 1965, but will continue to remain higher than normal because of the frequent scheduled shutdowns for the development program. During these shutdowns, all possible precautions are taken to ensure that the ensuing run(s) will be trouble-free; resultantly, manhours expended and equipment repaired or replaced are greatly in excess of those for normal Plant operation.

4-42. Maintenance costs for individual items of equipment are presented in Table 4-8. It should be noted that the figures in this Table are influenced by the following factors: "Engineering and Supervisory" costs (Account No. 510) are included; "Indirect Material" costs (Account No. 512.3) are not included; and some, but not all of "Other Maintenance Expenses" (Account No. 513) are included.

EQUIPMENT	TOTAL MANHOURS	TOTAL LABOR (DOLLARS)	TOTAL MATERIALS
INSTRUMENTATION:			
pH Recorder	25	95.98	-
Conductivity Cell and Recorder	79	303.31	11.59
Apparent HT Transfer Coefficient Meter	12	46.07	-
Level Transmitters, Indicators, and Controllers	71	272.60	92.27
Temperature Elements and Indicators	59	226.52	561.15
Flow Transmitters, Recorders, and Controllers	219	840.83	182.58
Condensate Purity Meter	1	3.83	-
Pressure Gages	39	149.74	131.14
Annunciator and Alarms	14	53.75	27.74
Bailey Meter	8	30.71	-
Compax Level Controls	276	1059.67	434.89
Misc. Instrumentation	77	295.63	312.68
TOTAL INSTRUMENTATION	880	3378.64	1754.04
EVAPORATORS			
Effect I	245	940.65	691.66
Effect II	84	322.50	233.10
Effect III	47	180.45	64.08
Effect IV	75	287.95	74.68
Effect V	68	261.07	62.40
Effect VI	55	211.17	35.45
Effect VII	63	241.88	36.33
Effect VIII	47	180.45	32.14
Effect IX	54	207.33	96.39
Effect X	137	525.99	512.79
Effect XI	278	1067.35	444.64
Effect XII	257	986.72	1371.89
TOTAL EVAPORATORS	1410	5413.51	3655.55
HEAT EXCHANGERS:			
201	21	80.62	20.42
202	4	15.35	-
203	4	15.35	-
204	5	19.20	-
205	7	26.87	-

TABLE 4-8. INDIVIDUAL EQUIPMENT ITEMS MAINTENANCE COST SUMMARY (SHEET 1 OF 4)

EQUIPMENT	TOTAL MANHOURS	TOTAL LABOR (DOLLARS)	TOTAL MATERIALS
206	5	19.20	-
207	5	19.20	-
208	-	-	-
209	11	42.23	-
210	-	-	-
211	-	-	-
212	28	107.50	181.68
213	40	513.57	246.44
214	50	101.97	246.44
215	59	226.52	242.75
301	31	119.02	-
302	53	203.49	100.75
303	2	7.68	-
304	24	92.14	-
305	2	7.68	-
306	5	19.20	-
307	10	38.39	-
308	-	-	-
309	-	-	-
310	1	3.84	26.50
311	56	215.00	256.75
312	46	176.61	115.30
TOTAL HEAT EXCHANGERS	469	1800.63	1437.03
OTHER VESSELS & TANKS:			
Deaerator	215	825.47	224.07
Product Storage Tank	132	506.80	123.95
Caustic Storage Tank	8	30.71	22.80
Sulphuric Acid Storage Tank	12	46.07	-
Sulfite Storage	5	19.20	12.50
TOTAL OTHER VESSELS	372	1428.25	383.32
MISC. EQUIPMENT:			
Clarifier-thickener	-	-	-
Steam Jet Ejectors	-	-	-
Traveling Screen	24	92.14	-
Sea Water Intake Pit	115	441.53	1249.56

TABLE 4-8. INDIVIDUAL EQUIPMENT ITEMS MAINTENANCE COST SUMMARY (SHEET 2 OF 4)

EQUIPMENT	TOTAL MANHOURS	TOTAL LABOR (DOLLARS)	TOTAL MATERIALS
MISC. EQUIPMENT (cont'd)			
Overhead Crane	44	168.93	10.00
TOTAL MISC. EQUIPMENT	183	702.60	1259.56
PUMPS:			
P-1	18	69.10	525.27
P-2	48	184.29	221.98
P-2a	57	218.85	15.80
P-3	31	119.02	-
P-4	13	49.90	-
P-5	15	57.59	10.04
P-5a	38	145.90	33.79
P-6	13	49.91	-
P-11	4	15.35	9.10
P-12	25	95.98	9.10
P-13	18	69.10	67.34
P-14	25	95.98	9.10
P-15	9	34.55	60.56
P-16	8	30.71	9.10
P-17	36	138.22	35.81
P-18	17	65.27	16.49
P-19	17	65.27	132.87
P-20	14	53.75	122.01
P-21	33	126.70	122.01
P-22	27	103.66	135.75
P-23	43	165.09	166.57
P-31	9	34.55	-
P-32	1	3.84	-
P-33	2	7.68	-
P-34	2	7.68	-
P-35	4	15.35	-
P-36	2	7.68	-
P-37	2	7.68	-
P-38	3	11.50	8.01
P-39	4	15.35	8.01
P-40	4	15.35	60.27
P-41	12	46.07	8.01

TABLE 4-8. INDIVIDUAL EQUIPMENT ITEMS MAINTENANCE COST SUMMARY (SHEET 3 OF 4)

EQUIPMENT	TOTAL MANHOURS	TOTAL LABOR (DOLLARS)	TOTAL MATERIALS
P-42	10	38.39	10.25
P-43	3	11.50	-
P-44	9	34.55	-
P-45	-	-	-
P-46	93	357.06	133.36
P-50	7	26.87	5.50
P-51	8	30.71	-
P-53	1	3.84	-
P-54	17	65.26	-
P-60	72	276.44	245.71
P-61	22	84.47	155.49
P-62	72	276.44	334.58
TOTAL PUMPS	868	3332.45	2729.89
ELECTRICAL:			
Switchgear and Starters	2	7.67	-
Electric Motors	-	-	675.00
Lighting (other than building)	-	-	69.69
Heat Tracing and Freeze Protection	113	433.85	136.48
Misc. Electrical Equipment	25	95.98	96.78
TOTAL ELECTRICAL	140	537.50	977.95
PIPING:			
Feed and Brine	755	2898.75	3140.54
Steam and Vapor	106	406.98	681.27
Condensate	198	760.20	1500.13
Gland Seals and Drains	-	-	-
Slurry	-	-	-
Control Valves	78	299.47	-
Chemical Piping	124	476.09	836.00
Compressed Air System	15	57.59	41.91
Misc. Piping	-	-	-
TOTAL PIPING	1276	4899.08	6199.85
GENERAL MAINTENANCE:			
Plant	1470	5643.91	57.13
Building	836	3209.74	107.62
TOTAL GENERAL	2306	8853.65	164.75
GRAND TOTALS	7904	30346.31	18561.94

TABLE 4-8. INDIVIDUAL EQUIPMENT ITEMS MAINTENANCE COST SUMMARY (SHEET 4 OF 4)

V. TECHNICAL EVALUATION

5-1 PROCESS ANALYSIS.

5-2 An analysis of the Long-Tube, Vertical (LTV) evaporation process is predicated upon the results of operating experience acquired during the last three years of operation and the various Development Programs that were applied in conjunction with the Plant operation. The process flow diagram (see Figure 7-1) provides a graphic description of the LTV process.

5-3 DEVELOPMENT PROGRAM.

5-4 The operation of Demonstration Plant No. 1 from April 1961 to approximately October of 1963 produced extensive and valuable operational and maintenance experience. Many of the mechanical problems which previously plagued the continuous operation of the Plant were resolved, at least to the extent that the XI and XII effects' scale formation became the limiting factor to any production run. Notable achievements during this period were:

a. Identification of realistic operating limits as originally conceived.

b. Successful, if not economical and completely understood, operation of the pH method of decarbonation and vacuum deaeration.

c. Elimination of scale formation in effect I.

d. Installation of tubing material which not only resists corrosion but also improves heat transfer.

e. Incorporation of recycling systems for the low-temperature effects resulting in better heat transfer and considerable reduction in scale formation rate.

f. Elimination of six condensate-to-feed exchangers with negligible loss in economy (2500 SQ-FT of heat transfer surface).

g. Many changes to redistribute pumping capacity and improve the reliability of pumping equipment.

5-5 The P-5 pump replacement had a very profound effect on the reliable operation of both the deaerator and the evaporator; and has contributed materially to the capability of studying other improvements to the process. In addition, many other problems were considered, including the never-ending, corrosion problem. The programs of prior years are adequately discussed in First and Second Annual Reports, and the reader is referred to these reports if more detail is desired.

5-6 Studies related to scaling, and the prediction and quantitative evaluation of scaling from on-stream data, were pursued diligently throughout the year. These studies have resulted in the development of excellent indications of the extent of scaling. Equipment performance and efficiency tests were continued, but were obviously of limited value until positive identification of cause and effect could be determined.

5-7 In December of 1963 and January of 1964, Stearns-Roger undertook an evaluation of the existing investigative program, the process, and the equipment. This evaluation resulted in the submission by Stearns-Roger, and subsequent approval by OSW, of a long-range development program. The long-range program was necessary to develop and improve the design, performance, and costs of this promising LTV process plant. The achievements of past operation have improved control and reliability to the point where sufficient variables can be held constant over a suitable period of time, and thereby, enable a positive identification of cause and effect. This program simulates a series of test runs of equal length under predetermined conditions. During each run, the variation of controllable operating conditions is limited to non-interacting ones (insofar as is practical).

5-8 The program has progressed approximately on schedule, with promising results from the first four runs. Many advancements have been made in many of the specified areas of emphasis, and the attack on all problem areas continues. The major problems of operation result from the fouling of equipment, attributable either to the formation of scale or the accumulation of silt. Corrosion and erosion problems are present and may become serious in the future. The cost of producing water has been the only criteria of process evaluation and it still remains the final standard. Full realization of the potential thermal advantages of the LTV will significantly reduce the cost of water produced by this process. It is necessary, therefore, to pursue the problems which limit operation, in a manner which does not yield the lowest immediate costs. This has been the type of operation carried out through the last six months of fiscal 1964. As the operating limitations are decreased or overcome completely, it will be possible to specify the operation of this Plant at significantly reduced costs. It will also be possible to design completely new and large-scale facilities with confidence.

5-9 Product cost is a function of energy requirements, equipment cost, operating labor, maintenance expense and treating chemical requirements. All of these items are affected by the variables of concentration ratio, deaerator operation, silt accumulation, over-all plant temperature difference and the reliable (continuous) operation of essential equipment. Corrosivity of very pure water is a problem to all evaporation-type desalinization plants. Improvements in any one of these areas permit improvements in others. The development program is concerned with the testing of operating methods, equipment changes, process changes, and the correlation of performance data with emphasis in the following phases:

a. Deaeration and Decarbonation

b. Feed Pretreatment

c. Feed Preheat and Product Condensation

d. Concentration Ratio (Extraction Ratio)

e. Brine Distribution and Scale Formation Control

f. Product Water Conditioning

g. Desuperheating Supply Steam

h. First Effect Operating Temperature

i. Materials of Construction (Corrosion and Erosion)

5-10 A formal report, separate from the normal monthly operating report, is prepared after each run at the Plant. These Development Run Reports are presented in a standard and established format to facilitate their evaluation.

5-11 Deaeration and Decarbonation.

5-12 The following investigations, equipment and process modifications have been accomplished or scheduled:

a. Determine and record the actual physical configuration of all components of the system, including the spray pattern.

b. Determine the flow rate, temperature and composition of all streams; and record liquid level, total and partial pressures, and pressure differentials for the various modes of operation. Heat and material balances, and an estimate of the chemical reactions occuring, will be developed from these data taken on all runs.

c. Runs have been accomplished with only the XI effect steam or II and III effect steam, and also with limited and controlled quantities of steam.

d. One or more runs are "scheduled" to determine the acidulation possibilities of hydrochloric acid.

5-13 <u>Feed Pretreatment.</u>

5-14 There are methods of scale prevention utilized in other processes or that have been proposed for this Plant, which should be investigated. Furthermore, the existing thickener tank can serve as a settling basin for the removal of silt and other suspended matter. The present development program or considerations include:

a. The rate at which silt removal can be achieved in the thickener without coagulation; and with coagulation.

b. The addition of polyphosphates, and possibly some organic chelating agents.

c. A circulation and recovery of an ion exchange resin which would accumulate troublesome ions in normal sea water and release them in strong brine.

5-15 <u>Feed Preheat and Product Condensation.</u>

5-16 Improvements in final condenser vacuum would result in a lower XII effect temperature. These improvements are possible with some loss of preheat on the incoming feed. The following development runs have been completed and will be repeated:

a. Apply the coldest sea water possible to the final product condenser and the barometric condenser. This action creates a maximum vacuum in the XII effect and contributes to a carryover problem which is being evaluated. This action also lowers the XII effect temperature, thus providing a net increase in the plant Δ T without raising the I effect temperature. Particular emphasis is placed in obtaining data that will provide the most accurate energy balance possible.

b. If results of a. above so indicate, one of the deleted 200-series heat exchangers could be connected to provide additional feed preheat from one or more of the streams leaving the Plant. This modification would replace heat sometimes recovered from warm condenser water. To date it is not indicated that the modification is necessary.

c. Data to evaluate the effect of recirculation at the sea water intake have been developed.

5-17 <u>Concentration Ratio (Extraction Ratio).</u>

5-18 A series of development runs at concentration factors from 2.5 to 3.5 are in progress. During these runs, emphasis has been placed on maximum production as limited by the capability of equipment with the final concentration factor held constant. Inlet concentration factor is adjusted to 1.0 by injection of strong brine. Each run is evaluated for:

a. Scale formation

b. Changes in Δ T

c. Heat exchange coefficients

d. Utilization of heat exchange surface

e. Energy utilization

f. Maintenance

5-19 Brine Distribution and Scale Formation Control.

5-20 Several runs have been accomplished to evaluate the performance of the notched weirs. Modifications to the distributor plates and inlet nozzles are planned. After each run, which is performed at essentially the same conditions, the loss of heating surface due to scale formation is evaluated. Emphasis is placed on the rate and pattern of scale formation, the effect of circulation rate, and the correlation of scaling to brine and vapor flow in the tubes. Flow rates appear to be exerting a primary influence on scale formation and a secondary influence on heat transfer; and therefore, a run will be performed with a predetermined quantity of additional plugged tubes in the effects being studied.

5-21 Product Water Conditioning.

5-22 Installation of a single, removal-cartridge type filter in the line to the City of Freeport has been accomplished. In addition, equipment for feeding water stabilization chemicals to the Freeport Product have been installed and connected. There is now a two-solution feed which results in 8 to 10 PPM of $CaC0_3$ in the product water. Inspections of the filter are made frequently. Bench scale work on mixtures of this water with Freeport well water will be accomplished to determine corrosivity and end composition. Corrosion coupons were inserted by NALCO and Stearns-Roger monitored the results of these tests. A full-scale, carefully controlled program to investigate six conditioning treatments has been programmed and is being implemented.

5-23 Desuperheating of Supply Steam.

5-24 In order to reliably control and evaluate the performance of effect I, it is necessary to have a saturated steam supply. The present desuperheater is designed to saturate the steam downstream of the pressure reducing station, but the throttling effect of the flow control valve causes it to return to the superheated state just prior to entry into the steam chest. Minor modifications have provided a method of correctly evaluating the degree of superheating. A further modification, to inject first effect condensate downstream from the flow controller, is improving saturation in the first effect chest.

5-25 First Effect Operating Temperature.

5-26 The improvements developed from prior test runs will be incorporated to the maximum extent before making runs at higher first effect temperatures. In addition, improvements in pumping capacity will be accomplished to provide increased feed capability. Higher operating temperature capabilities of this Plant will be established by the following development runs:

a. First effect steam temperature will be raised to 260°F and sea water feed held at 510,000 pounds-per-hour. Inlet concentration factor will be held to 1.0, but final concentration factor will be allowed to seek its own level.

b. First effect steam temperature will be held at 260°F but sea water feed will be varied to produce the optimum final concentration factor previously determined, except as limited by equipment capacity.

c. a. and b. above will be repeated in runs of 10°F increment temperature rises to a maximum of 300°F steam to the first effect steam chest.

5-27 Emphasis will be placed on obtaining data and comparing their results for the same factors listed in paragraph 5-17, Concentration Ratio.

5-28 Materials of Construction.

5-29 In addition to tests currently being conducted by alloy manufacturers, Stearns-Roger plans to insert metal test coupons and dissimilar metal assemblies in various portions of the Plant. Evaluation of plastic coatings is now in progress. Evaluation of the results is conducted in accordance with methods developed by NACE, so that they are comparable and consistent with existing practices.

5-30 THERMAL PERFORMANCE-EVAPORATOR

5-31 LTV Process.

5-32 The multiple effect evaporator is a process unit with wide application. Many variations are utilized. The one selected for Demonstration Plant No. 1 is well chosen for performance and potential. The economical use of heat transfer surface is an inherent advantage of the thin film evaporator. The falling film has the ability to keep the surface wetted and eliminates the detrimental effect of hydrostatic head on the flashing brine in other types of evaporation. The reverse distribution of concentration with respect to temperature is proper for the inverse solubility of the scaling components. The economy ratio is roughly proportional to the number of effects. The recovery of heat from condensate and bleed steam is sufficient to provide hot feed at or very near the boiling temperature in the first effect, and eliminates the necessity for a brine heater. This preheating feature is essentially automatic in this Plant through the use of bleed steam heaters at each effect. The use of condensate to feed exchangers versus flash tanks is optional, and both methods are utilized in this Plant. Concentration of the feed can be maintained at a sufficiently high rate to eliminate the need for inefficient mixing of strong and weak saline water for recycle.

5-33 Control of many conditions affecting flow over the heat transfer surface is inherently possible where interstage pumps are utilized to transfer brine from one effect to the next. No problem exists with bypassing steam; and economical production rates are possible over a wide range. Therefore, this process becomes a valuable tool in the design of dual purpose plants that are proposing to vary water and electrical output with demand. Valuable data relative to pressure drop in the two-phase tube flow is being obtained, which will provide an improved method of utilizing heat transfer surface through controlled recycling and bypassing of vapors on large units. The ability to condense final effect vapors at the approximate temperature of available heat sink, qualifies this process as one which can utilize low-grade heat. In addition, this process is easily vented and few problems result from the accumulation of noncondensable gases. The equipment is normally constructed in such a manner that air in leakage is almost nonexistent. A mechanical vacuum pump, in conjunction with a barometric condenser using raw sea water, is satisfactory for removing all noncondensables; and there is therefore, no requirement for high pressure steam.

5-34 An examination of the Plant Flow Diagram (see Figure 7-1), and a perusal of the Functional Description provided in the Second Annual Report, will provide a fairly complete concept of the process. For up-to-date temperatures, pressures, and flows, the Heat and Material Balance is provided (see Figure 5-1). The revisions incorporated into the Plant Flow Diagram as compared to the Second Annual Report are:

a. The correct flow of salt water around effects X, XI, and XII.

b. The product split flow around heat exchangers 214 and 215.

c. Condenser cooling water flow.

d. The connections for the barometric condenser and vacuum pump.

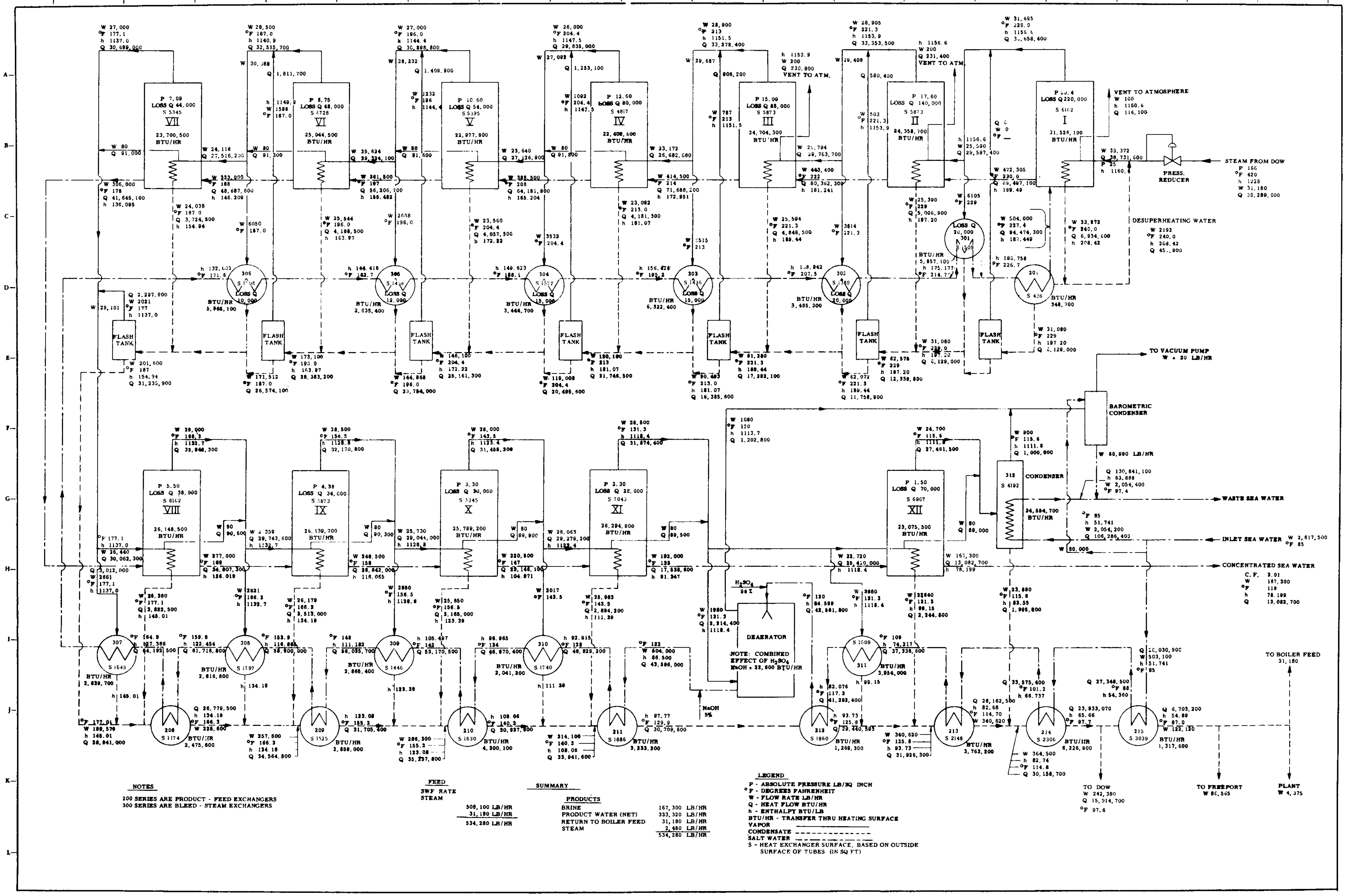

FIGURE 5-1. PLANT HEAT AND MATERIAL BALANCE FLOW DIAGRAM

The flow of desuperheating water and the positions of the pressure reducing valve and flow control valve in the heating steam supply are also depicted on Figure 5-1

5-35 <u>Over-All Thermal Performance.</u>

5-36 The operation of a multiple effect evaporator provides many interesting examples of the laws of Thermodynamics in operation. Tabulations are provided to illustrate results being achieved at Demonstration Plant No. 1 (refer to Tables 5-1 through 5-4). In presenting the economy ratio, care must be taken to specify the basis on which it is calculated. For this series of data the following definitions apply.

$$\text{Over-all Economy Ratio} = \frac{\text{Net Product}}{\text{Boiler Steam Production for Plant}}$$

Gross Product = Measured Product Flow

Net Product = Gross Product Minus Boiler Makeup Requirements.

5-37 By using these definitions, all losses are automatically charged to the process. The net economy ratio automatically charges the process for boiler feed water requirements. If, as others have done, the condensate from the first effect (or brine heater) is used as a measure of the steam to the first effect, the vent losses, line losses, and throttling losses are not accounted for. An economy ratio based on the condensate from the first effect (or brine heater in the case of MSF) is therefore not conservative or directly comparable to those shown in the following data.

5-38 The corrected economy ratio is provided at the direct request of OSW. It is calculated by applying a correction factor to the steam supplied in a superheated, high-pressure state. This factor is calculated from an enthalpy balance around the desuperheater. For Tables 5-1 through 5-3, the correction factor is established at 1.07. For Table 5-1, the steam supply is assumed to be saturated at 170 PSIG (for comparison to the San Diego Plant). The correction factor then becomes 1.036. It is applied as follows:

(Correction Factor) = (32,000) (1.036) = 33,152

24-HOUR AVERAGE FOR 3-27-64	FINAL CONCENTRATION FACTOR 3.
Product extracted, LBS	8, 134, 600
Steam fed to plant, LBS	751, 000
Gross product, LBS	8, 885, 600
Gross product, GAL	1, 069, 260
Economy Ratio Over-all (net)	10. 83
Economy Ratio Corrected (net)	10. 12
Final Concentration factor	3. 00
Product Temperature (312 Condensate)	115°F
Brine Overboard Temperature	120°F
Steam to First Effect	243°F
Sea Water Feed LBS	11, 640, 000

TABLE 5-1. THERMAL PERFORMANCE SUMMARY - DEVELOPMENT RUN OF 27 MARCH 1964

24-HOUR AVERAGE FOR 6-23-64	FINAL CONCENTRATION FACTOR 2.
Product extracted LBS	6, 958, 665
Steam Fed to plant, LBS	626, 300
Gross product, LBS	7, 584, 965
Gross product, GAL	910, 560
Economy Ratio Over-all (net)	11. 11
Economy Ratio Corrected (net)	10. 38
Final Concentration Factor	2. 63
Product Temperature	114°F
Brine Overboard	118°F
Steam to First Effect	229°F
Sea Water Feed LBS	10, 916, 000

TABLE 5-2. THERMAL PERFORMANCE SUMMARY - DEVELOPMENT RUN OF 23 JUNE 1964

ITEM	DATA
Product extracted, LBS/HR	333, 320
Steam fed to plant LBS/HR	31, 180
Gross product, LBS/HR	364, 500
Economy Ratio Over-all (net)	10. 69
Economy Ratio Corrected (net)	10. 01
Final Concentration Factor	3. 01
Product Temperature (at 312)	115. 6 ^{o}F
Briné Overboard Temperature	118. 0^{o}F
Steam to First Effect	240 ^{o}F
Sea Water Feed LB/HR	503, 100

TABLE 5-3. HEAT AND MATERIAL BALANCE DATA - ACTUAL

ITEM	DATA
Product extracted LBS/HR	350, 000
Steam Fed to Plant LBS/HR (at 170 psig SAT)	32, 000
Gross Product	382, 000
Gross Product, GAL per day	1, 103, 980
Economy Ratio Overall	10. 94
Economy Ratio Corrected (from 170 psig Sat Stm Basis)	10. 56
Final Concentration Factor	3. 3
Product Temperature	116^{o}F
Brine Overboard Temp.	119^{o}F
Steam to First Effect	250^{o}F
Sea Water Feed LBS/HR	495, 000

TABLE 5-4. OPTIMIZED DATA

Then the corrected Economy Ratio =

$$\frac{350,000}{33,152} = 10.56 \text{ (Net)}$$

5-39 The optimized data (refer to Table 5-4) is presented to fill the gap created by development type operations. All conditions specified therein have been realized using present equipment, but not at the same time. It would be an unfair evaluation of the process to omit this data. Sometime during the coming year, optimum condition runs will be tested using all the results of development work.

5-40 For comparison purposes, data from the Burns and Roe Annual Report[1] have been extracted and converted to the same basis as outlined above. The referenced report presents runs at 200°, 225° and 250° heater temperatures. Table 5-5 compares the economy ratios of Demonstration Plant No. 1 with Demonstration Plant No. 2; ratios for both Plants were calculated from identical bases.

DATA TAKEN AT (TEMPERATURE)	OVERALL ECONOMY RATIO		CORRECTED ECONOMY RATIO		REFERENCE	
	LTV Freeport	MSF San Diego	LTV Freeport	MSF San Diego	Table	Ref'd Report Page
200°	11.11	10.11	10.38	9.79	5-2	38
225°	10.83	10.81	10.12	10.38	5-1	40
250°	10.94	9.93	10.56	9.62	5-3	42

TABLE 5-5. COMPARISON OF ECONOMY RATIOS

5-41 The last two comparisons illustrate the penalty which the LTV, Freeport Plant pays for throttling and desuperheating high pressure, superheated steam.

[1] Burns and Roe, Inc., Annual Report, Saline Water Conversion Demonstration Plant No. 2, July 1, 1963 through February 26, 1963, pages 38, 40, and 42.

5-42 The operating parameters established in Development Report No. 1 are broader in their coverage of the over-all performance. The results achieved in Development Run Nos. 2 and 4 and the original design performance data are shown in Table 5-6. Run Nos. 2 and 4 are selected as being representative for two different final concentration factors. The variation of separation work with extraction ratio is illustrated by the differences in this parameter between Run No. 2 and Run No. 4. Overall work is the sum of both separation work and pumping work. Use of the available energy is a function of the over-all plant temperature difference, and in particular the heat rejection temperature. The variation of all of these things is shown as derived from the operating data and the original design data. Also shown in Table 5-6 are the established standards of attainable goals for the operation of this Plant.

5-43 An item of particular interest is the over-all economy ratio, wherein Run No. 2 was lower than in Run No. 4, indicating that better thermal economy is possible at lower concentration ratio. The decrease in electrical energy per pound of product is most likely an inverted occurrence attributable to the completion of the major painting program and the resulting standby status of the air compressor. Steam pressure in the first effect was lower, indicating that it was relatively easy to drive the process. This is also reflected in a lower over-all plant Δ T for the lower extraction ratio. It was also possible to reach design temperatures and pressures in the XII effect and 312 condenser at the low concentration ratio. No scale is formed in the LTV evaporator when final concentration factors are less than 2.8, even though polyphosphate was not added during Run No. 4. Run No. 3 formed no scale as a result of the addition of a proprietary blend of polyphosphates in P-18.

5-44 The available energy consumed in separation work was significantly less for the lower extraction ratio, but the pumping work was higher.[2] Even so,the total available energy consumption was less at an extraction ratio of .641 than at .704. This is predictable from theory[3] and

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[2]Dodge, B. F. and Eshaya, A. M., "Thermodynamics of Some Desalting Processes", Advances in Chemistry 5-15 Series No. 27 (1960) American Chemical Society.

[3]Kays, D. D., "A Minimum Work Determination Method for the Separation of Water from a Subsurface Saline Solution" (University of Wichita, Masters Thesis, January 1960).

DESCRIPTION	UNITS	ORIGINAL DESIGN	ESTABLISHED STANDARD	RUN NO. 2	RUN NO. 4
Total Production	Gallons			31,248,690	13,400,180
Operating 24 Hour Days				29.48	14.94
Average Rate	GPD	1,000,000	1,060,000	1,059,996	897,142
Net LBS water produced per LB steam fed		11.65	11.0	10.572	10.72
*LBS water extracted per KWHR electric energy		1,025	1,000	999.6	815
Steam pressure of Dow Steam	PSIA	174.7	164	169.11	143
Steam temp of Dow Steam	^oF	530	500	500.6	489
Steam pressure in first effect (Saturated)	PSIA	36	25	26.18	3.80
Steam pressure in XII effect V.H. (Saturated)	PSIA	1.47	1.50	1.299	1.18
Pressure in final condenser	PSIA	--	1.25	--	--
Incoming sea water temperature (to 312)	^oF	85	80	70.2	86
Discharge brine temperature (at P-22)	^oF	119	117	120.3	117
Dow product temperature	^oF	88	109	93.1	96
Freeport & Plant Use Product Temperature	^oF	88	88	84.8	92
Inlet Concentration Factor (1.0 = 35,200 PPM TDS)		1.0	1.0	.872	.92
Final Concentration Factor (1.0 = 35,200 PPM TDS)		4.0	3.0	2.943	2.56
Steam temperature in first effect chest	^oF	260.95	250.0	244	221
Steam temperature in XII effect V.H.	^oF	115	115	117.3	115
Average Total plant _T	^oF	145.95	135	126.7	106
SWF temperature to first effect evaporator	^oF	245.8	240	230.7	221
Maximum steam demand	LB/HR	31,500	31,500	32,700	30,300
Concentration of O_2 in D.A. Effluent	PPB	100	100	80	--
BiCarbonate Alkalinity of the D.A. effluent as equivalent $CaCO_3$	PPM	--	. 10	7	11
Work of Separation	Watt hours per LB	13.814	13.00	14.236	13.62
Work of pumping	Watt hours per LB	1.06	1.09	1.095	1.439
Extraction Ratio	LB H_2O/LB SWF	.75	.6667	.704	.641
Total Scale Produced	LB	--	--	none	none
Average Scale formation rate	LB/day	--	--	none	none
Beginning heat transfer surface	FT^2	65,400	66,882	66,812	66,882
Ending heat transfer surface	FT^2	--	66,882	66,882	66,882
Electric energy (pump work) to SWF	KWHR/M LBS	--	0.750	0.773	.915
66^o Be H_2SO_4 consumption	LB/M LB SWF	--	0.120	0.121	0.121
50% NaOH consumption	LB/M LB SWF	--	0.010	0.011	0.011
Estimated direct water cost	¢/M GAL			65.88	71.36
Estimated total GAL water cost	¢/M GAL			107.17	121.78

*All product at 88^oF for design

TABLE 5-6 PLANT OPERATING PARAMETERS

determines that the minimum available energy point consumption is reached at some extraction ratio less than .641. However, the cost estimates for Run No. 2 show that the economic minimum is above the .641 extraction ratio. Runs at 3.5 (extraction ratio .714) final concentration factor are in progress as this report is being written, and will complete the study of energy and economy as related to extraction ratio. All of the above-mentioned parameters were defined and/or derived in Development Report No. 1. The definitions and derivations are not repeated here, but do appear in the Appendix of this Annual Report.

5-45 Individual Equipment Thermal Performance.

5-46 The Heat and Material Balance Flow Diagram, Figure 5-1, is the primary source of data and calculated values for this Section. This heat and material balance represents the Plant in normal operation with an inlet concentration factor of 1.0 and an outlet concentration factor of 3.0. It illustrates the effect of the apparently arbitrarily chosen number of tubes installed in the various heat exchangers and evaporators. It also illustrates the use of flash tanks versus 200-series exchangers (condensate to sea water feed).

5-47 Evaporators. Table 5-7, Individual Evaporator Performance, illustrates the amount of extraction from each evaporator. This can be compared to Table 7 in Development Report No. 2 which is calculated from instantaneous operating data, and in particular the data of April 13, 1964. The heat and material balance is more rigorous because the results shown in Development Report No. 2 are not necessarily balanced. Therefore, we consider the heat and material balance to be more accurate.

5-48 Note, especially, the restriction imposed by the limited heat transfer surface in effect IV. The steam production from evaporator 2 and flash tank 3 can not be completely utilized in evaporator 4, so excessive load is placed on the condensing heater 303. Heater 304 is underloaded since steam production from effect IV is not adequate to meet the needs of V and the condensing heater 304. At present rates, the surface in effect IV is not imposing a serious restriction, because of ample surface availability and temperature differences in 303. If the full potential of the rest of the evaporation train is to be utilized, we suspect that additional surface will be needed in evaporator 4. The surface now installed in each evaporator is as shown in Table 5-8.

Effect No.	TDS (φ) In	TDS (φ) Out	Temperatures °F Cone Brine	Temperatures °F Vapor Hd. Steam	Temperatures °F Stm. Chest Stm.	Pressures psia Brine Eq. V.P.**	Pressures psia Vapor Head	Pressures psia Steam Chest	Flows (lb/hr) Brine In	Flows (lb/hr) Brine Out	Flows (lb/hr) Steam To Htr	Flows (lb/hr) Steam Fr Fltr	Flows (lb/hr) Steam To Next Chest	Water Extr.
I	.0352	.0376	230	229.0	240.0	20.48	20.40	25.0	504,000	472,305	(301) 6105	0	25,590	31,695
II	.0376	.0400	222	221.3	229.0	17.63	17.60	20.40	472,305	443,400	(302) 3614	503	25,794	28,905
III	.0400	.0427	214	213.0	221.3	15.07	15.00	17.60	443,400	414,500	(303) 6515	787	23,172	28,900
IV	.0427	.0456	206	204.4	213.0	12.06	12.60	15.00	414,500	388,500	(304) 3532	1092	23,640	26,000
V	.0456	.0490	197	196.0	204.4	10.67	10.60	12.60	388,500	361,500	(305) 2688	1232	25,624	27,000
VI	.0490	.0532	188	187.0	196.0	8.78	8.75	10.60	361,500	333,000	(306) 6050	1588	24,118	28,500
VII	.0532	.0579	178	177.1	187.0	7.099	7.09	8.75	333,000	306,000	(307) 2661	2021	26,440	27,000
VIII	.0579	.0640	168	166.3	177.1	5.58	5.50	7.09	306,000	277,000	(308) 2821	-	26,259	29,000
IX	.0640	.0706	158	156.5	166.3	4.39	4.36	5.50	277,000	248,500	(309) 2850	-	25,730	28,500
X	.0706	.0804	147	143.5	156.5	3.33	3.30	4.36	248,500	220,500	(310) 2017	-	26,063	28,000
XI	.0804	.0922	133	131.3	143.5	2.305	2.30	3.30	220,500	192,000	(311) 3880	-	22,720	28,500
XII	.0922	.1060	118	115.6	131.3	1.516	1.50	2.30	192,000	167,300	(312) 24,780	1980*	24,780	24,700

*Steam to Deaerator
**Gastaldo, UCLA Report 61-80

T ABLF 5-7. INDIVIDUAL EVAPORATOR PERFORMANCE
(Based on Heat & Material Balance)

Effect No.	Tubes	Inside Surface Area
I	490	5758
II	467	5487
III	467	5487
IV	383	4500
V	429	5041
VI	535	6286
VII	425	4994
VIII	490	5758
IX	467	5487
X	425	4994
XI	560	6580
XII	554	6510

TABLE 5-8. EVAPORATOR HEAT TRANSFER SURFACE

5-49 The same type of restriction exists at effect X and results in a lack of utility of the large surfaces in effect XI and XII. These effects were retubed in early 1963. The surface in effect XI was increased by 33 percent and by 42 percent in effect XII. At the time, the increase was accomplished to decrease vapor velocity and resultant carryover. The imbalance in heat transfer load can be seen in the drop in steam to heat exchanger 310 and heavy increase to heat exchanger 311. The withdrawal for deaeration also limits the utility of surface in effect XII.

5-50 A discussion of thermal performance of the evaporators must include an analysis of heat transfer characteristics. Table 5-9, Evaporator Performance Data, is representative of information obtained in a development run. It is used as the basis for making heat balances around the individual evaporators. The following discussion is presented as typical of the analysis performed after each development run. The results indicated are valid.

5-51 A typical heat balance diagram for effect I is shown in Figure 5-2. The heat balance data derivation is illustrated in the sample calculation sheets which follow Figure 5-2. Calculations were made on all five evaporators for the data of 28 February 1964. In addition to these calculations, the same set were made for effect XI on the data of 5 March 1964, since it had changed appreciably due to scaling between 28 February and 5 March 1964. The method of transferring

EFFECT NO.	DATE	NO. OF TUBES	FT^2 SCALE FREE SURFACE	FT^2 SCALE FREE FLOW AREA	BRINE RATE TO HTG ELEMENT LBS/HR	BRINE IN CONCENTR. CHLOROSITY	BRINE IN CONCENTR. ^oF TEMP	BRINE IN CONCENTR. $C.F._i$	BRINE OUT CONCENTR. CHLOROSITY	BRINE OUT CONCENTR. TEMP	BRINE OUT CONCENTR. $C.F._o$	STEAM IN HTG ELE. PR. PSIA	STEAM IN HTG ELE. ^oF TEMP	STEAM FLOWING FR. VAPOR HD PR. PSIA	STEAM FLOWING FR. VAPOR HD ^oF TEMP	E. R.	YIELD CALCULATION #VAPOR FLOWING OUT/HOUR
I	2-25	490	5758	9.344	475,000	15.40	□ 288	.784	16.41	230	.826	23.2	236.0	20.5◇	229.2	.051	24,225
	2-28					16.25	□ 288	.820	17.46	230	.886	23.95	237.7	20.5◇	229.2	.074	35,150
	3-5					17.46	□ 230	.886	18.60	232	.944	24.9	239.9	21.2◇	231.2	.061	28,975
VI	2-25	535	6286	10.202	322,295	21.98	196	1.110	23.92	188	1.208	10.51	195.5	8.76	187.1	.081	26,916
	2-28					23.43	199	1.186	25.61	190	1.292	10.75	196.6	9.14	*189	.082	27,248
	3-5					25.11	199	1.267	27.69	193	1.388	11.24	198.8	9.54	*191	.087	28,910
X	2-25	425	4994	8.105	210,560	33.20	158	1.648	38.00	144	1.870	5.25	*164.3	3.14	*143.2	.119	25,057
	2-28					35.78	160	1.771	40.68	145	2.001	4.47	*157.8	3.00	*141.5	.115	24,214
	3-5					38.18	161	1.881	44.02	146	2.151	4.57	*158.5	3.04	*142.0	.126	26,531
XI	2-25	560	6580	10.679	471,890	38.00	144	1.870	43.04	132	2.112	3.14	*143.2	2.16	*128.9	0.115	21,332
	2-28					40.68	145	2.001	46.62	134	2.275	3.00	*141.5	2.16	*128.9	0.120	25,100⊠
	3-5					44.02	146	2.151	49.64	134	2.416	3.04	*142.0	2.31	*131.4	0.110	28,000⊠
XII	2-25	560	6580	10.679	581,370	43.04	132	2.112	51.94	122	2.519	2.16	*128.9	1.33	*111.4	.162	
	2-28					46.62	134	2.275	56.62	120	2.735	2.16	*128.9	1.56	117.0	*.168	27,089
	3-5					49.64	134	2.416	60.46	120	2.908	2.31	*131.4	1.33	*111.4	.169	

⊠ Does not coincide with E. R. Calculation but does give Heat & Material Balance

* Calculated from measured saturation pressure

□ Taken from daily log sheets

◇ Calculated from BPR & Cone brine temperature

TABLE 5-9 EVAPORATOR PERFORMANCE DATA

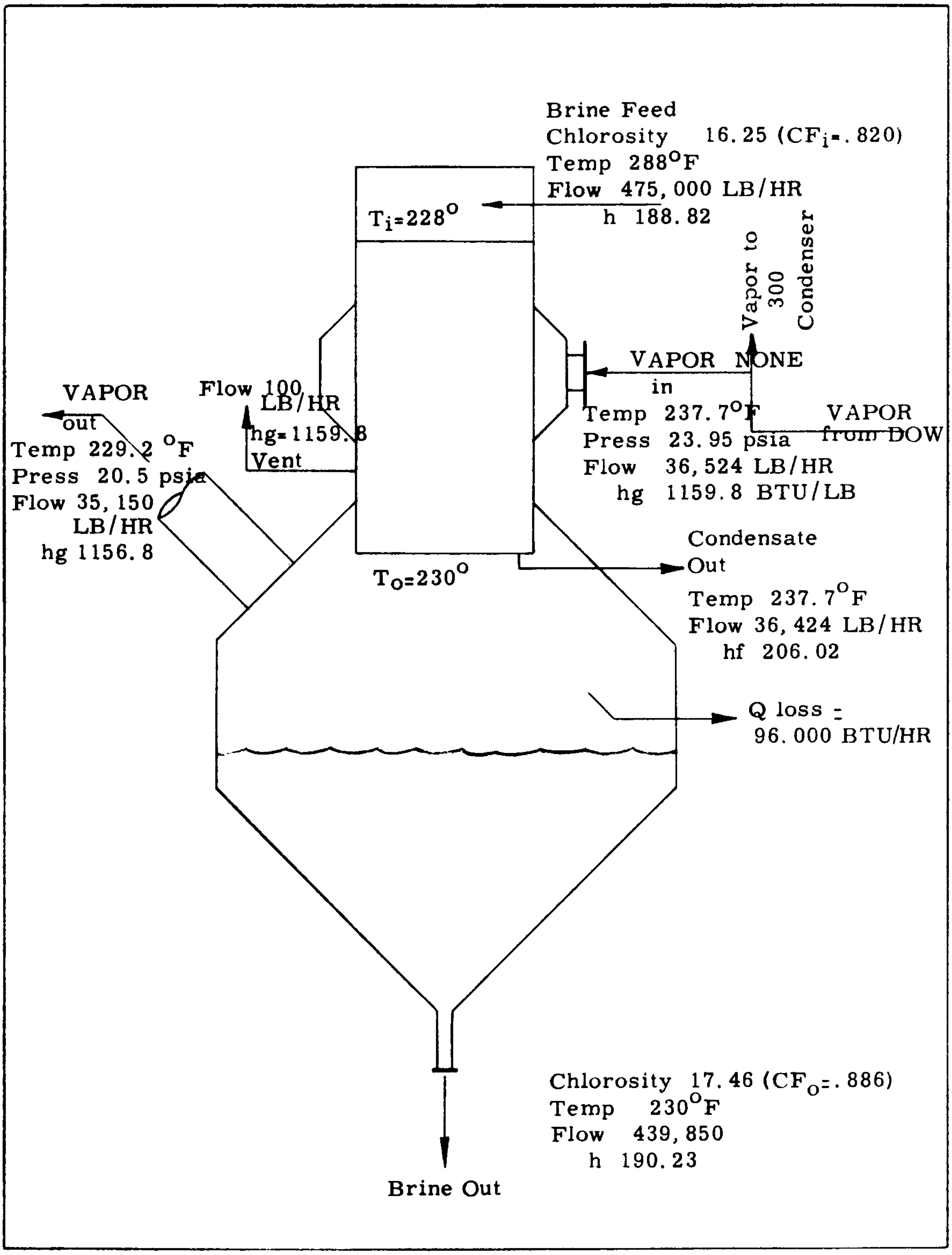

FIGURE 5-2. TYPICAL HEAT BALANCE DIAGRAM

brine from effects X to XI and from XI to XII make it essential to start with affect X and work through effect XI and then effect XII to determine yields from extraction ratios. Good approach to equilibrium is not indicated in effect XI. There is stratification in the cone brine which gives unreliable readings of chlorosity. The vapor temperature and vapor pressure measured in effect XII either indicated a superheated condition or an error in the pressure measurement. Since the temperature corresponded well with the brine temperature less the BPR, it was assumed correct and the pressure adjusted accordingly. The use of constant input does not account for changes in performance of the preceding evaporator except in effects XI and XII.

5-52 The quantities calculated are contained in Table 5-10, Evaporator Performance Comparisons. These quantities are self-explanatory to a certain extent. Pressure drops across the heating elements were not measured but were estimated by comparison to measurements made by the writers of R & D Report No. 74. The apparent Ua corresponds to the uncorrected "U" calculated by the method proposed in R & D Report No. 74, for the "Report No. 456" run (selected as being most comparable). The Reynolds No., (Re_2), Dimensionless Interfacial Shear, (B) Prandtl No. (Pr_2) and heat flux are all calculated by methods illustrated in the following sample calculations. Two sets of values are reported for effect XI for the data of 5 March 1964. The first set assumes no scaling. The second set accounts for the reduction in heat transfer surface and flow area in accordance with the measured scale condition.

5-53 The "U" reported in effect I was lower than that reported in the comparable run at WB. This can be accounted for by the fact that the SWF is actually subcooled in the operating situation whereas this condition probably did not exist and could not have been accounted for by the "U" calculation based only on tube exit brine temperature. The increase in coefficient for effect VI would normally be expected on the basis of the large increase in Reynolds No. The increased B also would have a beneficial effect on "U" according to Dukler[4], but entrainment is probably high. The low coefficient in effect X is difficult to explain, especially in view of the fact that heat flux seems normal or even slightly high. Effect XII falls closely in line, but this would probably not be the case without recirculation.

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[4]Prengle, Dukler and Crump, Inc., Saline Water Research and Development Progress Report No. 74, June 1963.

QUANTITY DESCRIPTION	SYMBOL	EFFECT I		EFFECT VI		EFFECT X		EFFECT XI				EFFECT XII	
		data of 2-28-64	Rept. 456 Run LWCF-2	data 2-28-64	Rept. 456 Run LWAT-5	data of 2-28-64	Rept. 456 Run No. LWCB-2	data 2-28-64 clean	data of 3-5-64 clean	data of 3-5-64 Scaled	Rept. 456 Run No. LWCK-26	data of 2-28-64 clean	Rept. 456 Run No. LWCA-1
Temp. Diff. at Tube Inlet °F	TI	9.7		5.1		15.8		6.5	8	8		8.9	
Temp. Diff. at Tube Outlet °F	Ta	7.7	6.76	6.1	6.5	14.0	11.21	5.5	7	7	14.28	6.9	12.47
Log Mean Temp. Diff. °F	LMTD	8.65		5.49		14.75		6.05	7.46	7.46		7.87	
Temp. of Vapors Leaving °F	Tv	229.2	222.4	189	193.3	141.5	139.3	128.9	131.4	131.4	124.8	117.4	120.0
Press. Drop Across Heating Element Inches Hg	Pw	.28	0.17	.28	.26	.28	.40	.08	.08	.39	.57	.08	.45
Apparent (Unoorrected) "U" (BTU/HR-FT^2 °F)	Ua	783.6	727	606.5	513	330	436	620	551	777	423	619	401
Corrected (LMTD) "U"	Uc	697.5	735	674	592	313	554	558	517	731	555	542	561
Theoretical "U"	Ut		744		612		571				595		560
Reynolds No. at Tube Exit (Dimensionless)	Re_2	9530	8220	5560	1217	2892	2860	5037	4839	6866	4924	5192	1825
Dimensionless Interfacial Shear	B	7.360	4.56	37.7	0.50	5.23	7.95	1.42	1.38	6.56	12.75	21.2	8.70
Prandtl No. at Exit (Dimensionless)	Pr_2	1.63	1.70	1.88	2.18	3.18	3.29	3.41	3.55	3.55	3.53	4.08	4.06
Heat Flux BTU/FT^2-HR	Fh	6034	4969	3700	3335	4619	4888	3406	3855	5451	6040	4262	5000

TABLE 5-10 EVAPORATOR PERFORMANCE COMPARISONS

5-54 The three sets of data for effect XI present some interesting possibilities. It is quite likely that some scale existed on 28 February and even more on 5 March 1964. The "U" calculated on the total surface (ignoring scale) decreased as was predicted in the monthly operating reports. When the heat transfer is considered to proceed without the utilization of scaled tubes, a marked increase, roughly proportional to the increase in Reynolds No., is seen. Even under the "clean" assumption, the coefficient is considerably improved by recirculation when compared to the Report No. 456 run. It is also worthy of note that no heat flux was as great as 10,000 BTU per hour per SQ-FT. Recently, published investigations indicate that nucleate boiling does not occur at fluxes below 10,000 BTU. The vaporization is therefore most likely occurring at the liquid vapor interface. This presents an interesting problem in how far one should go towards improvements of "U". If nucleate boiling were to occur as a result of improved "U" it might be difficult to maintain a continuous liquid film on the tube surface.

5-55 Effects XI and XII are not operating as true LTV evaporators. The liquor from effect X is transferred to effect XI by an equalizing pipe. It flows under the influence of gravity and the pressure difference between the two vapor heads. The same is true for the transfer from effect XI to effect XII. In both cases, a visual observation confirms that rather violent flashing occurs at the point of injection of the hot weak liquor. There is continuous mixing induced by the flashing of the incoming stream in the cone brine. Brine for feed to the heating element of effect XI is withdrawn from the cone of effect XI and circulated to the inlet water box. This is the principle of forced circulation. It necessarily contacts the heating surface with more concentrated brine than this surface would experience if it were receiving the effect X cone brine directly. These same statements can be made about the effect XII.

5-56 This mode of operation was established as a convenient method of accomplishing recirculation in these effects. It is certain to contribute to carryover. It may also be counteracting a part of the beneficial effect of higher flow rates on scaling tendency. Its degree of success does raise the possibility that future designs might be made purposely to utilize forced circulation in the latter effects. If this were proven to be desirable, it would appear that perhaps submerged tube heating might be more attractive. As it now stands, equilibrium conditions are not being closely approached in these effects with a resultant loss in full utilization of the equipment. Their performance is difficult to analyze and stratification of weak and strong brines is highly probable.

5-57 Instrumentation to measure temperatures of the liquid at the tube inlet or the tube outlet has not, as yet, been successfully developed. Also, vapor pressure measurements at the tube inlet are not yet possible. Arrangements are being made to measure the pressure between the tube sheet and the distributor plate. A long thermowell has been inserted into the vapor body to measure the impinging liquid leaving the tubes. A more accurate and reliable system will utilize thermocouples just inside the tube at each end, and will require a new readout instrument. The calculations and comparisons of Table 5-10 illustrate the value of establishing these conditions. The indicated correspondence strongly support the contention that such parameters may be reliably correlated for improved design techniques.

5-58 A major goal of this development program is to accomplish first effect temperature increases with the resulting increases in economy of both energy and capital investment. A reliable prediction of evaporator performance, without sacrificing the ability to control the formation of scale, is essential to achieve this goal. This will involve such things as staying below the nucleate boiling range and specifying the quantity of desirable recirculation. The ability to predict these and other factors under varying conditions will be determined by continued analysis of evaporator performance.

5-59 It is obvious that the thermal performance of these evaporators is a function of several mechanical variables such as: flow rate; pressure drop; and brine distribution. A large effort is being exerted into improvements here. It is our firm belief that the full utilization of this heat transfer surface will not be attained until production rates exceed 1,400,000 GPD.

5-60 Heat Exchangers. This discussion is devoted to the heat exchangers which perform the function of heating sea water feed by recovering heat from extracted water vapor and condensate. Their function is highly important since they recover approximately 68,000,000 BTU/HR from this source. This recovery is almost twice the heat that is supplied to the system by heated steam. Over a year ago, the 200-series exchangers 202 through 207 were removed from service, and the flash tanks associated with the first seven effects were put into continuous service.

5-61 The 300-series exchangers are all condensing heaters, in that they take steam from the vapor head and the associated evaporator and condense it in the shell while the tubes are flowing sea water feed. Exchangers 301 through 311 are heaters and recover heat from the product, but exchanger 312 is a heat rejection condenser. Table 5-11,

Hx	Surface Area	TEMP B_i	C_o	B_o	C_i	ΔT_i	ΔT_o	LMTD	Qe	"U"
301	1509.1	214.7	229	226.7	229	14.30	2.70	6.97	5,857,100	556.8
*302	1388.6	207.5	221.3	214.7	221.3	13.80	6.60	9.78	3,485,300	256.6
*303	1435.8	195.2	213	207.5	213	17.80	5.50	10.17	6,322,400	432.9
304	1572.0	188.1	204.4	195.2	204.4	16.30	9.20	12.46	3,444,700	175.9
305	1435.8	182.7	196	188.1	196	13.30	7.90	10.42	2,635,400	176.2
306	1708.2	171.6	187	182.7	187	15.40	4.30	8.70	5,965,100	401.4
307	1645.4	164.9	177.1	171.6	177.1	12.20	5.50	8.42	2,639,700	190.5
308	1797.3	153.9	166.3	159.8	166.3	12.40	6.50	9.13	2,816,800	171.7
309	1446.2	142	156.5	148	156.5	14.50	8.50	11.19	2,865,400	177.0
310	1739.7	128	143.5	134	143.5	15.50	9.50	12.30	2,041,200	95.4
311	1608.7	109	131.3	117.3	131.3	22.30	14.0	17.89	3,954,000	137.4
312	4192.0	85	115.6	97.4	115.6	30.60	18.20	23.94	24,554,700	244.67
201	536.0	226.7	229	227.4	240	2.30	12.60	6.06	348,700	132.0
208	1173.8	159.8	166.3	164.9	177.1	6.50	12.20	9.05	2,475,600	233.0
209	1524.8	148.0	155.2	153.9	166.3	7.20	12.40	9.81	2,859,000	195.1
210	1629.6	134	140.2	142.0	155.2	6.20	13.20	9.15	4,300,100	288.5
211	1886.4	122	129.9	128	140.3	7.90	12.30	9.89	3,233,200	173.3
212	1860.2	117.3	125.8	120	129.9	8.50	9.90	9.46	1,269,300	72.1
213	2148.4	101.2	114.7	109	125.8	13.50	16.80	15.35	3,763,200	114.1
214	2305.6	88.0	97.7	101.2	114.8	9.70	13.60	11.54	6,226,900	235.5
215	3039.2	85	87.0	88	97.7	2.00	9.70	4.87	1,317,600	89.05

Explanation of Headings

Hx = Heat Exchanger No.
Surface Area = FT^2
B_i = Brine in
C_o = Condensate Out
B_o = Brine Out
C_i = Condensate in
ΔT_i = $C_o - B_i$ in ^oF

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

TABLE 5-11 THERMAL PERFORMANCE, PREHEATING HEAT EXCHANGERS

Thermal Performance, Preheating Heat Exchangers, lists the process conditions, available surface, and the duty and transfer coefficient for each exchanger. The 300-series exchangers, exclusive of 312, have a total outside surface area of 17,286.8 square feet. The average heat flux is 2430.6 BTU/SQ-FT HR based on a total duty of 42,017,100 BTU/HR as shown on the Heat and Material Balance. Of this total, 30,339,700 BTU/HR is transferred in heat exchangers 301 through 307 with a total surface of 10,694.9 SQ-FT. The average heat flux here is 2836.8 BTU/SQ-FT HR. The heat transferred in exchangers 308 through 311 is 11,677,400 BTU/HR. These exchangers have a surface area of 6,591.9 square feet, so the heat flux is 1771.5 BTU/SQ-FT HR. This split is made to illustrate the effect of eliminating the 200-series exchangers, 202 through 207, and using the flash tanks in their place. The effect is a sizeable increase in the transfer load applied to the 300-series exchangers.

5-62 The heat exchangers 301 through 307 appear to be quite capable of handling the increased load. In fact, heat exchanger Nos. 301 and 303 are the only exchangers that approach a reasonable "U" value. The design of the internal baffling does not appear to be optimum, therefore, the increased turbulence of higher steam flow is probably helpful. The steam supply to these exchangers flows in the path of least resistance, and thereby, satisfies the requirements of the downstream evaporator rather than the exchanger. This accounts for the general existence of high LMTD. Exchanger Nos. 301 and 303 also illustrate the effect of locating exchangers between evaporators wherein the upstream evaporator has more surface than the downstream evaporator.

5-63 The 200-series exchangers can be similarly grouped. Exchanger 201 recovers only 348,700 BTU/HR from first effect condensate. It functions as a final boost to preheat the feed to boiling temperature at the pressure of the first effect. It could be eliminated without much effect on the operation. If it were eliminated, the first effect flash tank would become more functional (active) as a flash tank. The next group of 200-series exchangers is heat exchanger Nos. 208 through 213. The flow of condensate at each of these exchangers is augmented by the product from the preceeding (upstream) effect. These exchangers handle a total of 17,900,400 BTU/HR through 10,223.2 square feet of surface, or an average heat flux of 1751 BTU/SQ-FT HR. The "U" values are very low for a liquid-to-liquid heat exchanger, and indicate that baffling for proper flow across the sea water tube bank does not exist.

5-64 Heat exchanger Nos. 214 and 215 handle the product after (downstream of) the final condenser. Exchanger No. 214 handles the entire product flow, but exchanger No. 215 handles only the remainder after Dow product has been discharged. This is unfortunate, since a considerable quantity of heat is lost in the Dow stream (2,610,400 BTU/HR) and exchanger No. 215 has a very large heat transfer surface in comparison to any other 200-series exchanger. For maximum economy runs, it would seem desirable to operate at full flow through both heat exchanger Nos. 214 and 215. Exchanger No. 312 is the final product condenser. Its duty of 24,554,700 BTU/HR is heat rejected and irrevocably lost to the process. This represents a flux of 5858 BTU/SQ-FT HR. To take full advantage of the second law of thermodynamics, this unit should operate at the minimum temperature differential possible. Even without the consideration of efficiency, a closer approach to the temperature of the sea water circulating through the tubes should be achieved; which has the immediate effect of increasing the over-all Plant temperature difference without increasing the first effect temperature. This change in mode of operation would be reflected throughout the Plant as a greater Δ T across the surface, with more capability to transfer heat.

5-65 Condenser 312 does not perform within expected standards. At present production rates, this is the only undersized heat transfer unit in the Plant. The LMTD should be reduced to less than 15^{o}F, and the transfer coefficient should be in excess of 400. This problem is primarily the arrangement of tubes and baffling for good cross-flow and elimination of stagnant pockets. If this problem were corrected, the XII effect pressure would be reduced and carryover would be a problem. This problem, however, could be overcome with a suitable mist eliminator. The 312 condenser unit limits the achievable economy of this Plant. Its improvement would have a very favorable result.

5-66 Heat exchangers in general are not overloaded. From the duty standpoint, heat exchanger Nos. 208 thru 213 could probably be eliminated, and functionally replaced with flash tanks. At lower pressures, however, the flash tank and interconnecting piping would have to be liberally sized. If these five exchangers were eliminated and all of the heat recovered in heat exchanger Nos. 308 through 311, their duty would increase to 29,577,800 BTU/HR. This would represent a heat flux of 4,487 BTU/SQ-FT HR. The problems of vapor line sizing and flash tank adequacy would have to be evaluated. The available temperature difference would increase and the approach to the limiting transfer rate might be close. Note that larger ΔT's increase thermodynamic irreversibility.

5-67 Sufficient process calculations are not available to determine how the surface area was established for each exchanger. The pattern seems to be completely random and results in the same random pattern of LMTD, duty, and "U" factors. Adjustments of surface area figures may be in order when rates are increased.

5-68 Flash Tanks. The flash tanks are operating normally in accordance with the thermodynamic requirement. It should be noted that the weight of flashed vapor is continually increasing. This condition, coupled with reduced pressures and increased specific volumes, requires large flash tanks if they are to be used with the low temperature effects

5-69 Desuperheater. The desuperheater function is to reduce both the pressure and temperature of the Dow steam to levels that are useable in the first effect steam chest. The steam is first pressure-reduced and then desuperheated to saturation by the addition of condensate from the first effect. The steam then passes through a flow control valve. The throttling action of the flow control valve causes the steam to again become superheated. The desuperheater was designed to saturate steam at 36 psia, however, the Plant presently operates at 25 psia, and requires great quantities of desuperheating water Consequently, the desuperheating function was not being completed.

5-70 The system was modified to introduce additional desuperheating water downstream from the control valve. Special data-taking programs indicate that a temperature from $1/2^{o}$ to 2^{o}F above saturation is now being obtained. The benefits to heat transfer and process control in the first effect dictate that additional volumes of desuperheating water be added.

5-71 Re-Rating for Heat Transfer Surface.

5-72 A question was raised, after the submission of Development Report No. 1, concerning the capacity of the Plant. The changes made in heat transfer surface, first to effect No. XII (where the total number of tubes was increased) and then, in early 1963, to all effects containing carbon steel tubes, should have had a beneficial effect on operating economy. Full utilization of this benefit would require adequate pumping capacity and pipeline sizing, so that additional heat transfer could be realized without raising the final concentration factor. A better utilization of the available over-all ΔT can and does result from the decreased resistance to heat flow. A complete summary of the evaporator heat transfer surface, past and present, is provided in Table 5-12. The re-rated capacities are only a comparison of the prior and present resistance to heat flow, and an

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.
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.

Table 5-12 EVAPORATOR HEAT TRANSFER SURFACE SUMMARY accounting of total surface changes. The theory of this partial evaporator capacity re-rating follows:

The basic law of heat conduction is[5]: $q = k\frac{A}{L}T$

Over-all heat transfer is given by: $q = UA\,(t_1 - t_2)$

U is the over-all heat transfer coefficient and is related to the over-all resistance to the flow of heat by:

$$R = \frac{t_1 - t_2}{q} = \frac{1}{UA} = \frac{1}{h_{1a}A} + \frac{L}{K_{ab}A} + \frac{1}{h_{b2}A}$$

This can be simplified by eliminating A:

$$R = \frac{1}{U} = \frac{1}{h_1 a} + \frac{L}{K_{ab}} + \frac{1}{h_{b2}}$$

For comparison and re-rate purposes, calculate R for each effect before and after the change of tubes, and average the result over all 12 effects. Define improvement factor as follows:

$$\text{I.F.} = \frac{(\text{R ave before})\ (\text{Surface after})\ (.96^{**})}{(\text{R ave after})\ (\text{Surface before})}$$

Definition of Terms

h_{1a} and h_{b2} = film coefficients on the two sides of the tube wall

R = over-all resistance to heat transfer
U = over-all coefficient of heat transfer BTU/SQ-FT HR ^{o}F.
L = thickness of wall in feet
K_{ab} = thermal conductivity of material BTU/FT HR ^{o}F.
q = heat transferred BTU/HR
t_1 = steam-side temperature
t_2 = brine-side temperature
I.F. = improvement factor

[5] Jakob, Max, Heat Transfer, Volume I, John Wiley and Sons Inc., New York, 1949.

**This factor is applied because of the non-utility of surface added to effects XI and XII.

In Table 5-10, the revisions (present tubes) indicate an addition of 4, 320 square-feet, of which 3, 551 square-feet were added to the XI and XII effects. This was accomplished (as verbally reported to the writer) to decrease vapor velocity and entrainment, rather than to increase heat transfer capacity. The factor of .96 is applied to the improvement factor because of the above-stated reasons. Sample calculations follow:

Assume effect No. X represents a good average;

then for the data of 13 April 1964: $U_c = 524$ $L = \frac{.065}{12}$ $K = 58.0$

$$R = \frac{1}{524} = \left(\frac{1}{h_{1a}} + \frac{1}{h_{b2}}\right) + \frac{.065}{(12)(58.0)} \text{ or } \left(\frac{1}{h_{1a} + h_{b2}}\right) = .00182$$

For effect X before the change:

$$R = \frac{L}{12K} + .00182 = .0027 + .00182 = .00209$$

and after the change: $R = .0009 + .00182 = .00191$

The two film coefficients comprise approximately 90% of the total resistance. These coefficients are not a function of the tube material. The improvement factor calculation is:

$$\text{I. F.} = \left(\frac{.00199}{.00192}\right) \left(\frac{66,882}{62,562}\right) (.96) = (1.036)\ (1.069) = 1.06$$

5-73 Previous portions of this report have indicated that heat transfer surface is not limiting Plant production. Pumping capability limits the maximum sea water feed rate to 500,000 LBS/HR plus or minus 5,000 LBS/HR. (This was determined in Development Run No. 4.) Pumping capacity limitation and the pump which limits capacity are dependent upon mode of operation and inlet and outlet concentration factors. Improved capability to operate, scale-free, at higher temperatures and concentration factors, requires elimination of mechanical limitations. After overcoming mechanical limitations, the Plant will be re-rated, taking into account all factors including the proper placement of heating surfaces.

5-74 MECHANICAL FACTORS.

5-75 Mechanical factors are the key to many of the process problems, and bear direct influence on the Plant's productivity. The major mechanical factors are presented below.

5-76 Feed Water Distribution.

5-77 The successful use of recycle to increase flow rates and reduce scale formation in effects XI and XII confirms the need for better distribution to evaporator tubes. During Plant operation, the scaling pattern was consistently repeated indicating improper feed to the scaling tubes. A notched weir (see Figures 5-3 and 5-4) was designed to provide an even distribution of feed to all tubes irregardless of feed rates.

5-78 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 metal surface rather than spilling to the center. The 1-1/2-inch depth of the notch will provide even distribution of all expected flows without flooding. The design capacity of a single notch is 1.25 GPM at a liquid head of one inch, or a capacity of 3.75 GPM per tube. This rate is well above the maximum flow rate per tube which occurs in the effects utilizing recirculation.

5-79 The weirs are fabricated from sheet copper and designed to close tolerances. The fabrication process hardens the metal so that its elastic modulus is raised and it will hold itself tightly in the tube after being compressed for insertion. The butt gap serves two purposes, it allows the insertion in all tubes and provides a drain-down gap if operation is ceased. Gap dimension must be closely held to prevent excessive pull down of fluid at low rates of flow.

5-80 This weir was inserted into the top of all tubes in the X and XII effects for Development Run No. 2. The predictions were largely confirmed on 11 March 1964 by direct visual observation of the notched weirs through a plexiglass distributor plate located on the XII effect. An observation of the flow in the XI effect, verified that considerably more fluid falls down the center of the tube when the weirs are not used. For proper function, the tube sheets must be level, and the height of all the notches above the tube sheet must be the same. This requires that manufacturing tolerances be held to less than .005 of an inch for both the notches and the position of the insertion stops.

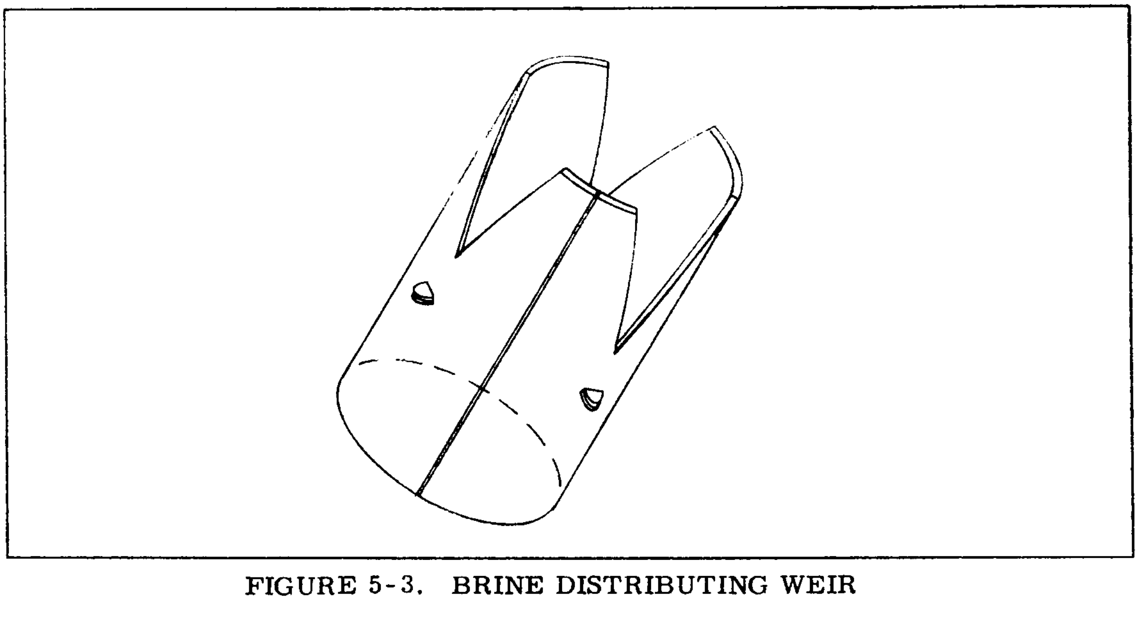


FIGURE 5-3. BRINE DISTRIBUTING WEIR

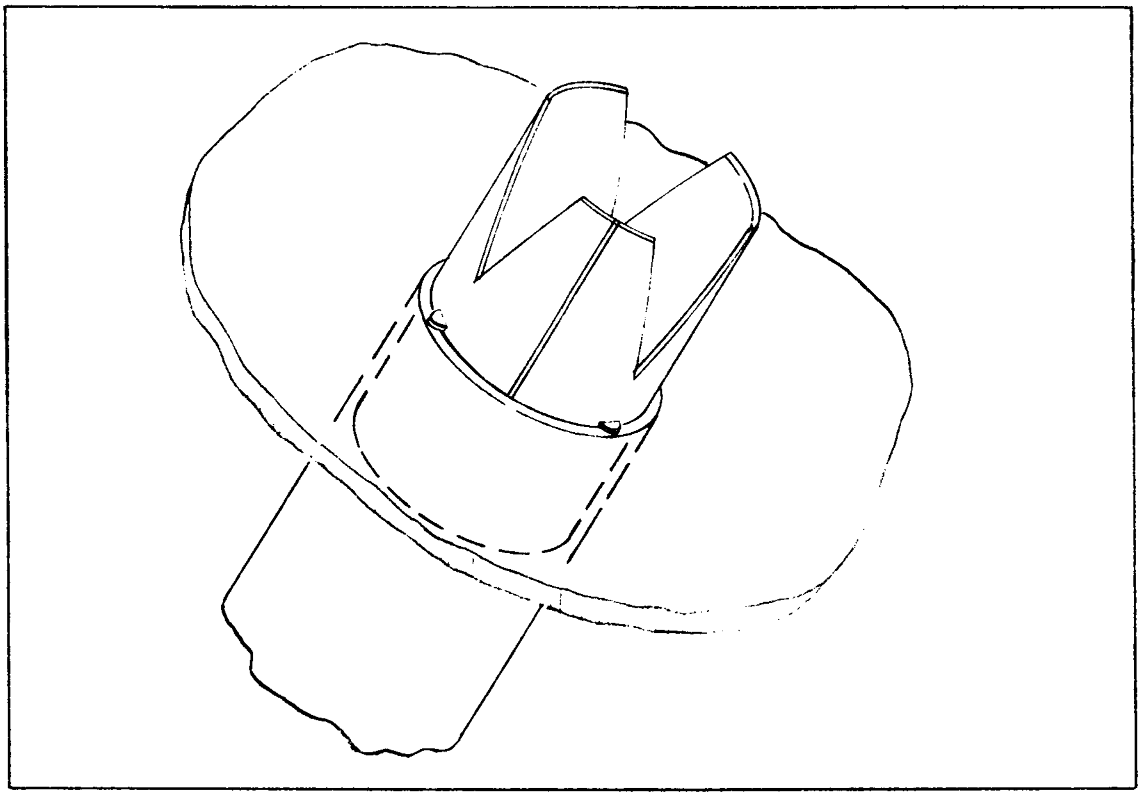


FIGURE 5-4. BRINE DISTRIBUTING WEIR INSERTED IN TUBE

5-81 No process changes were made for Development Runs No. 2 and No. 3. It was essential to hold process conditions as constant as possible at the same values as the previous run in order to evaluate the effect of the weirs on the formation of scale and the transfer of heat. Deaerator performance had been considered satisfactory since no scale was forming in the first effect.

5-82 Before process changes can be accomplished, a more complete and detailed knowledge of actual process occurrence is necessary. The results of Development Run No. 2 have significantly contributed to the data required to recommend process changes.

5-83 The success of the weirs in Development Run No. 1 and their favorable effect on heat transfer prompted the installation of these devices in effects I and XI for Development Run No. 3. (They were not installed during Development Run No. 2 since a positive, scale-effect, test result from polyphosphate compound addition at P-18 was desired.) During Development Run No. 3, polyphosphate was not added, and therefore, it was possible to positively identify the effect of the weirs in effect No. XI. The results were interesting, though not completely successful. Prior to the installation of weirs, more scale formed in effect XI than effect XII. When weirs were installed in effects X and XII, scale formation was practically eliminated in effect XII, but still occurred in effect XI. The installation of weirs in effects X, XI, and XII, moved all scale formation to effect XII. This occurred because the descaling effect on the cone-brine in effect XI removed $CaSO_4$ scale deposit. The combined effect of a lower concentration of calcium and sulfate ion and the elimination of starved tubes in effect XII prevented scale formation.

5-84 The installation of weirs in effect XI provided the distribution necessary to prevent the starving of tubes and subsequent scaling. Therefore, no calcium of sulfate ions were removed in effect XI and the scale came out in effect XII. Partial success is claimed because the scaling rate was greatly reduced. The scale was evenly distributed in the lower end of the tubes and plugging was limited to two tubes.

5-85 The value of adequate flow and good distribution applies to scale removal. When removing scale by circulating cold sea water, the use of weirs aid the process by the same distributing function.

5-86 Scale Measurement and Removal.

5-87 In order to evaluate the effect of process and mechanical changes, it was necessary to develop a method of quantitatively measuring the scale formed under a given set of conditions.

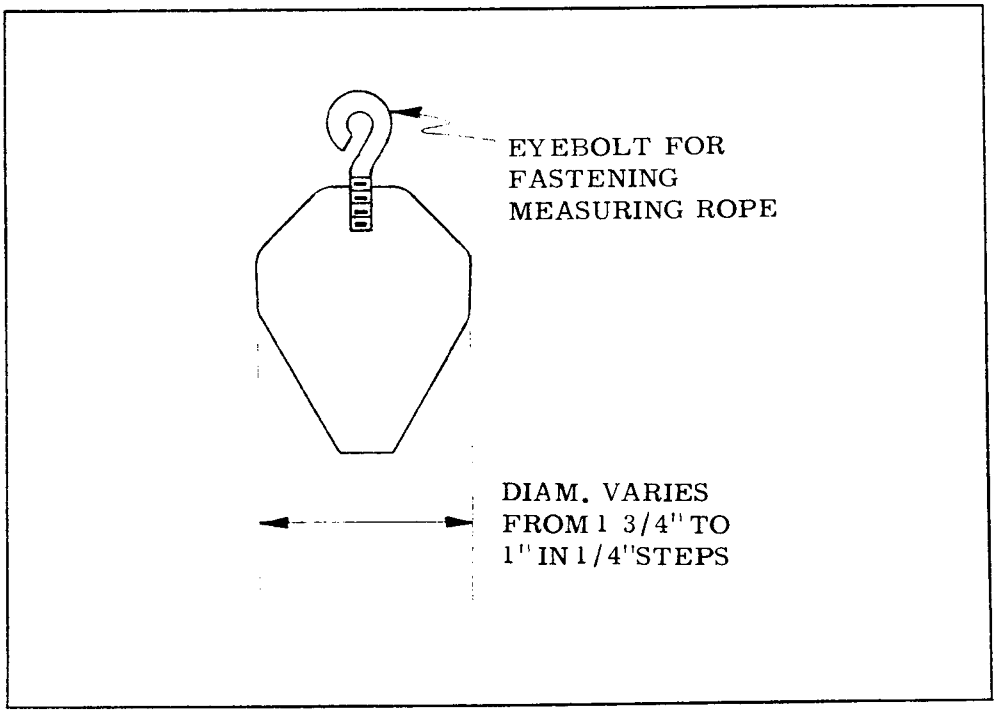


FIGURE 5-5. SCALE MEASURING GAGE.

5-88 The measurement of scale was performed by use of the scale gage pictured in Figure 5-5. This method is not entirely accurate, however, it is a compromise to reduce labor and downtime to a reasonable level and serves to affix a quantitative value to the scaling. No attempt is being made nor is it intended to account for scale formation in a material balancing which involve PPM ionic concentration in the discharge brine. (This would be the only required instance for accuracy beyond that yielded by the scale gage method.) The method incorporates a plan, whereby, only every other row of tubes is measured, and the results are doubled to obtain the total scale measurement. For tubes equipped with weirs, the 1-1/2-inch in diameter gage is inserted first (because the I. D. of the weir is less than 1-3/4-inch). Continued use of this method will produce comparable results.

5-89 If a tube will not accept the 1-inch diameter gage, it is assumed to be plugged below that point. This assumption has been generally verified by visual inspection. Such tubes are also considered unavailable for heat transfer, since it is impossible to pass sufficient liquid and vapor through them to accept what little heat that might be conducted through the built-up scale. All other tubes, scaled or not, are considered to be fully available for heat transfer. This last assumption counterbalances the error that might exist in the plugged-tube assumption. All scale is assumed to be gypsum.

5-90 Table 5-13 lists the scaling which occurred in effects XI and XII during two runs. The run from 6 February 1964 to 6 March 1964 is Development Run No. 1. Effects I, VI, and X did not scale during the run ending 17 January 1964. Effect XI was not equipped with weirs so results from effects XI and XII could be compared.

RUN PERIOD	OPERATING DAYS	TOTAL SCALE FORMED (LBS)		DAILY SCALE FORMATION RATE LBS/DAY		TUBES PLUGGED AT END OF RUN	
		XI	XII	XI	XII	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

TABLE 5-13. SCALING IN EFFECTS XI AND XII

5-91 The tabulation of Table 5-13 illustrates the effectiveness of better brine distribution. The results from effect XI are interpreted as an indication that the rate of scale formation is a function of the amount of existing scale. This is logical since scale forms from the concentration resulting from evaporation. Existing scale reduces heat transfer and, therefore, reduces evaporation, concentration, and scale formation. The reduction of total flow area increases the fluid flow rate and reduces scale formation. In the case of the run performed from 6 February to 6 March 1964, the tube-flow area available in effect XI at the beginning of the run was 10.68 SQ-FT and at the end of the run was only 8.28 SQ-FT.

5-92 Scaling occurs first in the high horizontal brine velocity areas of the distributor plate and last in the backwater areas. The use of the distribution weirs overcame this problem. It is almost certain that starving of tubes is a principal factor in promoting scaling. Improvement of the inlet and distribution configuration when combined with a more adequately devised, recirculation scheme could possibly eliminate the formation of scale in effects XI and XII. High distributor plate horizontal velocity results from single-point inlet and the rather primitive deflector plate.

5-93 The present recirculation scheme introduces brine to the top of the heating element. This brine has already flashed once in the evaporator at the heating element tube exit pressure. This brine is more concentrated than the brine from the previous effect which is introduced and mixed with the recirculation stream at the top of the heating element. The installation of a blinded 8-inch nozzle on each inlet water box would help to revise this system. This revision is anticipated.

5-94 Scale removal by circulation of sea water was also quantitatively evaluated by the same measurement scheme as the initial scale measurements. All scale was removed from effect XII and 11,033 pounds of scale were removed from effect XI in the first 31 hours of circulation at an average removal rate of 356 LBS/HR. After a total of 57 hours (an additional 26 hours) of circulation, only the scale in the six; tightly-plugged tubes remained. The average rate of scale removal for the last 26 hours was 30 LBS/HR, based on the doubtful assumption that total elapsed time was required to remove the scale. The six plugged tubes were not mechanically cleaned like the initial effect XI tubes were during Development Run No. 2.

5-95 <u>Two-Phase Flow Pressure Drop-Evaporators.</u>

5-96 This item is very important to future design. It has been difficult to determine methods by which actual pressures can be measured in the following locations:

a. Above the distributor plate.

b. Between the distributor plate and the liquid level on the top tube sheet.

c. In the evaporator tubes at the tube exit.

Development work will continue on this problem and the associated temperature measurement problem, since solutions to these problems will be very valuable to future design. Dukler[6] has explained the important effect it has on heat transfer.

[6]Prengle, Dukler and Crump, Inc., <u>Saline Water Research and Development Progress Report No. 64,</u> June 1963.

5-97 Scale Prevention.

5-98 In accordance with the development plan, a method was devised for the addition of a polyphosphate compound to the saline water stream. This was accomplished during Development Run No. 2. It was essential to maintain process conditions as nearly constant as possible, at the same values as the previous run, in order to evaluate the effect of polyphosphate addition in the suction of P-18. This pump transfers brine from the cone of effect VIII to the heating element of effect IX. This point was chosen primarily because the brine from this point on does not exceed 170°F. Reversion of polyphosphates to ortho-phosphate is slow at this low temperature and residence time in the evaporator is short enough to eliminate significant reversion which could form phosphate scale. The polyphosphate compound was furnished by Nalco and is a proprietary blend of polyphosphates. In addition to the reason stated above, the suction of P-18 was selected as the specific point of addition because the stream of brine is under a vacuum, therefore, no injection pump is required to achieve addition or to control rate of addition.

5-99 The control limits were established as follows:

a.	Inlet Sea Water Concentration Factor (at P-3 discharge)	1.0 ± 0.1
b.	Brine Discharge Concentration Factor (at P-22 discharge)	3.0 ± 0.1
c.	Pressure in 312 Condenser (PSIA)	1.43
d.	Temperature of Steam to Effect No. 1 (°F)	240
e.	Production rate (GAL/day total)	1,000,000
f.	Polyphosphate Addition (PPM to SWF $\pm$ 1 PPM)	4

5-100 No scale was formed during Development Run No. 2; actually, old scale was removed. Development Report No. 1 indicated two plugged tubes in effect XII which were still plugged on startup of Development Run No. 2. Likewise, Development Report No. 1 illustrated six plugged tubes in effect XI which were still plugged on startup of Development Run No. 2. All but two of the aforementioned eight tubes were clean at the end of Run No. 2. This reflects a removal of at least 92.5 LBS of scale in the period from March 3, 1964 to April 13, 1964. It is probable that (existent) old patches of scale and thin scale film were removed, especially from effect XI. The amount of chemical added was 4 PPM based on SWF to the evaporator. The calcium ion is complexed (chelated) with the polyphosphates and the polyphosphates usually have a greater effect than that expected from the stoichiometric relationships such as:

$$3\,Ca^{++} + 2\,Na_3\,Po_4 \qquad\qquad Ca_3\,(Po_4)_2 + 6\,Na^{++}$$

5-101 The combining of the calcium ion in a soluble compound reduced the concentration of $CaSO_4$. This is only a partial explanation. The results achieved indicate that a lower concentration of polyphosphates will be adequate. The cost of this treatment at 4 PPM was only 0.92 cents per 1000 gallons and may be reduced to the range of 0.5 cents per 1000 gallons of water produced. The use of polyphosphate will be repeated under more severe conditions for further evaluation.

5-102 <u>Silt Removal.</u>

5-103 The clarifier-thickener system was originally designed for use as a part of a scale seeding system. Since this proved unsuccessful, the system has been idle for a considerable period of time. Modifications to the system for use as a silt settling tank were completed in May of 1964. During Development Run No. 3, the feed flow passed through this modified system. Unfortunately sea water turbidities were extremely light and variable during this period. The residence time and rapid light variation in inlet turbidity completely masked results other than to prove that flow could proceed in this manner. One unexpected benefit was the improved operation of P-3 and the deaerator pH control. This was a direct result of the constant suction head applied to P-3 by the overflow from the clarifier-thickener.

5-104 <u>Steam Jet Evactors.</u>

5-105 The Steam Jet Evactors are incapable of maintaining sufficient vacuum to continuously operate the Plant at design rates. The evactors are normally used only to supplement the vacuum pump which is, at best, marginal in its capacity. It is not presently known if this lack of capacity is due to design or deterioration. This should be determined and the evactors repaired if necessary. Since this unit also acts as vacuum pump standby, it performs a very important function.

5-106 <u>Inlet and Sea Water Feed System.</u>

5-107 It is believed that considerable and frequent amounts of recirculation are experienced between the warm condenser water discharge line and the sea water intake line. Data were taken and on two occasions, it was found that the proper return valve was not fully closed. This proved to be the only identifiable source of recirculation and recirculation ceased to exist when the valve was closed.

5-108 Condenser Cooling Water Flow.

5-109 Since overboard brine has been routed to another flume and does not pass through the hot well, there is no detrimental recycling of high concentration brine. It is an established practice to purposely recycle some warm brine to prevent freezing during extremely cold weather. Under four-pass conditions (refer to paragraph 5-182), the flow through the condenser is slightly under 4,000 GPM. Pressure conditions on the steam-side of 312 are adversely affected by warm condenser water; therefore, recycling should be avoided under almost all conditions. An exception is made when the absolute pressure in 312, and its reflection to the vapor head of effect XII, is so low as to induce excessive carryover of suspended brine droplets. Very small changes in pressure have a very significant effect on the over-all Plant Delta T.

5-110 Pumping Capacity.

5-111 It is believed that pumping capacity may be a limiting factor to production at useable extraction ratios. The flows required by the Heat and Material Balance are listed in Table 5-14 Flow Rates and Pump Capacities. This table also lists the pump capacity ratings at the estimated head requirements.

5-112 In general, the pumping capacities are barely adequate. Some pumps in particular are presently at the overload point. Obviously it is necessary to run both P-1 and P-2 in parallel to supply the requirement of feed to the evaporator barometric condenser and the 312 condenser. With both pumps operating, capacity is adequate for requirements up to approximately 6500 GPM. Pump P-5 is near its capacity, but has 15 percent to spare for the present operation. Pumps P-11, P-17, P-21 and P-23 must be in perfect condition and adjustment to maintain the load. Pump P-31 operates significantly below design head and can maintain high requirement. Pump P-54 is overloaded. The condensate pumps P-38 through P-43 are capable of a 10 percent increase, if the flash tanks are not used. Pump P-41 would not be capable of handling the condensate from effects VIII, IX, and X, if flash tanks were utilized. Pump P-44 has ample capacity.

5-113 It is necessary to point out that all inter-effect brine and condensate pumps are assumed to have maximum impellers with respect to the drive motors. This was not the original design but has been the

FLOW RATES & PUMP CAPACITIES					PUMP RATINGS				
PUMP	SERVICE	FLOW PER HEAT & MATERIAL BALANCE			HP	GPM	TDH	RPM	REMARKS
		LB/HR	SP GR	GPM					
P-1 & P-2	Raw S. W. Supply to 312	2,617,500	1.025	5107	50	4000	22	1750	Each pump capable of rating
P-3	SWF to DA	503,100	1.02	986	40	1700	55	1750	Each pump capable of rating
P-5	SWF to EVAP	503,100	1.02	986	125	1100	275	1150	Ratings at Max. Impeller for 10 HP
P-11	Brine, 1 to 2	472,305	1.0	945	15	1000	45	1150	Ratings at Max. Impeller for 10 HP
P-12	Brine, 2 to 3	443,400	1.0	887	15	1000	45	1150	Ratings at Max. Impeller for 10 HP
P-13	Brine, 3 to 4	414,500	1.01	821	15	1000	45	1150	Ratings at Max. Impeller for 10 HP
P-14	Brine, 4 to 5	388,500	1.01	769	15	1000	45	1150	Ratings at Max. Impeller for 10 HP
P-15	Brine, 5 to 6	361,500	1.02	709	15	1000	45	1150	Ratings at Max. Impeller for 10 HP
P-16	Brine, 6 to 7	333,000	1.02	653	15	1000	45	1150	Ratings at Max. Impeller for 10 HP
P-17	Brine, 7 to 8	306,000	1.03	594	10	650	45	1150	Ratings at Max. Impeller for 10 HP
P-18	Brine, 8 to 9	277,000	1.04	533	10	650	45	1150	Ratings at Max. Impeller for 10 HP
P-19	Brine, 9 to 10	248,500	1.04	478	10	650	45	1150	Ratings at Max. Impeller for 10 HP
P-20	Out of Serv.	-	-	-	10	650	45	1150	*
P-21	Brine Recycle in 11	420,000	1.05	800	10	800	36	1150	-
P-22	Brine 12 to waste	167,300	1.067	314	10	500	55	1150	-
P-23	Recycle in 12	424,000	1.06	800	15	800	40	1150	-
P-31	Cond. from 1	32,207	.953	67.3	15	50	175	1750	Orig. desig. p. 47
P-54	Cond. from 1 thru VII	199,579	.973	410.1	10	400	60	1150	-
P-38	Cond. from 8	29,021	.973	59.6	3	65	70	1150	-
P-39	Cond. from 9	29,000	.976	59.4	3	65	70	1150	-
P-40	Cond. from 10	28,500	.980	58.2	3	65	70	1150	-
P-41	Cond. from 11	28,000	.983	57.0	3	65	70	1150	-
P-42	Cond. from 311 & 12	26,520	.987	53.7	3	65	70	1150	Sometimes not used
P-43	Cond. from 312	23,880	.991	48.2	3	65	70	1150	Sometimes not used
P-44	Cond. from 12 & 312	50,400	.990	101.8	10	400	60	1150	Spares both P-42 & P-43
P-40	Dow Product	242,380	.995	487.2	75	750	250**	1150	
P-51	Freeport Product	86,565	1.0	173.1	50	450	200	3550	
P-4	Spare SWF to DA	-	-	-	100	1050	265	1750	

*May be returned to Service
**Estimated

TABLE 5-14 FLOW RATES AND PUMP CAPACITIES practice during replacement. All of these pumps (P-11 through P-23, P-38 through P-44 and P-54) could be equipped for operation at 1750 RPM instead of 1150 RPM. This would significantly increase both head and capacity but would aggravate cavitation problems.

5-114 Actual operating experience has indicated that pump P-1 and/or P-17 are limiting the ability to pump brine. This was proven during Development Run No. 4 at a final concentration factor of 2.5 ($\pm$). The maximum feed rate achieved was slightly over 500,000 LBS/HR. Pump P-54 has shown signs of being overloaded during runs where the extraction ratio exceeds 0.70. During Development Run No. 5, the limitations of pumps P-38, P-39, P-40, and P-41 will be tested at an extraction ratio of approximately 0.72 and a final concentration factor of 3.5. A full evaluation of pumping requirements for a 700,000 LB/HR feed rate and a final concentration factor of 3.5 is now underway.

5-115 PROCESS LIMITATIONS.

5-116 The matter of process limitations is one of considerable interest to the Office of Saline Water and eventually the population of the world. The limitations, set forth below, recognize process limitations with regard to scaling, heat transfer and the cost of equipment, but do not recognize the obvious mechanical limitations of the Freeport Demonstration Plant No. 1. They represent the best estimates based on the Plant development work of this past year, however, future improvements may further reduce these limitations.

5-117 Over-All Temperature Difference.

5-118 The over-all temperature difference is limited to 170°F. This can be achieved by a first effect temperature of 270°F on the steam-side of the first effect condenser, a temperature of 100° for the brine overboard, and will require a condenser capable of condensing steam for heat rejection at 97°F. It is felt that the added cost of such a condenser and the larger final effect, will be easily recovered in improved thermal economies. This temperature differential would allow the use of 20 effects with a resulting economy ratio of approximately 20 to 1 gross and 19 to 1 net.

5-119 Final Concentration Factor.

5-120 The final concentration factor is limited to 3.5 when using a polyphosphate treatment to control scaling in the effects operating at 3.0 concentration factor and above. The extraction ratio for normal sea water would be 0.714; or, for each pound of water extracted, it is necessary to feed only 1.4 LB of sea water. Further concentration would require chemical ion exchange resins, and equipment for removing calcium ion from the sea water prior to feed to the evaporator. A 4.0 concentration factor would still require a feed of 1.33 LBS of sea water per pound of water produced, which is a negligible improvement.

5-121 Heat Flux.

5-122 Heat transfer coefficients can be improved to the 800 to 1000 BTU/HR ^{o}F range by controlling flow rates, distribution, and tube size. The heat flux, however, should be kept below a 10,000 BTU/HR SQ-FT level to avoid disrupting the film with nucleate boiling.

5-123 Demonstration Plant No. 1 can partially overcome all of these limitations. The program for the ensuing year will attempt to achieve improved operating conditions despite the restrictions imposed by time and equipment. Much valuable development type information is still to be obtained from the existing plant. So much, in fact, that it is doubtful that the time left in the originally contemplated 5-year period will be sufficient.

5-124 DEAERATOR.

5-125 The deaerator is more properly designated as a decarbonator-deaerator. Its function is not only to remove dissolved air (i.e. Oxygen and CO_2), but also to promote the decomposition of carbonate and remove the additionally released CO_2. The system must, therefore, include the injection of H_2SO_4 and a subsequent ability to adjust the final pH so as to avoid corrosion downstream.

5-126 During the life of the Plant, efforts have been made to improve performance. The performance is evaluated by measuring the effluent concentration of carbonate alkalinity, and dissolved Oxygen. From this standpoint, the Freeport deaerator has become quite successful with effluent carbonate alkalinities of 7 PPM or less and dissolved O_2 levels consistently less than 100 PPB. However, it is necessary to use considerable quantities of steam (up to 3000 LB/HR) and chemicals (H_2SO_4 and NaOH). In an effort to improve the economy of deaeration, a program was undertaken to attain information about the actual processes taking place.

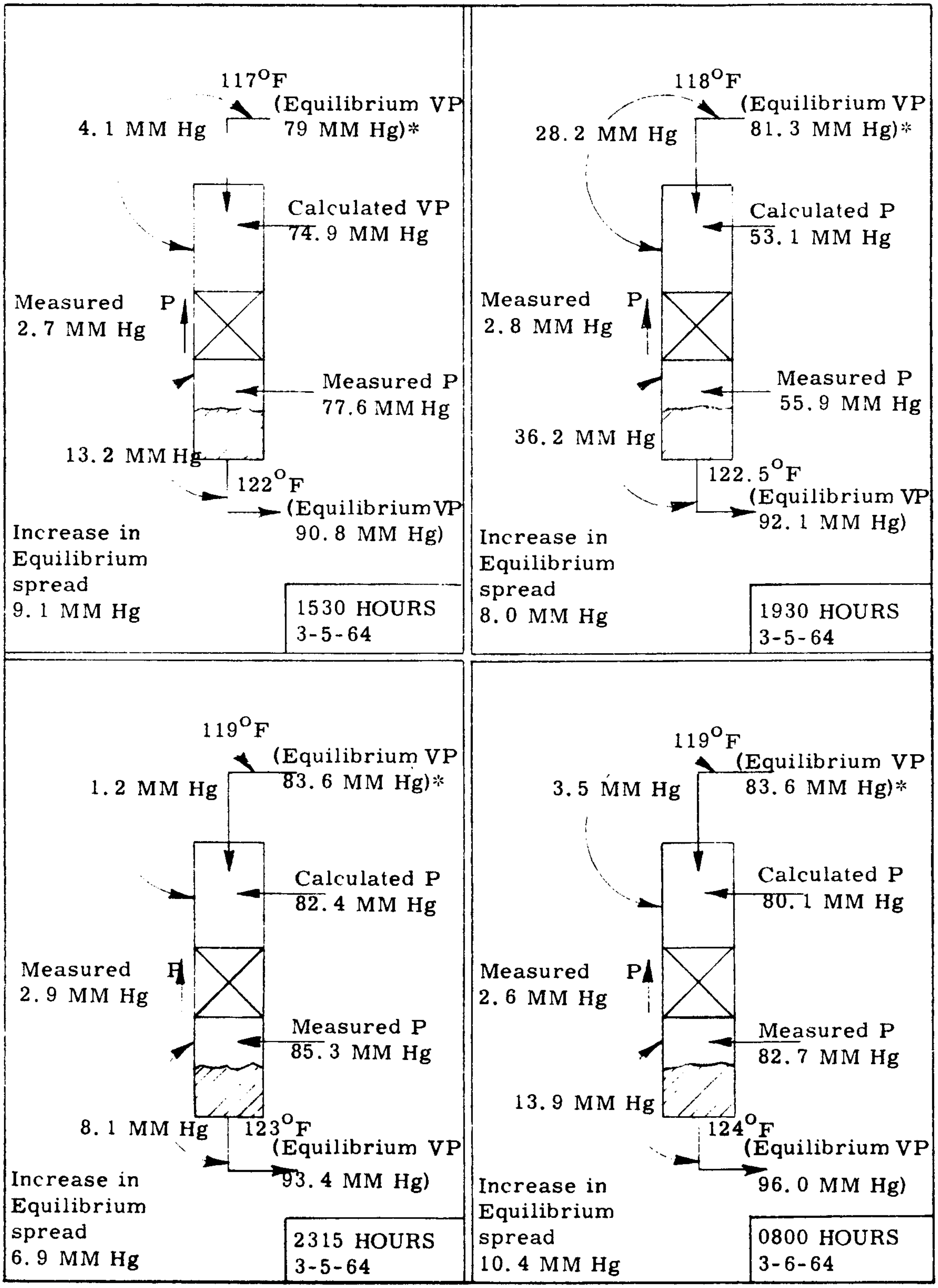
5-127 The first tests were conducted during Development Run No. 1. A preliminary discussion of the deaerator performance was issued to supply M. W. Kellogg with field performance data. The deaerator performance data are included in Table 5-15. Figure 5-6 presents

DATE	3-5-64			3-6-64
TIME	3:30 PM	7:30 PM	11:15 PM	8:00 AM
DESCRIPTION				
S. W. Feed Temperature after acid - ^{o} F	117	118	119	119
Steam from No. 11 ^{o}F	126	126	126	128
Steam from No. 2 & 3 ^{o}F	182	181	181	182
Sea Water Out after NaOH	122	123	123	124
Steam from 2 & 3 pressure psia	7.45	7.42	7.71	7.51
Steam from 2 & 3 pressure psia	1.90	1.70	1.70	1.75
S. W. Feed rate M LBS/Hour	487	487	487	487
Steam from No. 11 LBS/Hour	2115	2105	2095	2105
*Steam from No. 2 & 3 LBS/Hour	1022	997	1007	1007
Gas from Deaerator LBS/Hour	940	940	1020	995
pH of Sea Water after acid	4.1	3.65	4.5	3.80
pH of Sea Water after D. A.	4.0	4.0	4.0	4.05
pH of Sea Water after Caustic	7.3	7.05	7.1	7.05
Pressure in D. A. below packing psia	1.50	1.08	1.65	1.60
Pressure drop across packing " H_2O	1.45	1.50	1.55	1.40
**Sea Water outlet ppb O_2	200	190	210	180
**Sea Water Outlet ppm CO_2	10	8	10	7

*Upstream pressure not measured so this value is a figure resulting from the assumption of an upstream pressure that seems reasonable.

**Sea Water outlet analyses for O_2 & CO_2 are confused by the existence of an air leak to the system during the test.

TABLE 5-15 DEAERATOR PERFORMANCE DATA



*UCLA Dept. of Engineering Report 62-53.

FIGURE 5-6. DEAERATOR PRESSURE CONDITIONS DURING DEVELOPMENT RUNS OF 5 AND 6 MARCH 1964 an analysis of pressure conditions. Since it was not possible to attain gas stream composition data (from the system which is below atmospheric pressure) with equipment on hand, these data are missing. Heat and material balances made on the data were satisfactory, and therefore, some conclusions can be drawn with confidence.

5-128 The total pressure above the packing is lower than the equilibrium vapor pressure of sea water at the point of injection. This is as it should be to promote the rapid removal of CO_2 saturated vapor. The packing should ideally decrease the difference between equilibrium vapor pressure and actual vapor pressure, by promoting the intimate contact of rising vapors and falling liquid. The inability to do so is a strong indication of channeling and/or bypassing. This indication exists in all four sets of compiled data (refer Table 5-15). The difference between equilibrium vapor pressure and actual total vapor pressure increases a minimum amount of 6.9 millimeters of mercury as measured above and below the packing. For efficient stripping, there should be a decrease.

5-129 The apparently, excessive rise of liquid temperature is probably necessary to accomplish adequate removal of CO_2 and O_2. The deaerator unit does achieve a satisfactory level of these two constituents in the sea water effluent. The steam introduced from effects II and III probably promotes heating (since it often contains significant quantities of free CO_2) and inhibits the normal stripping process. The steam requirement of slightly over 3000 LBS/HR is excessive and results from one or more of the previously mentioned problems.

5-130 Vacuum sampling equipment was obtained for the sampling of vapor streams during tests of Development Run No. 2. The deaerator performance data obtained during this development run are presented in Tables 5-16 and 5-17. The pressure conditions presented therein, do not closely correspond with those of the previous run. It is necessary to make a selection of times for full data coverage of the deaerator, since all data were not taken simultaneously (due to lack of equipment and manpower). The data utilized in the determination of steam composition from effect XI were those taken on April 12, 1964, from 0800 to 0900 hours. The data utilized in the determination of vapor composition leaving the deaerator were those taken on April 12, 1964, from 1300 to 1400 hours. The liquid data utilized for the heat and material balance were those taken on April 11, 1964, at 1330 hours.

Stream & Location of Measurement	Date	Time	Temp °F	Flow LB/HR	Pressures			Composition					
					Measured		Calc. Equil. mmHg	H_2O		CO_2		Air	
					psia	mmHg		Vol. %	Wt. %	Vol. %	Wt. %	Vol. %	Wt. %
Steam From 2 & 3 (After Orifice)	4-11	0930	196	932	10.0	517	-----	NOT RUN					
	4-11	1030	196	927	10.0	517	-----						
	4-11	1130	196	926	9.8	506.7	-----						
	4-11	1230	196	921	9.8	506.7	-----						
	4-11	1330	196	921	9.8	506.7	-----						
Steam From 11 (before flow tube)	4-11	0930	128	1550	1.9	98.1	-----						
	4-11	1030	128	1540	1.9	98.1	-----						
	4-11	1130	128	1780	1.9	98.1	-----						
	4-11	1230	128	1580	1.9	98.1	-----						
	4-11	1330	128	1590	1.9	98.1	-----	99.995	99.987	.005	.013	0	0
Vapors From Deaerator (above packing)	4-12	1300	120	1040	1.68	87	86*	99.52	98.941	.36	.865	.13	.194
	4-12	1500	121	1040	1.68	87	89*	99.86	99.773	.01	.021	.13	.206
	4-13	0700	120	1040	1.68	87	86*	99.60	99.121	.27	.669	.13	.210
	4-13	0930	120	1030	1.68	87	86*	99.76	99.513	.11	.279	.13	.208
Vapors Below Packing	4-11	0930	122		1.72	89	92*	NOT RUN					
	4-11	1030	122		1.72	89	92*						
	4-11	1130	122		1.72	89	92*						
	4-11	1230	124		1.72	89	97*						
	4-11	1330	124		1.72	89	97*						

NOTE: Temp. Accuracy $\pm 1/2\,^{\circ}F$

Press. Accuracy $\pm 1/10$ "H_2O

*from UCLA Report No. 61-80 (Ref. 7)

TABLE 5-16 DEAERATOR OPERATING DATA LIQUID STREAMS

Stream & Location of Measurement	Date	Time	Temp °F	Flow LB/HR	Composition: Conc. Factor	Oz ppb	Alkalinity ppmas Ca Co$_3$	PH	Calculated Equilibrium Vapor Press mmHg	Measured Vapor Pressure mmHg	Δ P* mmHg
Sea Water	4-11	0930	120	480M	.80			3.8	86.4	87	-.6
Inlet After	4-11	1030	120	480M	.80			3.9	86.4	87	-.6
Acid Addition	4-11	1130	120	480M	.80			3.9	86.4	87	-.6
& Before Flash.	4-11	1230	121	480M	.80			3.85	89.0	87	+2
	4-11	1330	121	480M	.80			----	89.0	87	+2
Sea Water	4-11	0930	122	480M	.80	nil	7	3.9	92	89	+3
Effluent before	4-11	1030	122	480M	.80	nil	7	4.9	92	89	+3
Caustic	4-11	1130	122	480M	.80	90	7	4.05	92	89	+3
Addition	4-11	1230	124	480M	.80	80	7	4.10	97.2	89	+8.2
	4-11	1330	124	480M	.80	80	7	4.0	97.2	89	+8.2

NOTES:

1. Conc. factor was taken from daily operator logs and, therefore, does not show any variation.
2. Calculated equilibrium vapor pressures are taken from UCLA Report No. 61-80 by Charles Gastaldo, and do not take into account the effect of acidulation.
3. The variation of calculated equilibrium vapor pressure with temperature is quite rapid over the range measured, being 86.4 mmHg at 120°F and 97.2 mmHg at 124°F, so fractional variations of temperature not recorded would probably correct the negative differential at inlet conditions.

*Δ P Equilibrium VP - Actual V.P.

TABLE 5-17 DEAERATOR OPERATING DATA VAPOR STREAMS

5-131 The unit heat balance calculations are presented below (see Figure 5-7):

HEAT AND MATERIAL BALANCE

$$(480,000)\ (85.323) + (1590)\ (1117.0) + (921)\ (1144.4) =$$

$$(481,476)\ (88.489) + (1035)\ (1114.4)$$

$$43,785,032 = 43,758,734$$

5-132 The calculations depict a very close heat balance; probably closer than the accuracy of the data used. The required use of concentration factor data from the operators' log contributes to situations such as an apparent negative differential between equilibrium vapor pressure and actual vapor pressure. It is, therefore, difficult to draw any conclusions from these data. The closeness of the two pressures would again indicate that heating (such as is actually occurring) is probably necessary to achieve good decarbonation and deaeration.

5-133 The vapor phase composition data are interesting and indicate some necessity for better sea water data. The weight-percent CO_2 data form the basis of the conclusion that the vapors from the deaerator contain more than 10 times as much CO_2 as the inlet steam from effect XI. Even more interesting is the possibility of developing k values for CO_2 in sea water. If this were possible, it would enable a more sophisticated and effective deaerator design.

$$\text{let } k = y/x = \frac{\text{mol fraction of } CO_2 \text{ in vapor}}{\text{mol fraction of } CO_2 \text{ in sea water}}$$

5-134 If a range of k values covering the proper pressures and temperatures are known, it is possible to design a deaerator that will attain equilibrium conditions above and below a packed section or a series of trays. The theoretical stage is one in which equilibrium is reached between the liquid and the vapor. Conditions are varied by heat addition, etc. to change the equilibrium and to gain the desired final compositions of the liquid and vapor. A y value equal to 0.0036 has already been calculated from the vapor stream data but x values are not readily available.

5-135 If these x values become available and valid over a range of temperatures, they would permit an accurate prediction of performance, based on design, to achieve the desired conditions. These interface concentrations are useful in packed or trayed column design.

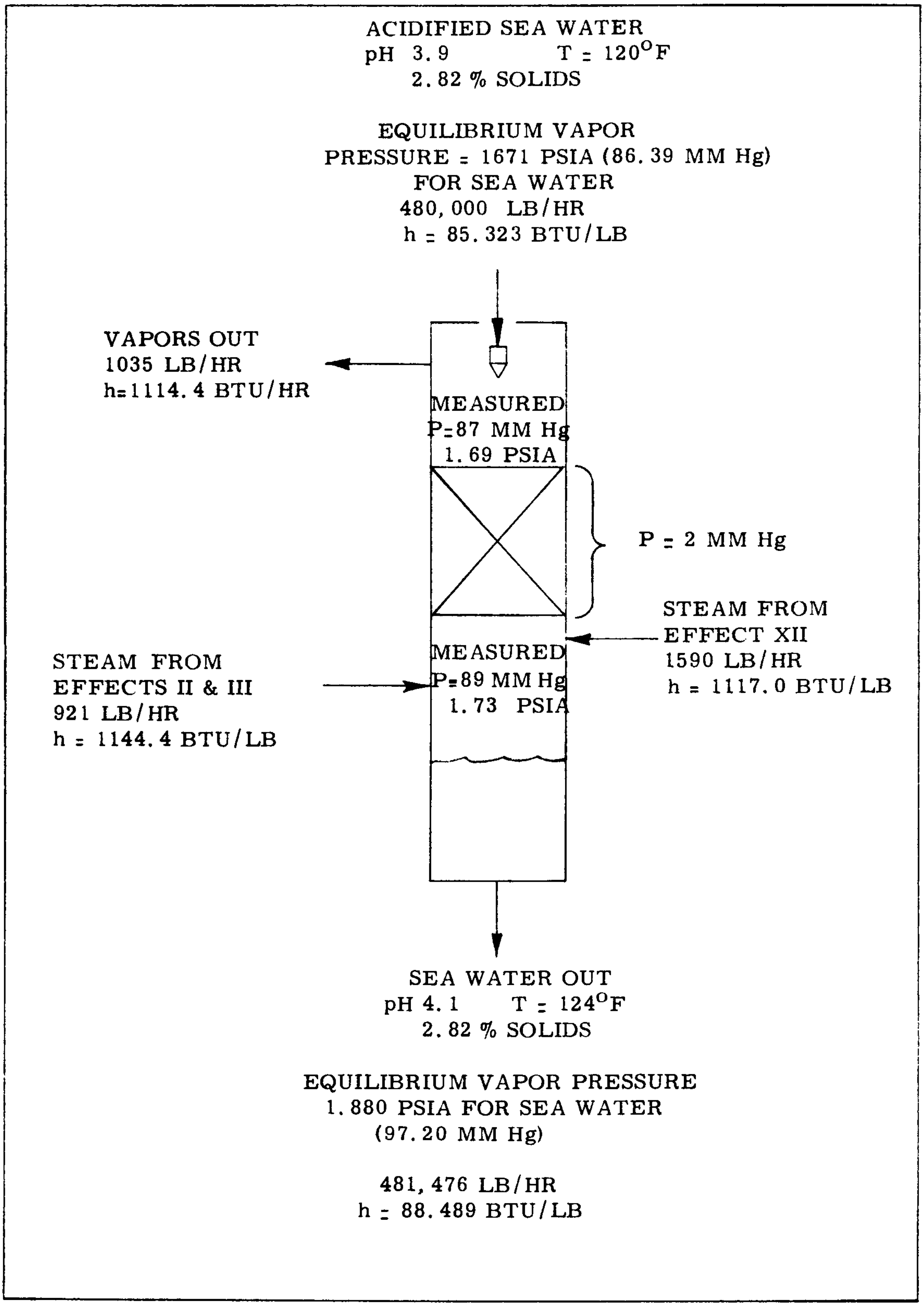


FIGURE 5-7. HEAT BALANCING SCHEMATIC FLOW DIAGRAM

5-136 The partial pressure calculations are summarized as follows:

COMPONENT	VOL. %	PARTIAL PRESSURE
H_2O	99.52	86.59 MM H_g
CO_2	.36	.31 MM H_g
Air	.12	.10 MM H_g
	100.00	87.00 MM H_g

These pressure calculations represent the condition above the packing. If equilibrium partial pressures were known, the value and/or probability of stripping with effect XI steam could be more reliably estimated. This could also be used to determine whether heating should be accepted as desirable. All of these comments are based on fragmentary data, and therefore, the figures should not be used for anything beyond comparisons to future data. It is hoped that studies currently under way by other OSW contractors will produce such information.

5-137 At the recommendation of the Dow Chemical Co., the packing was removed from the deaerator for Development Run No. 4. Also, acid addition was decreased and an attempt was made to maintain a pH of 4.1 at the sea water inlet to the deaerator. Results achieved here were surprisingly good. The Oxygen content was maintained below 100 PPB and carbonate alkalinity to maximum of 12 PPM in the effluent. It is further noted that steam from No. 2 and 3 vents was not used for this run, thereby reducing steam consumption to approximately 2000 LBS/HR.

5-138 The question at the moment is, what is the detrimental effect of the additional carbonate alkalinity under the conditions of Development Run No. 4? Run No. 4 was conducted at first effect temperatures of less than 230°F, so it cannot be considered a fair test in determining whether scaling in the first effect could result from this level of carbonate. A lower level (7 PPM) will be required to prevent scale formation at 270°F first effect temperatures. Also, the presence of any significant amount of free CO_2 in the steam from effect I will require excessive venting from effects II and III. Accordingly, a new type of packing (Dow Maspac) was ordered and will be evaluated in the near future.

5-139 PRODUCT WATER TREATING.

5-140 To reduce the corrosivity of the product water, a system was devised to inject Na_2CO_2 and $CaCl_2$ into the Freeport water, resulting in a final $CaCO_3$ concentration of approximately 10 PPM. In addition to this step, a complete converted water stabilization test program was proposed and approved. The program is presented below.

5-141 Converted Water Treatment Test Program.

5-142 A converted water treatment test program has been instituted at the Freeport Demonstration Plant to develop a product water which, when used in average municipal water systems, is sufficiently stable to be acceptable. The effect of various treatment methods on the degree and rate of corrosion of construction materials is being analyzed and recommendations for permanent product water treatment will be made.

5-143 Water Treatment Methods to be Tested. Several methods of water treatment are available to decrease the corrosiveness of converted water to materials used in the construction of municipal water distribution systems. These methods may be categorized as follows:

a. Increase calcium hardness.

b. Increase carbonate and bicarbonate alkalinity.

c. Stabilize iron and water.

5-144 Table 5-18 presents the effect (on construction materials) and the water treatment methods that are being evaluated.

5-145 The treatments presented in Table 5-18 are added to converted product water upstream of test loops containing corrosion coupons made of the various construction materials to be tested. Untreated, converted water will also be tested to serve as a reference point for comparison of all results.

	TREATMENT	CHEMICALS ADDED	CONTROL CONC.	TYPE OF ACTION*
(1)	Sodium Carbonate-Calcium Chloride	$Na_2CO_3 + CaCl_2$	10-15 PPM as $CaCO_3$	A, B
(2)	Calcium Carbonate Filter	$CaCO_3$ as oyster shell	5-10 PPM as $CaCO_3$	A, B
(3)	Polyphosphate	Nalco 918 or equal	5-10 PPM	C
(4)	Sodium-Silicate	2:1 SiO_2 + Na_2O	5-10 PPM	C
(5)	$CaCO_3$ Slurry	$CaCO_3$ (pigment grade)	5-10 PPM as $CaCO_3$	A
(6)	Recarbonation	$Ca(OH)_2 + CO_2$	35-40 PPM as $CaCO_3$	B

*A Increase calcium hardness
B Increase carbonate and bicarbonate alkalinity
C Stabilize iron and water

TABLE 5-18. SCHEDULED WATER TREATMENT DEVELOPMENT PROGRAMS.

5-146 Construction Materials to be Tested. The following construction materials, commonly used in municipal water distribution system components, are tested in each water treatment:

a. Carbon Steel

b. Galvanized Steel

c. Cast Iron

d. Copper

e. Concrete

f. Polyvinylchloride

g. Transite

h. Ni-Cr Alloy (comparable to Ni-Resist Type 1-b)

i. Steam Bronze ASTM-B62

j. Cast Si-Bronze ASTM-B198-13b

5-147 Materials a through g listed above are normally used in municipal water systems as mains and distribution piping. Materials h through j are used in the construction of pumps and valves for water systems.

5-148 Installation of Materials to be Tested. Test loops of one-inch pipe have been fabricated in which corrosion coupons of each material to be tested is installed. The test coupons are mounted in plastic holders to electrically insulate the coupon; and the holders are inserted into the test loops, parallel to the flow of water (see Figure 5-8). Treated water is passed through the test loops at a controlled rate of approximately three feet per second (8 gallons per minute). Coupons are exposed to the water for a period of thirty days. One test loop flows untreated water to serve as a standard for comparison of the effects of treated water upon the corrosion coupons.

5-149 Test Loop Configuration. Figure 5-9 defines the installation and equipment necessary to conduct these test-runs. The test system includes four test loops. Three loops pass treated water and one loop passes untreated water. Two 30-day test runs are required to analyze the six water treatment methods to be tested.

5-150 Evaluation of Test Results. The test coupons are examined and evaluated at the end of a thirty-day exposure period. The corrosion rate of each treated water on each material is evaluated qualitatively, for general etching, localized attack, and pitting attack; and evaluated quantitatively by calculation, based on the loss of weight of the test coupons. On the basis of this evaluation, recommendations will be made for the permanent treatment necessary to reduce the activity of converted water to average municipal water distribution systems.

5-151 Test Schedule. The procurement and installation of most materials and equipment required or the test program was completed during the third and fourth process development runs. The first test run is scheduled to be completed concurrently with Process Development

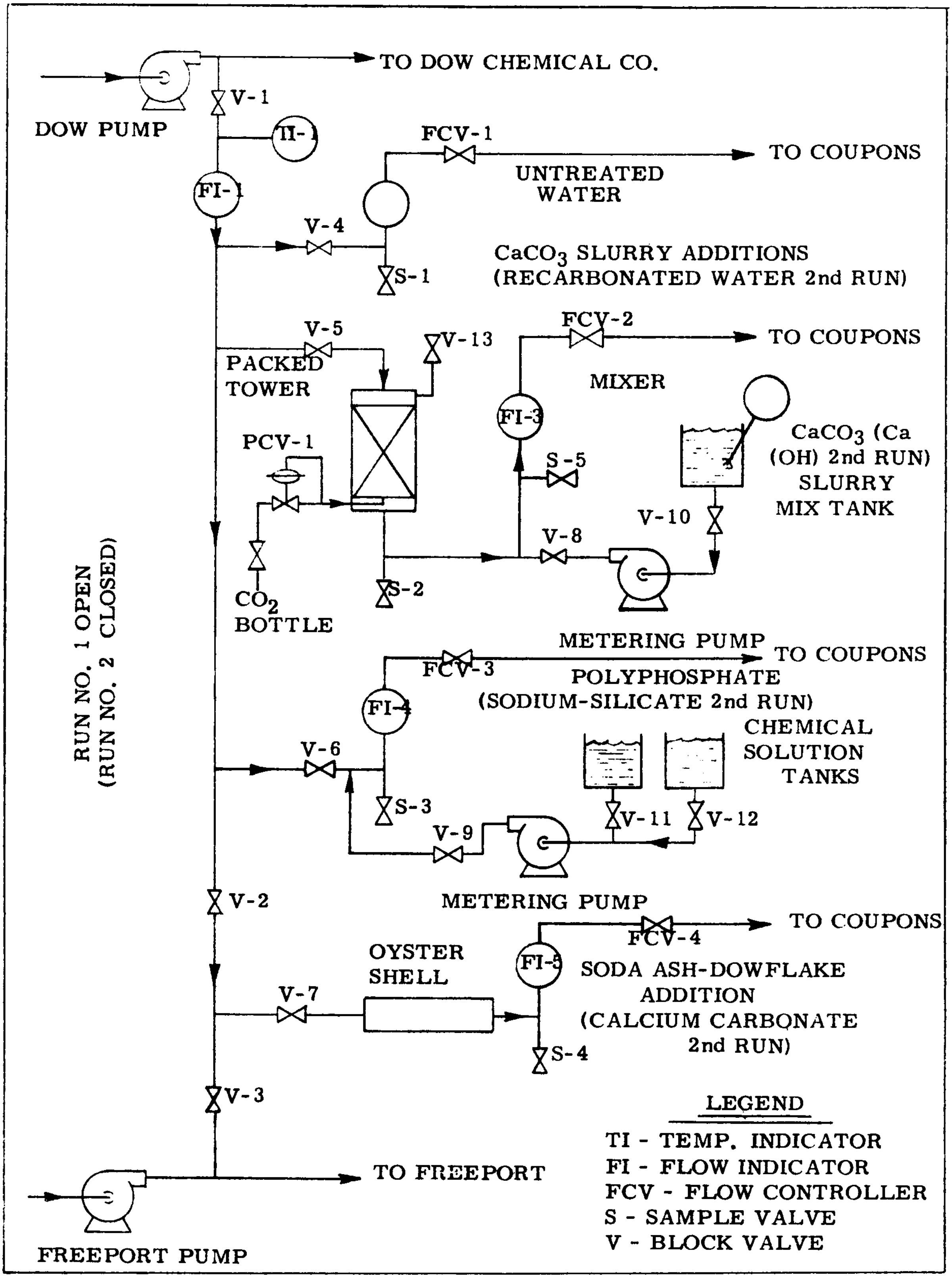


FIGURE 5-8. CHEMICAL ADDITION FLOW SHEET

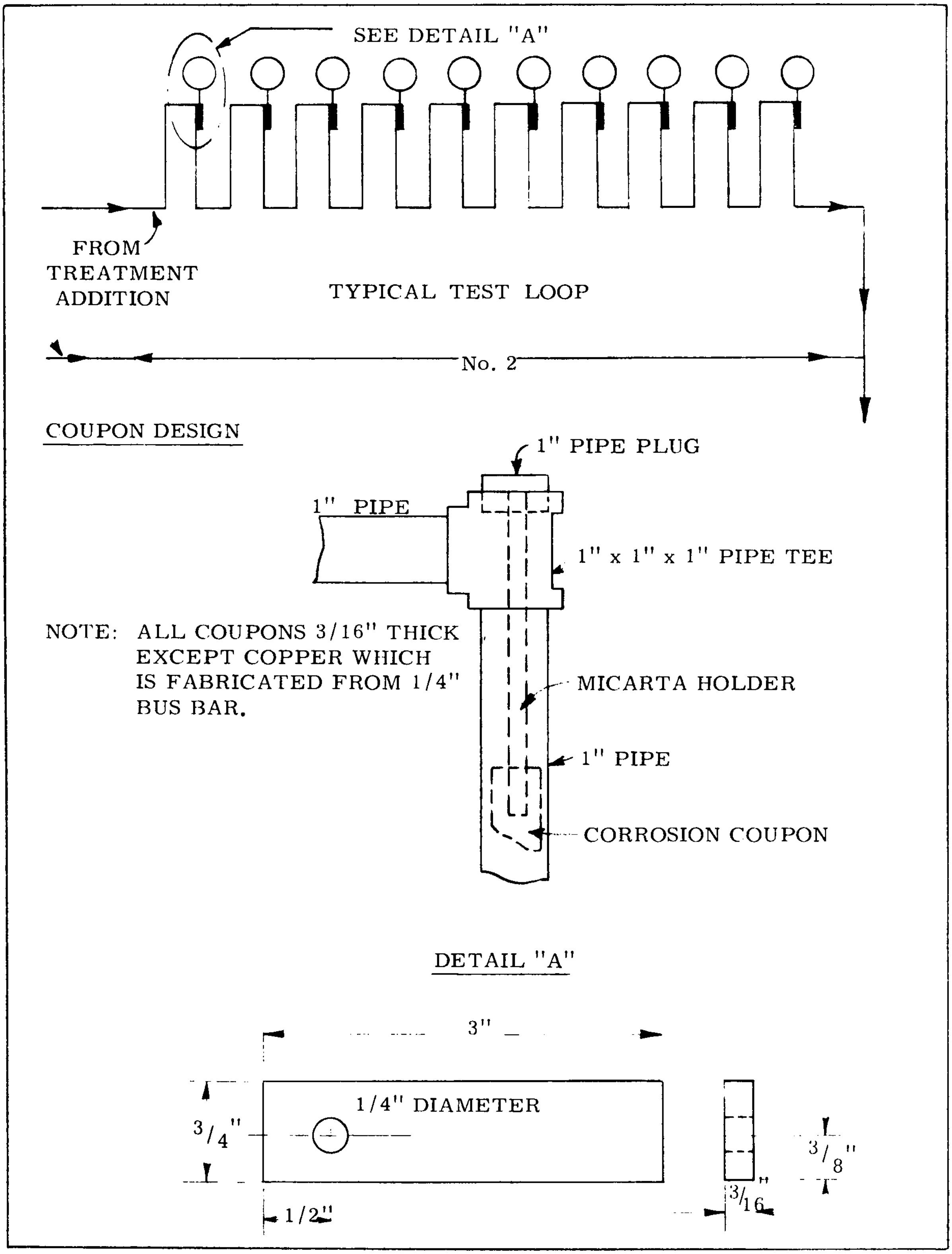


FIGURE 5-9. WATER TREATMENT STUDY - TEST SPOOL CONFIGURATIONS AND DETAILS

Run No. 5, beginning approximately 20 July 1964 and ending approximately 20 August 1964. The second test run will be completed concurrently with Process Development Run No. 6, beginning approximately 1 September 1964 and ending approximately 30 September 1964. Additional test runs will be scheduled as indicated by the results of first two test runs.

5-152 Preliminary tests run by Nalco Chemical Co., and Dow Chemical Co., indicate that the carbonate treatment does not substantially reduce the corrosivity. The mixture of converted, treated water and Freeport well water is only slightly more corrosive than well water alone. The problem to be solved here is not just a Freeport, Texas problem. Water converted by any type distillation process will be of a purity which will adversely affect the corrosion of municipal and residential water systems. The program is designed to acquire the necessary information to specify treatments and predict results for all receiving systems.

5-153 TECHNICAL EVALUATION.

5-154 EQUIPMENT EVALUATIONS.

5-155 It is difficult to separate the discussion of process and mechanical factors from equipment evaluations. Some conditions, however, are more properly associated with the various process and classes of equipment. (The reader is referred to Figure 7-1, Flow Diagram, Revision 2 to assist in understanding the following discussion.)

5-156 Intake.

5-157 The intake was evaluated for recirculation during Development Run No. 2 and no recirculation was found at that time. This configuration and arrangement would present problems for conditions other than flume conditions. It is suggested that a large part of the silt problem could be resolved in future plants by redesigning the intake system. The intake is in good physical condition.

5-158 Clarifier-Thickener.

5-159 The evaluation of the clarifier-thickener as a device for removal of silt is only partially complete. Certain mechanical problems involved with feed and withdrawal are only now being solved. The available head from pumps P-1 and P-2, makes it impossible to feed warm return condenser water to the clarifier-thickener. This situation is not likely to be remedied unless P-1 and P-2 are replaced. A modification to provide adequate suction volume for pump P-3, and still allow use of the laundering weir, is now in development. This modification must be completed before an evaluation can be made, concerning clarification with a coagulant. The clarifier-thickener unit is in good physical condition.

5-160 Deaerator.

5-161 The control of deaerator was reversed for Development Run No. 4. Until that time, the deaerator level was controlled by throttling the inlet stream. The flow rate control to the evaporator was located on the discharge from the deaerator. It was felt that this configuration allowed for many fluctuations in flow to the deaerator but caused difficulties in controlling acid injection for a specified pH. To accomplish the reversal, it was necessary to provide a new source of barometric condenser water which had previously been supplied from a connection downstream of the deaerator level control valve. The new connection was made close to the discharge from pump P-3. If this had not been done, the feed flow measurement would have included barometric condenser water. The results of this change were beneficial to pH control and chemical consumption.

5-162 The removal of the ceramic rashig rings from the deaerator, permitted an insepction of the internal lining. This inspection revealed damage and minor failure at ring contact points. Replacement of the coating is scheduled. The use of plastic packing is expected to decrease the tendency to damage the coating, but an inspection should be made in 18 months (the life of the present coating). The inspection also revealed that there was an accumulation of silt and debris in the bottom of the vessel. The level of this accumulation was at the bottom of the P-5a suction nozzle. This emphasizes the need for either removal of silt in the clarifier-thickener and/or a system for on-stream removal of the accumulation in this vessel. The chemical feed systems are in good condition. The deaerator does not experience flooding and has a capacity which substantially exceeds the requirements.

5-163 Pump P-5a.

5-164 The evaporator feed pump P-5a, a Worthington, 2-stage, vertical turbine pump, was placed in service in late 1963 to replace the double-suction, horizontally-split case, centrifugal pump, P-5. Pump P-4 continued to function as standby for pump P-5a. After six months continuous service, pump P-5a was removed from service because of excessive vibration caused by wear to the top bowl bearing. Continued pump operation was possible without repair; however, it was decided to modify as well as repair the pump since inspection revealed defects in the Ni-Resist bowls and the suction bell. Pump P-5a has not been returned to service because of the difficult in obtaining satisfactory Ni-Resist castings.

5-165 A meeting was held at the Worthington pump factory on 7 July 1964 for the purpose of inspecting the pump parts. The results of this meeting are presented below:

a. Top bowl bearing was worn sufficiently to cause the vibration which was the immediate reason for pump shutdown and removal.

b. Shaft sleeve was worn at the packing but wear was negligible at the stuffing box bearing. The snap ring grooves were determined to be satisfactory.

c. The impeller shaft was extensively worn at the top bowl bearing and worn to a lesser degree at the other three bearings.

d. The first-stage impeller displayed some signs of minor cavitation. The 316 stainless steel impeller did not show damage from this cavitation.

e. The line shaft displayed packing wear only.

f. The Ni-Resist bowls were both welded, but only one was cracked. The suction bell definitely showed graphitization in the ring fit area. Some question exists as to whether the suction bell was actually Ni-Resist as ordered.

g. The impeller-bowl assembly bearings, other than the top bearing, did not appear to be excessively worn. These graphitar bearings have been damaged by impact and will require replacement.

5-166 On the basis of previous experience and descriptions of operating problems, Worthington and/or Stearns-Rogers recommended modifications to evaluate other materials and mechanical features. The recommendations selected are expected to promote a longer service life. Since the presence of silt and even trace quantities of industrial wastes could invalidate the results of previous experience, a detailed analysis of flume sea water will be attempted in order to identify any objectional industrial waste.

5-167 The top bowl bearing wear is by far the most serious problem discovered. Worthington feels that operation at varying rates, along with the slight cavitation, is probably causing some slight up-and-down movement of the impeller shaft. This could cause some deflection in approximately 10 feet of line shaft, and induce whipping which would cause abnormal wear of the top bowl bearing. It is recommended that an intermediate line shaft bearing be installed. This would decrease vibration tendencies when clearances at the top bowl bearing increase due to normal wear. If whipping does exist, installation of the intermediate bearing will prolong the maintenance free operating period.

5-168 The inability to keep the stuffing box sleeve in place is another problem. The sleeve itself shows an appreciable amount of packing wear. The function of this monel sleeve is to allow for preferential wear of the sleeve rather than the shaft against the bearing and the packing. The sleeve can be eliminated by applying a very hard and smooth ceramic coating to the shaft. The stuffing box can be revised to eliminate the sleeve, snap rings, and set screw.

5-169 To return the pump to good operating condition, the bearing contact areas of both the impeller shaft and the line shaft must be resurfaced. Worthington recommends ceramic coating the shafts at all bearing locations and recommends installing new bearing at all locations. All bearings will be graphitar, except for the intermediate line shaft bearing which will be Teflon.

5-170 Repairs and modifications now in progress are summarized as follows:

a. The impeller shaft will be undercut and ceramic coated in the bearing areas.

b. The suction bell and both bowl castings will be replaced with new Ni-Resist castings by Worthington.

c. The line shaft will be modified to accept an intermediate bearing. This requires the column pipe to be modified to support the bearing housing.

d. The stuffing box is being modified to accept a ceramic-coated shaft and a new graphitar bearing.

5-171 These modifications will provide experience with ceramic coating against graphitar and packing; teflon against stainless steel; and the effect of the center line shaft bearing on top bowl bearing wear. The 316 stainless steel parts of the pump are in good condition, although slight cavitation exists at the first-stage impeller. The Ni-Resist bowl castings appear to be in good condition except for manufacturing defects.

5-172 Pump P-5a had a very beneficial effect on the process because it was capable of providing a smooth continuous flow to the evaporator. Apparently, pump P-5 was surging and required frequent maintenance (six week average service life). The packing position (high pressure side) on pump P-5a, greatly improved the quality of feed to the evaporators from the standpoint of dissolved oxygen. Because of this improvement, the use of sulfite was discontinued. Unfortunately, Development Runs Nos. 3 and 4 were adversely affected by the loss of pump P-5a.

5-173 Preheat Circuit.

5-174 The preheat circuit equipment is in good condition up to the point where the feed temperature reaches 200°F. From this point on, the carbon steel sections are experiencing frequent failure due to general thinning and some pitting-type corrosion. This is most evident in the piping spools between heat exchangers, but has recently become a problem in the sea water channels and baffles. Perforations have occurred, which has resulted in bypassing of the exchanger. Stainless steel liners have been installed in two of the 300-series exchangers in this area. It is possible that they will experience chloride stress corrosion and may require a second replacement with a cupro-nickel. There was a period during which the preheat circuit was operated without anodes. We have re-installed these anodes. The replacement of some metal which corroded badly in the past will aid in evaluating the effect of these anodes on corrosion. A study is contemplated on the use of anodes as related to this Plant, and sea water heat exchange equipment in general.

5-175 As previously noted, the use of heat exchanger 201 is marginal, and should it fail, we would seriously consider not replacing it. Its function would be easily replaced by the first effect flash tank. Likewise a method may be devised for elimination of the 208 through 211 exchangers which involves use of the latter stage flash tanks. It appears that heat exchangers 214 and 215 represent a lot of surface which could be better utilized. If it is not practical to put all product through both exchangers, a method might be devised which would replace 215 with 201 and allow use of 215 for exchanging heat between the feed and the blowdown from effect XII. These developments would provide valuable knowledge.

5-176 Evaporators.

5-177 The evaporators are difficult to modify. In accordance with previous discussions in this report, we would like to be able to vary the amount of surface in each evaporator. We also feel that much valuable knowledge would be achieved if additional brine distribution modifications could be accomplished in effects XI and XII. We also feel that mist elimination should be studied in effect XII. In addition to the conventional woven-wire mesh, pressure controlled bypassing of vapor is a possibility. It would not be logical to vary tube size and/or length in the present units, but this is a desirable future design possibility. In fact, many economies for future plants are indicated by designing new physical configurations and combinations into the evaporators. Development data of significant value, to aid in these future designs, will be obtained when it becomes possible to measure pressures and temperatures at critical local points in the present evaporators. The problem of recycling and transfer of brine around effects I, X, XI, and XII should also be given considerable study.

5-178 The corrosion problem affecting the hot end of the preheat circuit is also affecting the water box in the first two effects heating elements. The coating failure problem here is anticipated to be resolved in the coming year. Carbon-steel plugs, originally placed in the unused holes of evaporator tube sheets, are failing due to corrosion. This failure bypasses steam to brine and adversely affects the economy rather than product quality. We expect to replace all such plugs with aluminum-brass plugs during the next few months. The upper plugs are to be filed with an epoxy potting compound. This will avoid pocketing brine in these "cups" with subsequent concentration and possible corrosion.

5-179 The absence of insulation on effect XII causes performance to vary extensively with weather conditions. This is especially true during rainfall and cold wind.

5-180 Final Condenser.

5-181 Previous discussions of this condenser have indicated that its process performance leaves much to be desired. Its physical condition, however, is good. The cupro-nickel cladding and aluminum-brass tubes are in excellent condition. One problem, which was thought to be fouling, turned out to be anode covers which had come loose, were laying on the tube sheet and obstructing flow.

5-182 In an effort to increase heat transfer through this unit, we recommend and are currently using 2-pass flow rather than 4-pass, as shown in Figure 5-10, 312 Condenser Flow Patterns. This obviously increases the pump work, but when condensing temperatures are noticeably lowered, the over-all economy of production is improved.

5-183 For future design of plant final condensers, we recommend using the wealth of design information available from steam turbine condenser design. The function and effect on economy is very similar for both units. After considering this problem, we feel that the lack of performance is primarily due to lack of proper steam distribution to the tubes. The flow of steam through the tube banks and the accumulation of condensate is of prime importance to the functioning of a surface condenser. The most beneficial improvement to the Freeport condenser is improved baffling arrangement.

5-184 Vacuum Equipment.

5-185 This Plant is equipped with both steam jet evactors and a mechanical vacuum pump. It also has a barometric condenser for handling the steam and a limited quantity of noncondensibles from the 312 condenser and the deaerator. Attempts at operation, with steam jet evactors only, met with failure due to a lack of capacity. The operation of all this equipment is heavily dependent on the function being performed by the final condenser. We are not yet ready to declare this equipment marginal, since improvement in final condensation may solve capacity problems for this system.

5-186 The vacuum pump and barometric condenser have been quite reliable, even though often overloaded. The steam jet evactors probably have an internal defect, rather than a low of design capacity. Complete inspection of this equipment is scheduled.

5-187 MATERIALS AND CORROSION.

5-188 Metals and Alloys.

5-189 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 evolution 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.

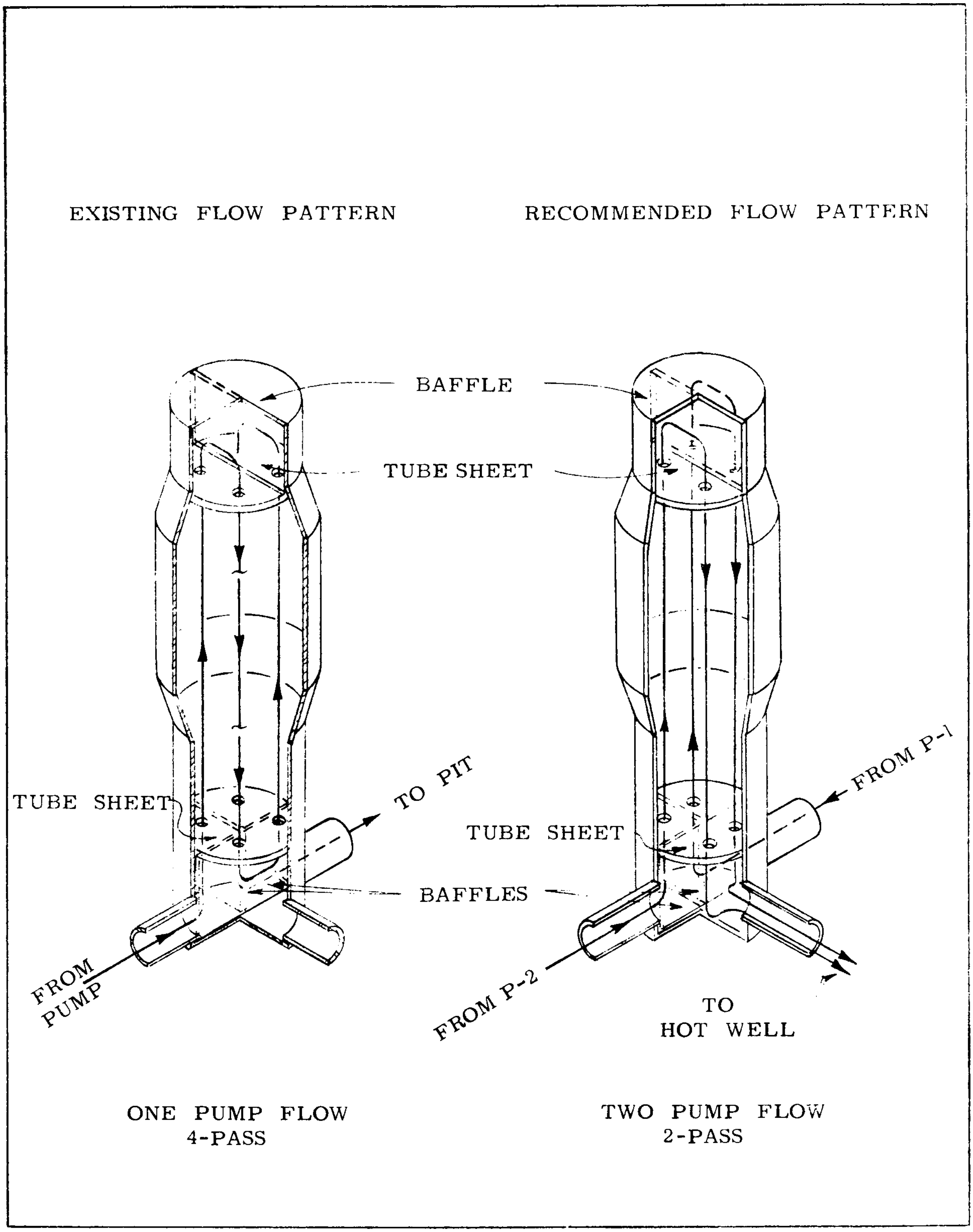


FIGURE 5-10. 312 CONDENSER FLOW PATTERNS

5-190 Carbon Steel. This material will serve satisfactorily in the non-deaerated sea water 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 sea water concentrate service up to 250°F. For satisfactory service in sea water 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 tube sheet material, the performance of carbon steel can only be termed as minimal. Cupro-nickel (70-30) cladding of tube sheets is recommended to lessen possible galvanic action and erosion, and minimize corrosion. Carbon steel is not satisfactory as a heat exchanger tube material for sea water service.

5-191 Cupro-Nickel. This material has performed excellently when provided 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 water boxes and tube sheets and also as cladding for the evaporator tube sheets and water boxes 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 occurs at the upper tube sheets) 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.

5-192 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 limits 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.

5-193 300-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 very corrosion resistant while in this form. Since the structure is a supercooled 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[7]. Work-hardening, however, provides the excellent resistance to cavitation damage exhibited by these alloys.

5-194 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[7]. This action, in conjunction with work-hardening, is the cause of rapid deterioration of stainless steel pump, impellers and shafts, in stagnant sea water. Also contributing to failure of these materials is the susceptibility to chloride stress corrosion cracking, primarily from the presence of calcium chloride, magnesium chloride, and sodium chloride[8].

5-195 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 sea water is continuous. If continuous sea water flow cannot be obtained, provisions should be made, during pump outages, for removing the pump from contact with sea water or flushing the pump with fresh water. This would prevent impeller or shaft contact with sea water 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.

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[7] Lewis, Charles F., Chemical Engineering, "Why Metals Fail-1", April 13, 1964.

[8] The International Nickel Co., Corrosion Resisting Properties of Austenitic Chromium-Nickel Stainless Steels, Copyright 1949.

5-196 Ni-Resist Irons. This material has exhibited good corrosion resistance when used as non-rotating castings in vertical turbine pumps. However, it has some of the same tendency as stainless steel to corrode in stagnant sea water. 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 indicates 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.

5-197 Non-Metals.

5-198 In general, use of non-metals has been very limited in this Plant and it is recommended that these materials be employed more extensively. Although special piping and equipment arrangements may be required to use these materials, the following service applications of non-metals are presented for consideration:

a. Glass-reinforced epoxy and/or polyester pipe should be considered for the non-deaerated, low pressure, feed water circuit (including the 312 condenser supply and return).

b. Extreme service grades of glass-reinforced centrifugally-cost epoxy pipe presumably can be used for service in the feed water circuit at temperatures as great as $250^{\circ}F$.

c. Epoxy-lined transite pipe could be used in the condensate system with a beneficial effect as to iron pickup.

d. Reinforced concrete pressure pipe could provide adequate service as gravity intake piping.

e. The product (split) tank could be a filament-wound epoxy or polyester vessel.

5-199 The existing non-metal corrosion spools installed between the 215 and 214 heat exchangers and experience of others using non-metals in these types of service, indicate these materials will perform satisfactorily. Costs of non-metals are, in general, in the range of carbon steel and plastic or metallic-coated (clad) carbon steel and are less costly than the solid alloys (such as monel and cupro-nickel).

5-200 Coatings.

5-201 The only successful coating in high temperature service has been a baked-on phenolic, when shop-applied to pipe. Baked-on epoxy (field-applied) to effect I has been unsatisfactory, as previously noted. External coatings (paint) are presently being applied over carefully prepared surfaces. A chlorinated rubber-based paint is giving satisfactory service. Other recommended systems are vinyl or epoxy applied over an inorganic based paint. Use of aluminum base paints is not recommended.

VI. PROPOSALS

6-1 DEVELOPMENT WORK.

6-2 The development runs described in the following paragraphs are part of the program given over-all approval in February, 1964. Some modifications have been made in the light of information developed from the work of fiscal year 1964. The description of each run is followed by modifications and additions required to achieve the objectives of the run. Numbering begins with Development Run No. 5 because the first four development runs were performed in fiscal year 1964. Development Run No. 5 is in progress during the writing of this Third Annual Report. All development runs are expected to requite approximately 30 days to complete unless otherwise noted.

6-3 Each development run will be studied for scale formation, changes in ΔT, heat exchange coefficient, distribution of evaporator load, energy utilization, and maintenance factors. Special emphasis on other systems will be noted individually with the applicable development run.

6-4 DEVELOPMENT RUN NO. 5.

6-5 This Run will evaluate the effect of polyphosphate addition at higher final concentration factor. It will also establish thermal and economic factors at the higher final concentration factor. The conditions will be:

Final Concentration Factor	3. 5
Production Rate	maximum possible
Effect I Steam Temperature	255^{o} F Maximum

6-6 If scaling occurs in effect XII, there may be a tendency to carryover brine into the product stream. If this occurs, the upper limit to product water conductivity for delivery to Freeport will be 250 microhms. The upper limit to ΔT from vapor-to-vapor across effect XII will be 28^{o} F; this is to avoid plugging tubes in effect XII.

6-7 If scale forms, polyphosphate will be added to the circulating cold sea water to study its effect on scale removal after the run is completed.

6-8 DEVELOPMENT RUN NO. 6.

6-9 This Run will study the effect of elevating temperatures in effect I. All efforts to provide maximum flow of sea water feed shall be made. The deaerator process will also be studied, using a new type packing.

A study will be conducted on the composition of vent steam and product condensate relative to carbon dioxide from effects II and III. The conditions are to be:

Final Concentration Factor	3. 3 maximum
Production Rate	maximum possible at final concentration factor
Effect I Steam Temperature	260° F $\pm 1^{\circ}$ F

6-10 It will be necessary to have pumping capability increased to an estimated 530, 000 LBS/HR sea water feed if an over-all benefit is to be realized from the increased temperature. Otherwise, it will be necessary to arbitrarily elevate effect XII temperature until extraction is reduced sufficiently to avoid scaling in the latter effects. Effect I ΔT will be closely monitored to ensure that the maximum upper ΔT limit of 15° F is not exceeded to prevent tube plugging.

6-11 DEVELOPMENT RUN NO. 7.

6-12 This will be an extended run utilizing all the information achieved to date. It will emphasize maximum scale-free production at a maximum effect I temperature of 260° F. Conditions will be establishedby considering all previous development work. During this period, intake, product water treating, silt settling, and deaeration subsystems will be studied in detail.

6-13 For this Run, it will be essential to have adequate pumping capability throughout the Plant. It will also be desirable to have additional heat transfer surface in effects IV and X. The recycle systems for effects XI and XII should be modified to provide for measured variable recycle flows to study this effect. Adequate mixing and distribution should also be provided. Effect XII should also be provided with a vapor bypass line equipped with a differential pressure control valve to study the result of vapor bypassing from effect XII water box to the 312 condenser. A mesh-type mist eliminator should also be provided in effect XII. The 312 condenser baffling modifications should be complete and maximum possible vacuum obtained in this unit. Use of the latter effect flash tanks may also be attempted during this Run.

6-14 DEVELOPMENT RUN NO. 8.

6-15 This Run will evaluate the operation at effect I temperatures of 265°F on the steam side and 255°F on the brine side. At these temperatures it becomes desirable to consider the use of recycle in effect I. It would therefore be advantageous to have the capability of flowing measured quantities of recycle to the water box of effect I. There is some indication that liquid-filled tubes, with flashing in the vapor body would be desirable. This Run should be repeated in five-degree-rising-temperature increments until anhydrite scaling is encountered. The maximum service temperature for effect I steam chest (steam side) is conservatively set at 280° F.

6-16 DEVELOPMENT RUN NO. 9.

6-17 This will be used as the trial run for either ion exchange or anhydrite scale seeding as a sea water softening and scale prevention technique. It will require significant equipment additions and modifications. The actual conditions will be established in close cooperation with the people who are proposing them.

6-18 During all these Runs, efforts will be made to design and evaluate better distributing devices, instrumentation, and control methods for process evaluation. The materials and corrosion problems will be continually scrutinized, with the ultimate objective of correcting them. Solution of the product water stabilization and filtrations problems will also be prime objectives in this program.

6-19 If the general outline presented here does not seem to be comprehensive or sufficiently detailed, the reader should consider the extensive data gathering, evaluation, and analyses required to have a knowledge of the operation procedures in these plants. It is obvious that sea water can be desalted in large volumes by evaporation, but the economics will not be improved except by more efficient use of energy sources and equipment. The improved design of future plantsis all-important to economic improvement. The present state-of-the-art is such that detailed knowledge of reaction mechanisms, mass and heat transfer, and flow of fluids are required to improve processes and equipment. This program, if conducted at this Plant, will yield much valid information on a reliable scale (as opposed to laboratory or pilot scale) for the obvious needs of future desalting plants.

6-20 MODIFICATIONS AND MAJOR MAINTENANCE.

6-21 Modifications in support of the development program are not necessarily aimed at the process only. They are intimately associated with maintenance as well. To provide increased over-all Plant efficiency, the following recommendations are made.

a. Study the use of anodes throughout the system and take appropriate action. This study must include many aspects of sacrificial anode use, galvanic corrosion and cathodic protection as well. The original anode calculations are not known and may have been arbitrary. Continued use of anodes, insulating gaskets, clad materials, etc., should have a basis in theory from which to evaluate their performance.

b. Provide equipment modifications and piping connections for on-stream flushing of the deaerator bottoms and exchangers 214 and 215. The former will improve silt removal prior to evaporation, and the latter will improve product filter life.

c. Provide an additional suction connection for pump P-4. This will improve net positive suction head and allow parallel operation of pumps P-4 and P-5a.

d. Test vinyl and epoxy system exterior coatings on new equipment or complete replacements. This will provide a comparison to the chlorinated rubber system now being used.

e. Eliminate heat exchanger 201 when the maintenance is next required on this unit.

f. Provide the following instrumentation on or around evaporators of effects VI and XI:

(1) Temperature measurement of the vapor above the distributor plate, the liquid on the tube sheet, and immediately inside the tube exit.

(2) Pressure measurement of the vapor above the distributor plate, between the distributor plate and the liquid on the tube sheet, and in vapor body, accurate to 0.5 inch of water.

(3) Temperature measurement of the cone brine in evaporators of effects V, VI, X and XI accurate to 0.5 degree F. This instrumentation, along with flow measurement capabilities already installed will greatly assist in detailing the actual conditions, for, and locations of, heat and mass transfer in the evaporators. It will allow accurate correlation of; and assist in, defining the parameters of design as well as performance prediction.

g. Replace effect I water box with a 70-30 cupro-nickel clad unit. This water box would serve for a period of not less than 18 months with no more than continued replacement of a protective coating (at \$2.00 per square foot). However, complete unit replacement of the water box will be required well in advance of the other Plant equipment. It is urgently recommended that this replacement be made immediately, with the clad material to confirm the selection of this material, and its serviceability under the existing and planned high-temperature high-velocity conditions.

h. Provide a suction drum for pump P-3. This addition will permit lowering the operating level of the clarifier-thickener to the point where the laundering weir can function. It will retain the obvious benefits of a constant suction head on pump P-3.

i. Replace sections of the sea water feed piping with a suitable non-metallic pipe in the high-temperature area of the Plant. These materials should be evaluated because of their significantly lower cost when compared to corrosion-resistant metallics.

j. As pump replacements are made in the low-pressure end of the Plant, equip some of the replacements with mechanical seals, for evaluation of both serviceability and maintenance requirements.

k. Provide a means of on-stream silt removal from the intake pump bay.

l. Design and install permanent stabilization chemical injection equipment. This must be necessarily delayed until results are obtained from the stabilization evaluation program.

m. Replace the present cartridge-type production water filter with a backflush, self-cleaning type. Present filter life is of such short duration that significant maintenance time is consumed in changing and cleaning cartridges.

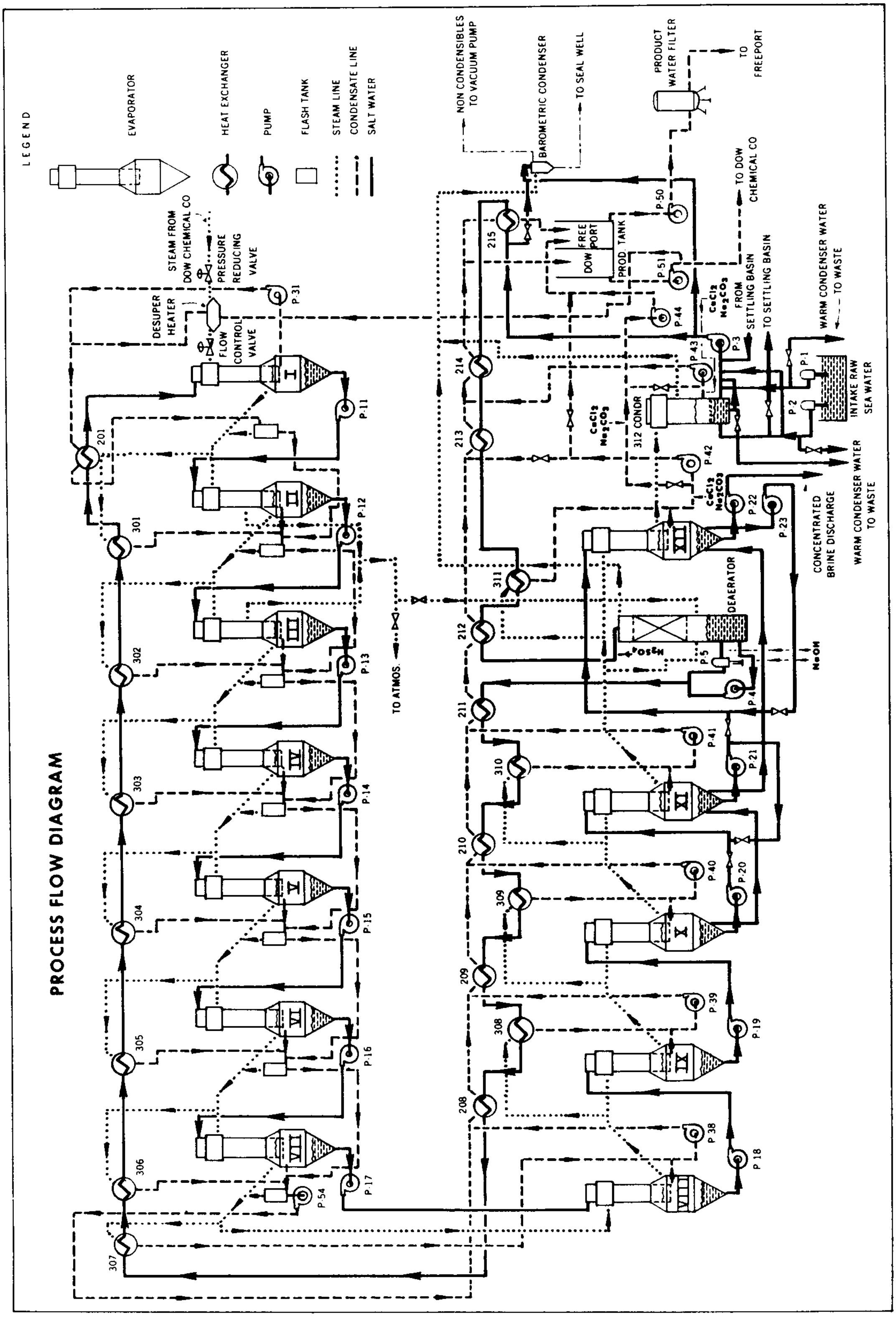
n. Provide a mist element with appropriate instrumentation for the effect XII evaporator. This device must be studied for future designs. This has a carryover problem at this time which is aggrevated by the flashing that now occurs in the cone.

VII. APPENDIX

7-1 INTRODUCTION.

7-2 This Section contains an assortment of data that support the Third Annual Report for the Sea Water Desalting Demonstration Plant No. 1. The order of data presentation is not necessarily compatible with the order that they have been referenced in the preceding Sections. However, an Index to the Appendix is provided below which will facilitate the location of any referenced data.

7-3 INDEX TO THE APPENDIX.



PROCESS FLOW DIAGRAM
LEGEND
EVAPORATOR
HEAT EXCHANGER
PUMP
FLASH TANK
STEAM LINE
CONDENSATE LINE
SALT WATER
STEAM FROM DOW CHEMICAL CO
PRESSURE REDUCING VALVE
DESUPER HEATER
FLOW CONTROL VALVE
NON CONDENSIBLES TO VACUUM PUMP
BAROMETRIC CONDENSER
TO SEAL WELL
PRODUCT WATER FILTER
TO FREEPORT
TO DOW CHEMICAL CO
DOW
FREE PORT
PROD. TANK
FROM SETTLING BASIN
TO SETTLING BASIN
WARM CONDENSER WATER TO WASTE
INTAKE RAW SEA WATER
312 CONDR
CONCENTRATED BRINE DISCHARGE
DEAERATOR
NaOH
TO ATMOS.

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LEDGER SUMMARY FOR MONTH OF JUNE 1964
(100 AND 200-SERIES ACCOUNT NUMBERS)
OSW DEMONSTRATION PLANT NO. 1, FREEPORT, TEXAS

ACCOUNT NUMBER	DESCRIPTION	LABOR HOURS	LABOR MONEY	MATERIAL	OTHER EXPENSE	MONTHLY TOTALS		CUMULATIVE TOTALS START OF MONTH	CUMULATIVE TOTALS FY TO DATE
ASSETS									
100	Plant								
100.1	Lands							34143.50	34143.50
100.2	Water Rights & Interests in Lands (other than title fee)							1102.00	1102.00
100.3	Roads and Walks							7400.00	7400.00
100.4	Buildings							76351.00	76351.00
100.5	Process Plant							1191278.45	1191278.45
100.6	Other Structures							2500.00	2500.00
100.7	Grounds and Improvements							2100.00	2100.00
100.9	Plant Not in Operation							59130.00	59130.00
100.16	Moveable Equipment							26307.50	26307.50
100.21	Distribution System							47050.48	47050.48
	Total Plant						$	1447362.93	1447362.93
101	Bank Account					(7588.11)	$	20505.08	12916.97
105	Drum Deposits					54.00	$	262.00	316.00
110	Inventory								
110.1	Spare Parts							14138.44	14138.44
110.2	Process Chemicals					61.91		1015.83	1077.74
	Total Inventory						$	15154.27	15216.18
111	Undistributed Plant Acquisitions								
111.3	pH meter							330.00	330.00
111.4	Fence					1159.00			1159.00
	Total Undistributed Plant Acquisitions						$	330.00	1489.00
115	Work in Progress								
115.1	Structure Modifications	58	192.89	700.02		892.91		697.13	1590.04
115.2	Painting Program	1653	4941.88	1961.22	1222.71	8125.81		21203.59	29329.40
115.5	Product Water Filter System	30	103.20	24.56		127.76		2487.80	2615.56
115.6	Clarifier-Thickener Modification	59	207.20	620.20		827.40		1565.99	2393.39
115.7	Corrosion Test Facilities	16	50.00	2032.34		2082.34			2082.34
	Total Work in Progress					12056.22	$	25954.51	38010.73
116	Retirement OF Property								
116.1	Pumps P-2 and P-5							9350.00	9350.00
116.2	200 Series Heat Exchangers							1623.54	1623.54
	Total Retirement of property						$	10973.54	10973.54
TOTAL ASSETS							$	1520542.33	1526285.35
LIABILITIES AND OTHER CREDITS									
201	Reserves for Depreciation								
201.1	Unallocated					5862.16	$	219308.84	225171.00
202	Government Equity								
202.1	Plant							1217849.27	1217849.27
202.2	Other					7448.33		208523.97	215972.30
	Total Government Equity						$	1426373.24	1433821.57
203	Stearns-Roger Equity								
203.1	Funds					27607.48		235226.95	262834.43
203.2	Non-reimbursable Denver charges					35.50		842.76	878.26
	Total S-R Equity					27642.98	$	236069.71	263712.69
210	Accrued Payroll					721.78	$		721.78
TOTAL LIABILITIES & OTHER CREDITS							$	1881751.79	1923427.04

LEDGER SUMMARY FOR MONTH OF JUNE 1964
(500 and 900-SERIES ACCOUNT NUMBERS)
OSW DEMONSTRATION PLANT NO. 1, FREEPORT, TEXAS

ACCOUNT NUMBER	DESCRIPTION	LABOR HOURS	LABOR MONEY	MATERIAL	OTHER EXPENSE	MONTHLY TOTALS	CUMULATIVE TOTALS START OF MONTH	CUMULATIVE TOTALS FY TO DATE
DIRECT EXPENSE								
Operations								
500	Supervision & Engineering	136	609.09			609.09	6086.38	6695.47
501	Labor	535	2171.04			2171.04	36285.84	38456.88
502	Materials & Supplies							
502.1	Process Chemicals			518.40		518.40	8149.69	8668.09
502.2	Laboratory Supplies			496.49		496.49	1343.26	1839.75
502.3	Product Treating			97.98		97.98	414.50	512.48
503	Utilities							
503.1	Steam				5847.53	5847.53	87925.33	93772.86
503.2	Electricity				1600.80	1600.80	20929.60	22530.40
	Total Operations Cost					11341.33	$ 161134.60	172475.93
Maintenance								
510	Supervision & Engineering	50	269.90			269.90	4020.82	4290.72
511	Labor							
511.1	Process Plant	816	2771.51			2771.51	17297.39	20068.90
511.2	Other Plant	76	256.19			256.19	5031.66	5287.85
511.3	General	226	708.98			708.98	70.86	779.84
512	Materials & Supplies							
512.1	Direct - Process Plant			589.54		589.54	14735.92	15325.46
512.2	Direct - Other Plant			87.26		87.26	3149.22	3236.48
512.3	Indirect			887.25		887.25	2571.38	3458.63
513	Other Maintenance Expense				1013.52	1013.52	1853.55	2867.07
	Total Maintenance Cost					6584.15	$ 48730.80	55314.95
Extraordinary & Experimental & Research Expense								
531	Extraordinary: Labor						77.23	77.23
540	Experimental & Research: Supervision & Engineering	214	935.78			935.78	4637.84	5573.62
541	Experimental & Research: Labor						726.93	726.93
542	Experimental & Research: Material			230.97		230.97	1013.33	1244.30
	Total Extraordinary and Experimental & Research Cost					1166.75	$ 6455.33	7622.08
TOTAL DIRECT COST						19092.23	$ 216320.73	235412.96
INDIRECT COSTS								
General & Administrative Expense								
920	Administrative Salaries	136	386.36			386.36	3731.91	4118.27
921	Payroll Burden	160	624.52		1046.02	1670.54	13293.20	14963.74
923	Reports and Procedures	128	559.10		14.76	573.86	3660.45	4234.31
924	Travel						767.27	767.27
925	Safety - Labor						396.34	396.34
926	Public Relations & Union - Labor	56	225.35			225.35	1554.55	1779.90
930	Office Supplies			222.20		222.20	941.71	1163.91
931	Safety Supplies						257.69	257.69
932	Communications				366.38	366.38	2438.21	2804.59
933	Public Relations & Union - Expense				134.43	134.43	111.88	246.31
935	Plant Transportation				8.77	8.77	363.02	371.79
936	Freight						794.15	794.15
937	Other G&A Expense						190.74	190.74
940	S-R Denver Office Expense		709.23		485.74	1194.97	20510.33	21705.30
950	Contractor Fee				1489.00	1489.00	16379.00	17868.00
951	Contractor Overhead Percentage				2209.34	2209.34	13920.68	16130.02
	Total G&A					8481.20	$ 79311.13	87792.33
Depreciation								
980.1	Total During FY					5862.16	$ 63708.99	69571.15
TOTAL INDIRECT COSTS						14343.36	$ 143020.12	157363.48
						$ 33435.59	$ 359340.85	392776.44
MAJOR MAINTENANCE								
992.1	Building and grounds					$ 2262.47		2262.47
TOTALS						$ 35698.06	$ 359340.85	395038.91

STARTUP PROCEDURE GUIDE

1. Charge caustic tank and check acid tanks.
2. Put top vents at specified settings.
3. Open bottom vents 2 or 3 turns.
4. Vent vapors from II and III downstream.
5. Start pit pumps to circulate water through heat exchanger 312.
6. Start P-3 pump to establish level in deaerator.
7. Use seal water for pump glands from Freeport Municipal Water system through P-51.
8. When level is established in deaerator, start P-5a (or P-4).
9. Establish sea water feed rate at approximately 200,000 lbs/hour.
10. As levels appear in evaporators, start brine pumps in sequence. Level controls will maintain proper levels.
11. Start acid to deaerator and adjust effluent water to a PH of 6.5-7.5 without addition of caustic.
12. Start vacuum pump and establish vacuum throughout system.
13. Set steam temperature regulator at proper setting.
14. Cut in steam - 12,000 lbs/hr.
15. Start condensate pumps as required in sequence.
16. Start Dow and Freeport product water pumps.
17. Raise sea water feed rate to 275,000 lbs/hr.
18. Raise steam to 20,000 lbs/hr.
19. Raise sea water feed rate to 350,000 lbs/hr.
20. Raise steam rate to 25,000 lbs/hr.
21. Raise sea water feed rate to 450,000 lbs/hr.
22. Raise steam rate to 31,000 lbs/hr.

STARTUP PROCEDURE GUIDE (CONT'D)

23. Flush Freeport and Dow supply line with converted water. When water quality is satisfactory, take integrator readings.

24. Adjust sea water feed rate and steam rate to achieve specified production rate and concentration factor.

25. Begin chemical feed to Freeport product.

26. Adjust vents to proper settings.

FREEZE PROTECTION CHECKLIST

1. Drain condensate pumps.
2. Drain P-4.
3. Drain idle condensate header.
4. Drain discharge line and check valves on idle condensate pumps.
5. Break and bleed seal water lines on idle pumps.
6. Turn on faucet inside of building.
7. Increase level control purge water flow rate as much as possible.
8. Drain plant utility water lines.
9. Plug in all heating elements.
10. Turn on all heat tracing lines.
11. Run P-42, P-43, and P-44. Crack discharge valves on P-42 and P-43
12. Drain fire hose lines.
13. Feed desuperheater from Dow pump.
14. Open all vents on steam chests.
15. Check water in monometers.
16. Open valve on purge water line at northwest end of structure.
17. Run P-53 to barometric condenser and to screen. Leave P-3 on barometric condenser also.
18. Isolate and drain all gage glasses.
19. Cut in unused portions of flash system.

SAMPLE CALCULATIONS AND DATA ON SCALING VERSUS APPARENT "U" COEFFICIENT

Quantitative Prediction of Scale from Changes in Apparent Heat Transfer Coefficient.

In a period of sixty (60) days the heat transfer coefficient of the effect XII evaporator went from a clean value of 280 to a value of 235. This loss in heat transfer rate is due primarily to the build-up of scale on the heat transfer surface.

It would be desirable to be able to calculate from operating data the rate at which this scale is forming. If it is assumed that the rate of scale build-up is proportional to the amount of evaporation from the start of the run, then one possible means of measuring this rate would be to apply the following equation

$$\frac{1}{U^2} = \frac{1}{U_o^2} + C\,\theta$$

U = Heat transfer coefficient
U_o = Clean heat transfer coefficient
θ = Time, days
C = Constant

In order to establish the validity of this equation, two (2) graphs were made (Fig. 1). The first was a plot of U versus θ and the second a plot of $1/U^2$ versus θ. If this equation is valid for this application, the graph of $1/U^2$ versus θ will be a straight line.

A word of explanation of the graphs in Figure 1 is in order. The plant operated for a period of thirty-four (34) days then was shut down for twenty-four (24) hours in order to install the replacement pump for P-5. After starting back up, a rise in the heat transfer coefficient was observed. This would be expected if a portion of the scale, already deposited, re-dissolved. Experience has shown that this scale is soluble in seawater. During the period of shutdown and start-up, the XII effect would "see" normal seawater and perhaps some of the scale did dissolve.

Although there is considerable scatter in the measured points, nevertheless it appears there is some merit in the application of this technique for measuring scaling rates. In order to verify this, additional data must be taken and evaluated.

Scaling in XI

In this evaporator the heat transfer coefficient went from a clean value of 297 to a value of 232 in a period of sixty (60) days. A similar plot to the one for XII is shown in Figure 2. Note again the rise in the coefficient corresponding to the short shutdown period.

SAMPLE CALCULATIONS AND DATA ON SCALING
VERSUS APPARENT "U" COEFFICIENT (CONT'D)

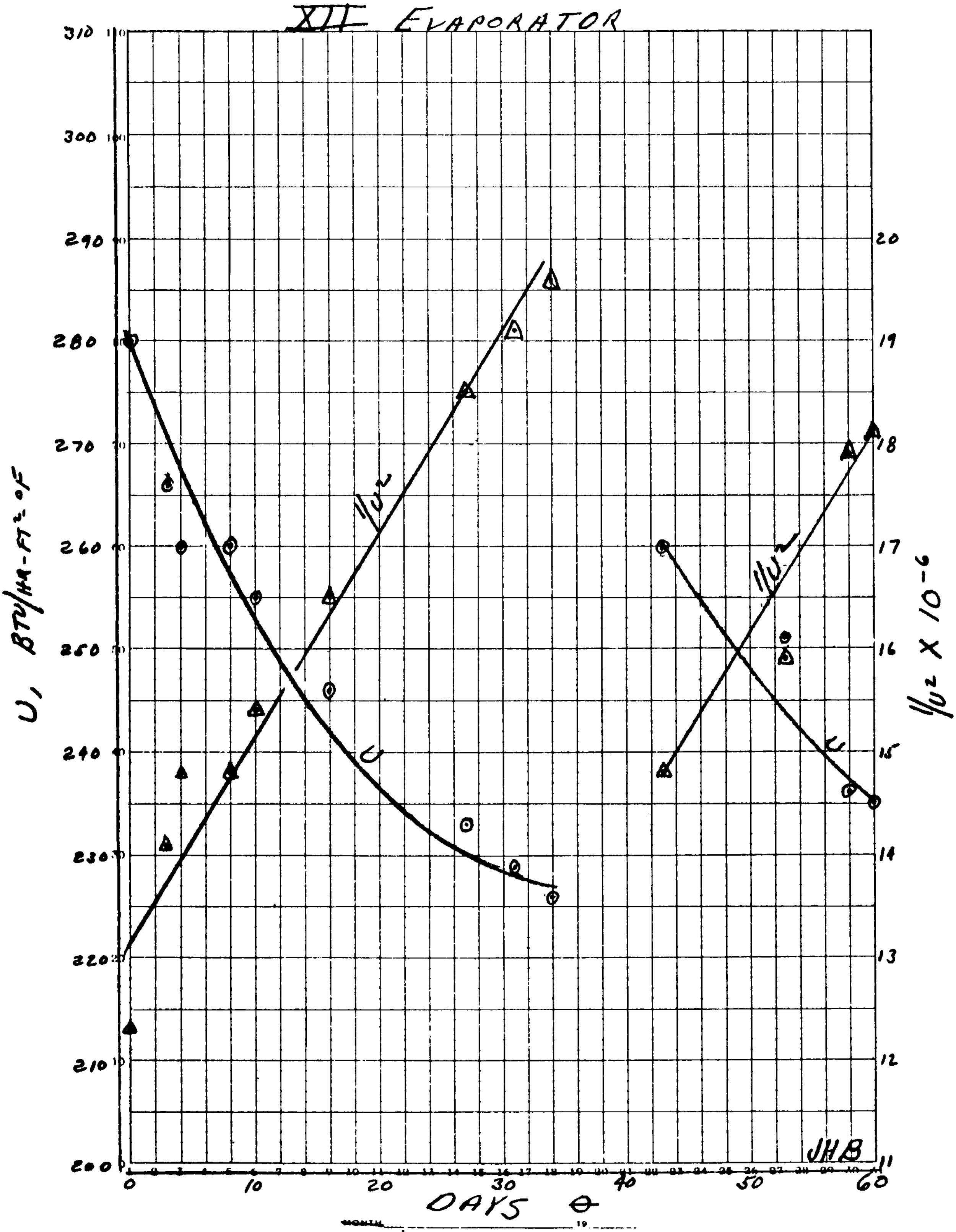

FIGURE 1

SAMPLE CALCULATIONS AND DATA ON SCALING
VERSUS APPARENT "U" COEFFICIENT (CONT'D)

If this method of predicting scaling rates is valid for this process, its value will be twofold. First of all, it can be used to calculate the optimum time of operation between periods of scale removal that will give maximum overall capacity. However, more important than this, it can be used as an experimental tool. A given set of operating conditions can be evaluated as to their scale forming potential merely by maintaining them long enough to measure four or five points and establish the slope of the $1/U^2$ versus θ line. This slope then is a measure of the scaling rate.

SAMPLE CALCULATIONS AND DATA ON SCALING
VERSUS APPARENT "U" COEFFICIENT (CONT'D)

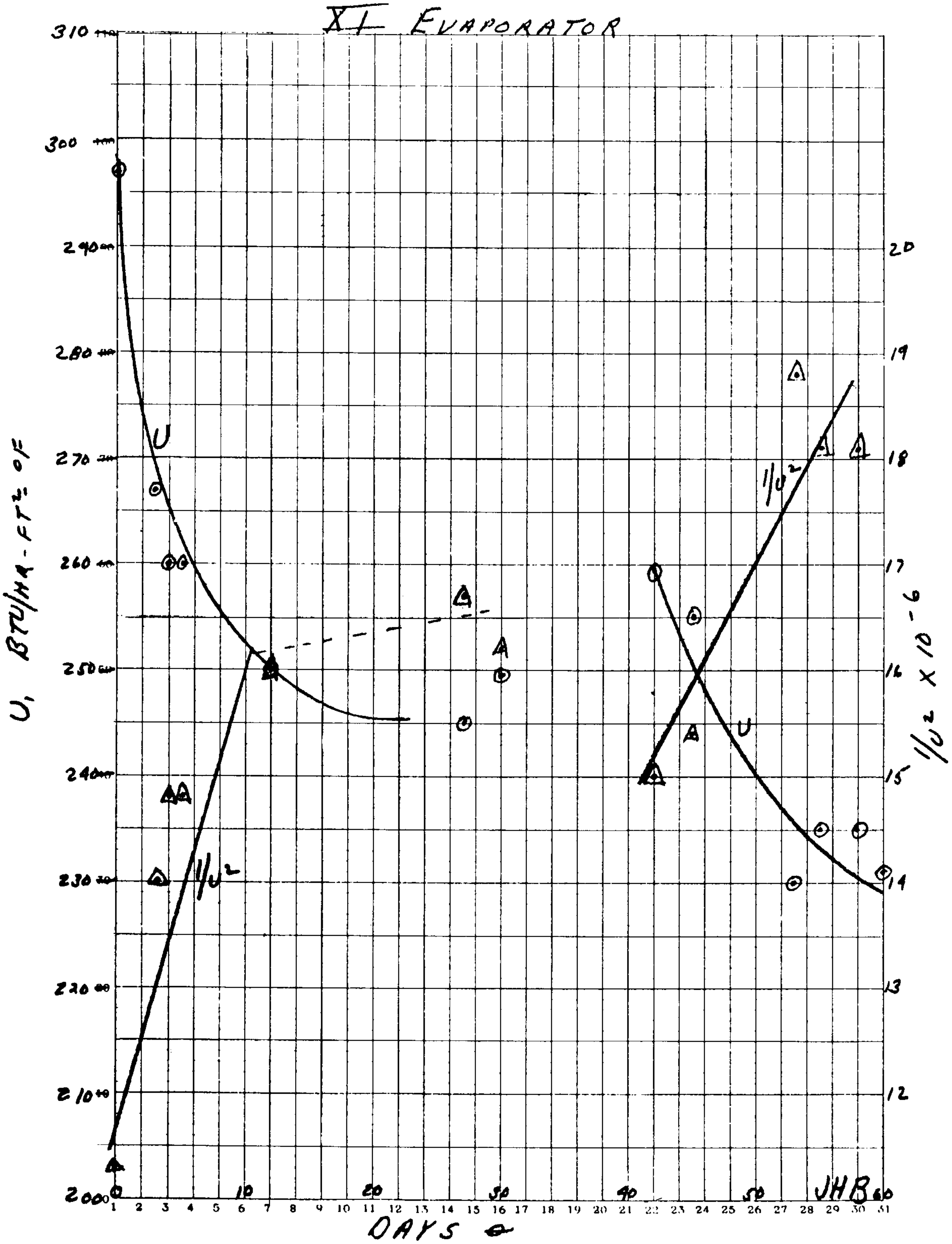

FIGURE 2

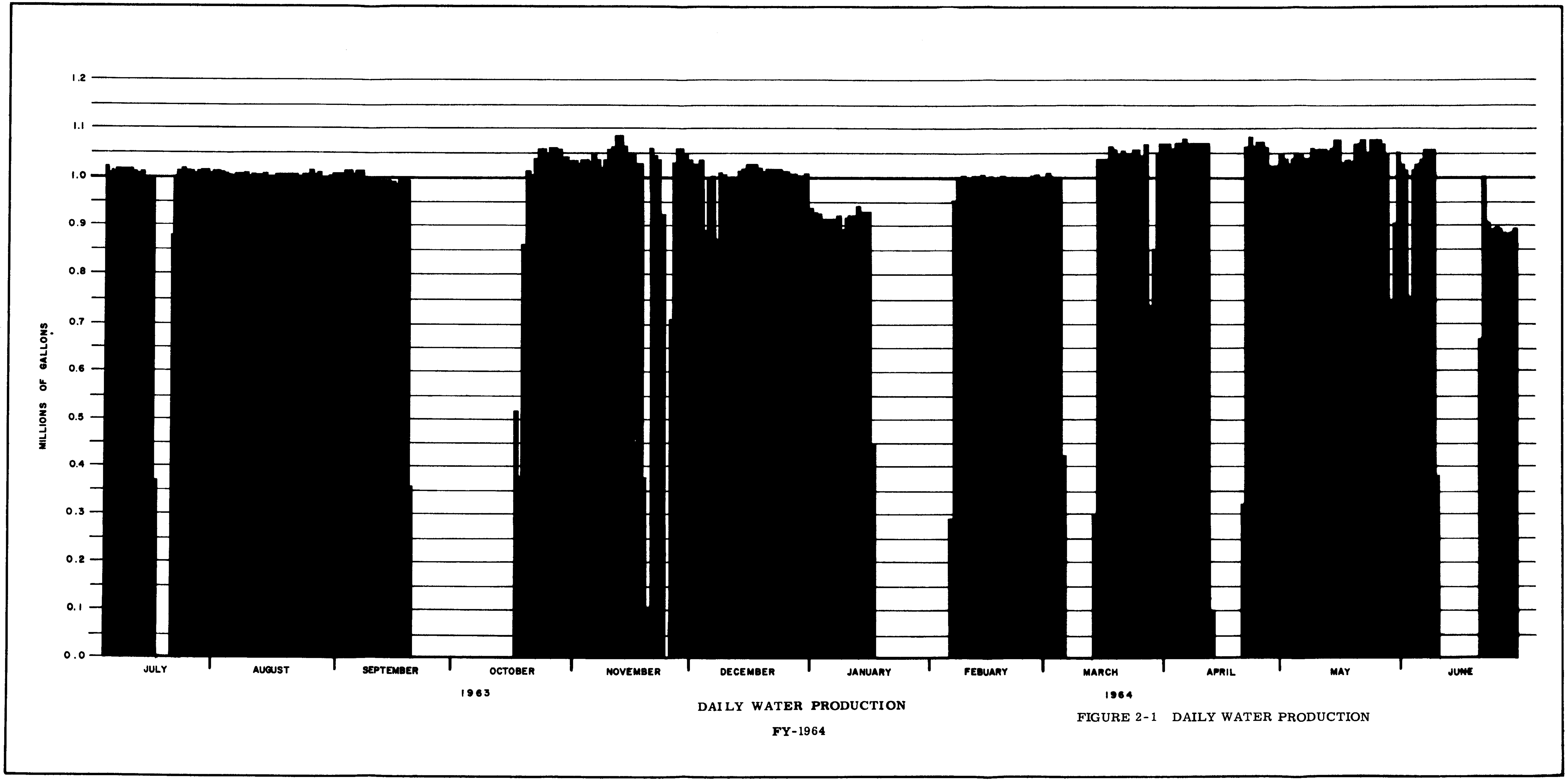

FIGURE 2-1 DAILY WATER PRODUCTION

SAMPLE CALCULATIONS AND DATA ON SCALING
VERSUS APPARENT "U" COEFFICIENT (CONT'D)

Sample Heat Transfer Calculations

5-19-64 DATA

SWF	14.03/1.017	=	13.80
IX	34.40/1.043	=	32.98
X	40.78/1.052	=	38.76
XI	48.20/1.061	=	45.43
XII	60.84/1.077	=	56.49

Brine Rate From IX = $\frac{470,000 \times 13.80}{32.98}$ = 196,665

Brine Rate From X = 167,337 #/hr
Brine Rate From XI = 142,769 #/hr
Brine Rate From XII = 114,817 #/hr

Brine To XII = 142,769 #/hr

Vapor To XII = $(167,337 - 142,769) - 1904 - \frac{(470,000)(7)(0.972)}{1019.4}$

= 24,568 - 1904 - 3137 = 19,527

Brine From XII = 114,817 #/hr
Vapor From XII = 27,952 #/hr
Cond. From XII = 19,527 #/hr

Heat In

(142,769)(101)(0.907)	=	13,079,000
(19,527)(1118.3)	=	21,837.000
		34,916,000

Heat Out

(19,527)(98.90)	=	1,931,000
(27,952)(1111.6)	=	31,071,000
(114,817)(85)(0.889)	=	8,676,000
		41,678,000

No check

Use Alternate Calculation Method

Sample Heat Transfer Calculations (Cont'd)

Vapor Rate From X = 196,665-167,337 = 29,328

Stm. to 310 = $\frac{(470,000)(7)(0.973)}{1011.7}$ = 3164 #/hr

Vaporization in Stm. Chest of XI = $\frac{(29,328-3164)(1011.7)}{1018.8}$

= 25,982 #/hr

Brine Flashed in XI

Let X = lbs Brine flashed at 133° F

(167337)(113) = 1118.3 X + (167,337-X)(101)

X = 1974 #/hr

Total Vapor From XI = 1974 + 25,982 = 27,956 #/hr
Vapor To XII = 27,956 - 1904 - 3137 = 22,915 #/hr
Brine To XII = 167,337 - 27,956 = 139,381 #/hr
Vap. In Stm. Chest of XII = $\frac{(22,915)(1,019.4)}{(1027.5)}$ = 22,734 #/hr

Brine Flashed In XII

Let X = # Brine Flashed at 117° F

(139,381)(101) = 1111.6 X + (139,381-X)(85)

1026.6 X = 2,230,096

X = 2172 #/hr

Vapor From XII = 2172 + 22734 = 24906 #/hr
Brine From XII = 139,381 - 24,906 = 114,475 #/hr

Heat In

(139,381)(101)(0.907)	=	14,032,000
(22,915)(1118.3)	=	25,626,000
		39,658,000

Heat Out

(114,475)(85)(0.889)	=	8,650,000	
(24,906)(1111.6)	=	27,686,000	
(22,915)(98.90)	=	2,266,000	
		38,602,000	Close Enough

$$U_{XII} = \frac{23,360,000}{(6574)(16.0)} = \underline{\underline{222}}$$

ECONOMY RATIO CALCULATIONS

Table in Sec. 5.	Table 5-1	5-2	5-3	5-4
Net Product	8,134,600 lbs	6,958,665	333,320	350,000
Act Boiler Stm to Plant	751,000 lbs	626,300	31,180	32.000
Correction factor	1.07	1.07	1.07	1.036

$$\text{Overall economy ratio} = \frac{\text{Net Product}}{\text{Boiler stm to Plant}}$$

$$\text{Corrected economy ratio} = \frac{\text{Overall economy ratio}}{\text{Correction factor}}$$

for 5-1 $\text{OER} = \frac{8,134,600}{751,000} = 10.83$ $\quad \text{CER} = \frac{10.83}{1.07} = 10.12$

for 5-2 $\text{OER} = \frac{6,958,665}{626,300} = 11.11$ $\quad \text{CER} = \frac{11.11}{1.07} = 10.38$

for 5-3 $\text{OER} = \frac{333,320}{31,180} = 10.69$ $\quad \text{CER} = \frac{10.69}{1.07} = 10.01$

for 5-4 $\text{OER} = \frac{350,000}{32,000} = 10.94$ $\quad \text{CER} = \frac{10.94}{1.036} = 10.56$

NORMALIZED COST CALCULATIONS

Reference OSW R&D Report No. 72, Bechtel Corporation Pages 29, 30, 31 and 31a. Since all the applications of escalation and the additions for Site and Design Criteria are applicable, the values of page 31a are used for a starting point.

I PIE

a.	Evaporator effects	no change	$ 609,740
b.	Deaerator Equipment	" "	16,600
c.	Vacuum Equipment	" "	14,700
d.	Heat Exchangers - reduced by $20,455 for removal of HK's 202, 203, 204, 205, 206 and 207.		243,645
e.	Pumps and Drivers		31,400
	reduced by $6,810 for removal of Pumps P-31, P-32, P-33, P-34, P-35 and P-36 no longer in service.		65,545
f.	Desuperheater and Seal Water Pump	no change	31,400
g.	Water Cooling Equipment	" "	11,500
h.	Air Compressor Equipment	" "	7,000
i.	Acid and Caustic System	" "	5,630
j.	Concrete Work	" "	21,800
k.	Miscellaneous	" "	66,340
	Sub Total		$ 1,093,900

Note:

from Page 31a for PIE 1. of Table 3-2

Item 13 Electrical is subtracted	-70,275
Item 14, Boiler is subtracted	-138,200
Item 11, Ranney Water System is subtracted	-120,000
Item 16, Concrete Work is added	+ 21,800
Item 21, Miscellaneous is added	+ 66,340
Adjusting Item 4 subtract	- 20,455
Adjusting Item 5 subtract	- 6,810

then total is
1,362,000-70,275-138,700-120,000+21,800+66,340-20,455-6810 = $ 1,093,900

2. Standard Engineering Equipment

a.	Spare parts	no change	21,200
b.	Overhead crane	" "	9,360
	Sub Total		30,560
	Total PIE		$ 1,124,460

NORMALIZED COST CALCULATIONS - OPERATING COST (Cont.)

Energy	first 80,000 KWhr @ $.0070	560.00
	next 120,000 KWhr @ .0055	660.00
	next 32,500 KWhr @ .0045	146.25
	fuel adjustment 232,500 KWhr @ .00145	337.12
	Sub Total	1,703.37
	Gross Bill	2,173.37

Credit 370x360+36,000 = 169,200

232,500-169200 = 63,300 at 1 mil (63.30)

$ 2,110.07

NORMALIZED COST CALCULATIONS

¢ per 1000 gallons.

$$\frac{12x2110.07}{330x1,010} = \frac{7673}{1,010} = 7.60 \text{ ¢/1000 gallons}$$

$ per Stream day = $76.73

b. Fuel (oil, gas)

Assume a net overall economy ratio of 10.57 per Dev. Rept. No. 2

This means that .0946 lbs steam at 169 psia and 500°F must be fed to plant for each lb of product produced.

Converting this to lbs per gallon (8.31 lbs/gal) yields (8.31) (.0946) = .786 lbs of stm/Gal H_2O net

Converting from 169 psia and 500°F to 36 psia saturated

$$(.786)\,\frac{(1272.0-229.60)}{(1167.6-229.60)} = .872 \text{ lbs Sat Stm/gal } H_2O$$

If condensate is returned to boiler at 88°F h_f = 66.0 than Δh = 1167.6 - 66.0 = 1101.6 Btu/lb.

Per Stream Day, then

$$\text{Boiler heat Output} = \frac{(1101.6)(\text{-}872)(1,010,000)}{10^6} = 970.201 \text{ Million Btu}$$

$$\text{Boiler heat Input} = \frac{970.201}{.81} = 1,197.78 \text{ Million Btu}$$

$$\text{Fuel required} = \frac{1,197.78}{6.3} = 190.12 \text{ bbl per stream day}$$

NORMALIZED COST CALCULATIONS (Cont.)

II Process Facilities

1.	Raw Feed & Cooling Water Intake Ranney Water System (Item 11 of 31a)		\$ 120,000
2.	Site Development	no change	36,800
3.	Insulation	" "	40,200
4.	Painting	" "	18,700
5.	Electrical (Item 13 of 31a)		70,275
6.	Piping is included in PIE		----
7.	Instruments	no change	50,300
8.	Buildings	" "	42,800
9.	Boiler Plant (Item 14 of 31a)		138,700
		Total Process Facilities	517,775

III Other Plant Costs

1.	Engineering	no change	120,000
2.	Interest on Investment 2% x 1,642,235		32,845
3.	Startup Expense	no change	30,000
4.	Cost of Site (Item 24 of 31a)		20,000
		Total Other Plant Costs	202,845
		Total Plant Costs	\$ 1,845,080

Capital Cost per Gallon of Daily Capacity (net) \$ 1.827

NORMALIZED COST CALCULATIONS - OPERATING COST

Reference OSW R&D Report No. 72 pages 42 thru 50.

I. Direct Operating Costs

1. Energy
   Electric power
   From Dev. Run No. 2 use

$$257{,}000 \text{ KwHr/mo } \frac{330}{365} = 232{,}350 \text{ KwHr per month}$$
(for 330 Stream Days/Year.)

$$\frac{257{,}000 \text{ KvHrs}}{720 \text{ Hr}} = 357 \text{ KVA ave demand.}$$
use 370 KVA max demand.

then monthly electrical charges are

Demand	first 100 KVA @ \$2.00	200.00
	next 270 KVA @ \$1.00	270.00
	Sub Total	470.00

\$ per Stream day = 190.12x\$2.35 = \$446,782

¢/1000 gal - 44.24

The balance of the costs shown in table 4-3 are adequately explained in the text.

QUANTITIES APPEARING ON MAIN DATA SHEET

Production

Gallons Total Production = Dow Gallons + Freeport Gallons + Plant Use & Excess Gallons

$$10^6 \text{ Pounds Total Production} = \frac{8.31 \times \text{Gallons total production}}{10^6}$$

Sea Water Feed

M LBS Total = M LBS Rate x Operating hours (M=1000)

Concentration Factor In & Out — Arithmetic mean of readings taken

Temperatures, Pressures & Others — Arithmetic mean of readings taken

Ambient Conditions - Taken from operators written Log.

Extraction Ratio - The pounds of H_2O extracted from each pound of sea water feed.

$$E.R. = 1 - \frac{\text{Concentration Factor in}}{\text{Concentration Factor out}}$$

(see derivation P 7-23)

NOTE: This does not include any condensate from heating steam

M LBS H_2O extracted from Sea Water

$$(E.R.)(\text{Sea Water Feed}) = \text{M LBS } H_2O \text{ extracted} = P_E$$

M LBS Steam vented or used in barometric condenser & vacuum jets

$$S_V = \text{M LBS } H_2O \text{ extracted} + \text{M LBS Steam Fed to plant} - \text{M LBS Total Product}$$

$$S_V = P_e + S_F - P_T$$

LB H_2O per KW HR pump work

$$W_p = \frac{\text{total LBS } H_2O \text{ produced}}{\text{total KW HR Used}}$$ (includes heating steam condensate)

$$W_p' = \frac{\text{total LBS } H_2O \text{ produced} - \text{LBS steam feed}}{\text{total KW HR used}}$$ (does not include heating steam condensate)

Design Work of Separation

h_g = 1287.5 — enthalpy of inlet steam condensate
h_f = 56.0 — leaving as product @88°F SAT LIQ
h_{fg} = 1231.5

Work available from this heat input is given by

$$\left(\frac{T_i - T_o}{T_i}\right) \frac{Q\ in}{3413} = \text{KW HR/LB Steam}$$

$$\frac{(990 - 548)}{(990)} \quad \frac{1231.5}{34.3} = \underline{.161}$$ KW HR/LB Steam or 161 Watt HR/LB Steam

Separation work Not counting heating Steam Condensate as product

LB Steam/ LB H_2O = $\frac{1}{11.65}$ = .0858

then (.0858)(161) = 13.814 watt hours/LB H_2O produced

Design Work of Pumping not counting heating steam condensate as product

$$\frac{8.107 \times 10^6 \text{ watt hours}}{(8.31 - .657)\ 10^6 \text{ LBS}} = \frac{8.107}{7.653} = 1.06 \text{ watt hours/LB}$$

Design Over-all Temperature Difference

(vapor temperature to No. 1 - Vapor temperature in No. 12) = ΔT
261 - 115 = 146

Question: To determine the relationship of the ratio of water produced per pound of steam fed where this heating steam is considered as product and where it is not.

with condensate in product
LBS H_2O/LB Steam = $\frac{P}{S}$

where P = Total Product
S = Heating Steam

without condensate as product
LBS H_2O/LB Steam = $\frac{P-S}{S}$

let X = the difference $\frac{P}{S} = \frac{P-S}{S} + X$ or $X = \frac{P}{S} - \frac{P-S}{S} = 1$

so LBS H2O/LB Steam (without condensate) = LBS H_2O/LB Steam with- 1

Let N_s = LBS of salt per pound of sea water feed
N_{wi} = LBS of H_2O per pound of sea water feed
N_{wo} = LBS of H_2O per pound of concentrated brine discharge
Define Concentration Factor

$$CF = \frac{\dfrac{N_s}{N_s + N_{wi}}}{.0352} \quad \leftarrow \text{Eq. I}$$

Let CF_i = Concentration factor in
CF_o = Concentration factor out

by definition $N_s + N_{wi} \equiv 1$ LB for each pound of feed $\leftarrow$ Eq. II

also by definition $N_s \equiv .0352\ CF_i$ LBS

and $N_{si} = N_{so} = N_s$ because only the water is removed in evaporation.

Substituting in II $\quad .0352\ CF_i + N_{wi} = 1$

or $\quad N_{wi} = 1 - .0352\ CF_i \quad \leftarrow$ Eq. III

by substitution in I $\quad CF_o = \dfrac{N_s}{N_s + N_{wo}}$ or rearranging

$$N_{wo} = \frac{N_s}{.0352\ CF_o - N_s}$$

<u>Define extraction ratio</u> as E.R. $\equiv N_{wi} - N_{wo}$ = LBS H_2O Separated/LB SWF

Substituting for N_{wi} from III and for N_{wo} from IV

$$ER = 1 - \frac{CF_i}{CF_o}$$

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