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DEPARTMENT OF THE INTERIOR

DEVELOPMENT OF THE DIRECT FREEZE SEPARATION PROCESS

BY
CARRIER CORPORATION
UNDER CONTRACT 14-01-0001-286

OFFICE OF SALINE WATER

RESEARCH AND DEVELOPMENT PROGRESS REPORT NO. 113

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UNITED STATES

DEPARTMENT OF THE INTERIOR

STEWART L. UDALL, SECRETARY

KENNETH HOLUM, ASSISTANT SECRETARY
FOR WATER AND POWER DEVELOPMENT

SALINE WATER RESEARCH AND DEVELOPMENT PROGRESS REPORT NO. 113

DEVELOPMENT OF THE DIRECT FREEZE SEPARATION PROCESS

BY

WILFRED J. HAHN, RODNEY C. BURNS.
ROBERT S. FULLERTON AND DEWEY J. SANDELL
CARRIER CORPORATION

FOR

OFFICE OF SALINE WATER

CHARLES F. MACGOWAN, DIRECTOR

Joseph J. Strobel, Chief, Processes Development Division

Walter W. Rinne, Chief, Special Projects Branch

Stephen C. Verikios, Project Engineer

JUNE 1964

Created in 1849, the Department of the Interior--America s Department of Natural Resources--is concerned with the management, conservation, and development of the Nation's water, wildlife, mineral. forest, and park and recreational resources. It also has major responsibilities for Indian and Territorial affairs.

As the Nation's principal conservation agency, the Department of the Interior works to assure that nonrenewable resources are developed and used wisely, that park and recreational resources are conserved for the future, and that renewable resources make their full contribution to the progress, prosperity, and security of the United States--now and in the future.

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 of sea and other saline waters.

Except for minor editing, the data herein are as contained in reports submitted by Carrier Corporation under Contract No. 14-01-0001-286, covering research carried out through May 1963. The data and conclusions given in this report are essentially those of the Contractor and are not necessarily endorsed by the Department of the Interior.

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1.0 INTRODUCTION & SUMMARY

This report presents results of project activity for the period August 17, 1962 through May 17, 1963, under Contract #14-01-0001-286 entitled "Development of the Direct Freeze Separation Process".

The saline water conversion pilot plant was originally constructed and operated in Syracuse, New York under a previous contract. The plant was dismantled in June, 1960 and re-erected at the Wrightsville Beach site. Technical feasibility was established in Syracuse using artificial sea water and demonstrated again during initial operation at the sea coast site.

The program for this contract was essentially an extension of work under the previous contracts with emphasis on investigation of design concepts which would lead to economic improvements of capital and operating costs in a production size plant. Among the advances made along these lines at the pilot plant was a significant increase in the production capacity of the separation column while simplifying the column structural design and control scheme.

In parallel with the pilot plant test program, the practical range of plant capacities for this process was evaluated. A feasible modular plant capacity was determined to be approximately 165,000 gallons per day. Larger capacities are practical and are discussed in the text. Owning and operating costs have been estimated in detail for the above plant capacity and are presented in this report.

1.1 OBJECTIVES

The general objective of this contract was to further evaluate the 15,000 gallons per day pilot plant at Wrightsville Beach, North Carolina, for the purpose of determining the technical and economic feasibility of the direct freeze-separation process for the conversion of sea water to fresh water. The following procedures were specified by the contract.

(1) Continue operational evaluation of the water vapor absorption type freezing process components and the complete system to give reliable, continuous, trouble-free operation.

(2) Continue, and amplify, tests on materials, automatic controls, and associated equipment.

(3) Undertake studies that will lead to simplification of components, operation, maintenance and operating personnel requirements.

(4) Develop and test design concepts that will contribute to economic improvements of capital and operating costs.

(5) Evaluate the practical range of plant capacities suitable to this process.

1.2 BACKGROUND

During the last seven years since 1956, Carrier Research and Development Co. has developed the freezing process of saline water conversion from the stage of initial feasibility studies to successful operation of a 15,000 GPD pilot plant. This work has been done by joint participation under several contracts from the Office of Saline Water.

Following is a very brief summary of the contracts since June 1956.

Contract 14-01-001-86, June, 1956 to November, 1958

This contract covered the initial feasibility and laboratory studies, and also the construction and testing of a 300 gallon per day experimental model.

Contract 14-01-001-169, December, 1958 to July, 1960

This contract covered the design, construction and testing of a 15,000 GPD pilot plant at Syracuse, New York. It also included disassembly and shipment of this pilot plant to the sea coast at Wrightsville Beach, North Carolina.

Contract 14-01-001-213, August, 1960 to August, 1962

This contract covered erection and operational testing of the 15,000 GPD pilot plant at Wrightsville Beach.

Carrier activity has been concentrated on a direct freeze process of the primary refrigerant type. The indirect freezing process was considered but was discarded because of the higher cost for additional heat transfer surface and the higher energy requirements of this type process.

The direct freeze process using a secondary refrigerant had not been considered seriously until just lately. In Carrier's opinion such processes were not desirable because the secondary refrigerants heretofore available presented excessive hazards of toxicity and explosion. Recently a new refrigerant has become available which would eliminate these hazards. Carrier now is studying the feasibility of a secondary refrigerant process using this new refrigerant.

The primary refrigerant type of direct freezing process may be classified further into the vapor compression or the vapor absorption type. Both of these types were examined very early in the initial feasibility studies. It was found that the vapor compression type theoretically required substantially less energy and, therefore, could have a substantially lower operating cost potential than the vapor absorption method.

However, due to the high vacuum and high capacity required, no compressor suitable for the conditions was available on the market. On the other hand standard absorption type refrigerating machines were available which could be adapted to the required conditions for the vapor absorption type process. Also, at that time the basic functions of freezing and separating of the ice from the brine had not yet been demonstrated. Therefore, it was decided to use the vapor absorption type of process as the most economical way to prove out these latter fundamentals. The development of a special compressor was not considered worthwhile when the feasibility of the basic process had not been established.

Since the feasibility of the basic direct freezing process now has been proven, the vapor compression process also is being investigated further along with the secondary refrigerant process.

2.0 DESIGN OF THE PILOT PLANT

A preliminary design for the pilot plant was described in the final report under contract 14-01-001-86 (OSW Research & Development Report No. 32). Complete drawings covering the general design of the plant and the major items of special equipment were submitted in subsequent reports and have not been repeated in this report. The process piping diagram presented on RD1039-9193 shows the pilot plant as it was at contract termination.

In July, 1961, the capacity of the auxiliary refrigeration was increased to allow operation with 83°F and higher sea water temperatures. All other equipment proved satisfactory for production at the design capacity of 15,000 gallons per day at the Wrightsville Beach site. A detailed discussion of the evolution of the separation column, which ultimately led to reduction in plant capacity is covered in this report.

Figure 1 is a photograph of the pilot plant at Wrightsville Beach taken in March, 1963 near the end of the present contract period. Figure 2 shows the top of the separation column with the conveyor type scraper developed during this period.

2.1 THE PROCESS FLOW

The process flow may be followed on RD1039-9193 on which the transition from sea water to potable product has been accented as a heavier line. It also is shown schematically on drawing RD1039-9153.

Sea water is pumped from Banks Channel, an arm of the ocean adjacent to the plant, through coarse (3/32") and fine (40-mesh) strainers into a cold deaerator under vacuum. Deaeration minimizes corrosion as well as promoting better refrigeration performance.

The sea water feed is metered and split into parallel streams through two exchangers where it is precooled by streams of waste brine and fresh water product. Although other routings are possible, the feed is currently pumped into the freezer standpipes with the slurry recycle stream.

The recycle slurry consists of a mixture of 9 or 10 per cent flocculent ice crystals in 6 to 7 per cent salt recycle brine. The slurry overflows standpipes in a freezer vessel which is maintained at about 3 mm Hg absolute.

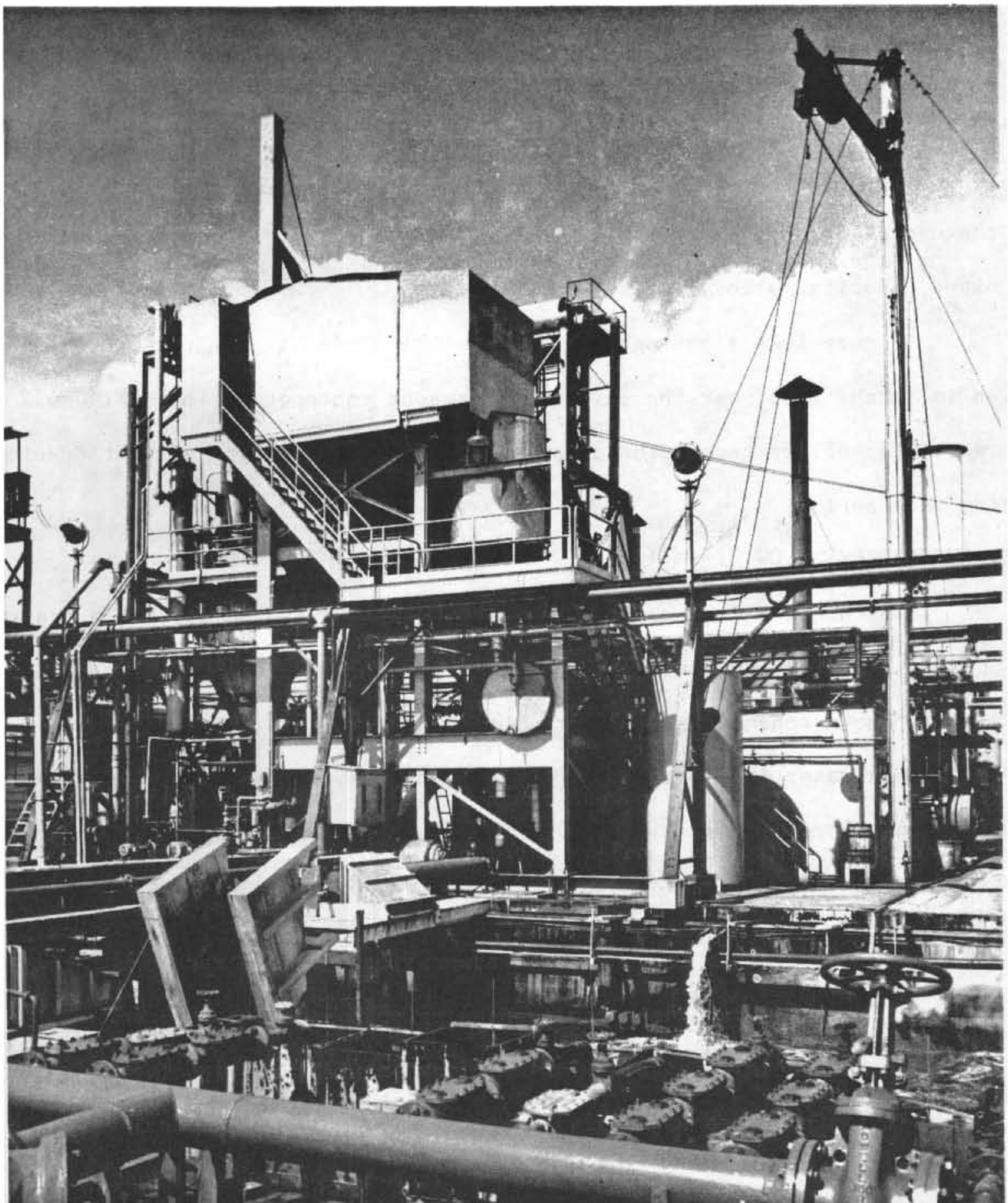


FIGURE 1: PILOT PLANT AT WRIGHTSVILLE BEACH, NORTH CAROLINA - MARCH 1963

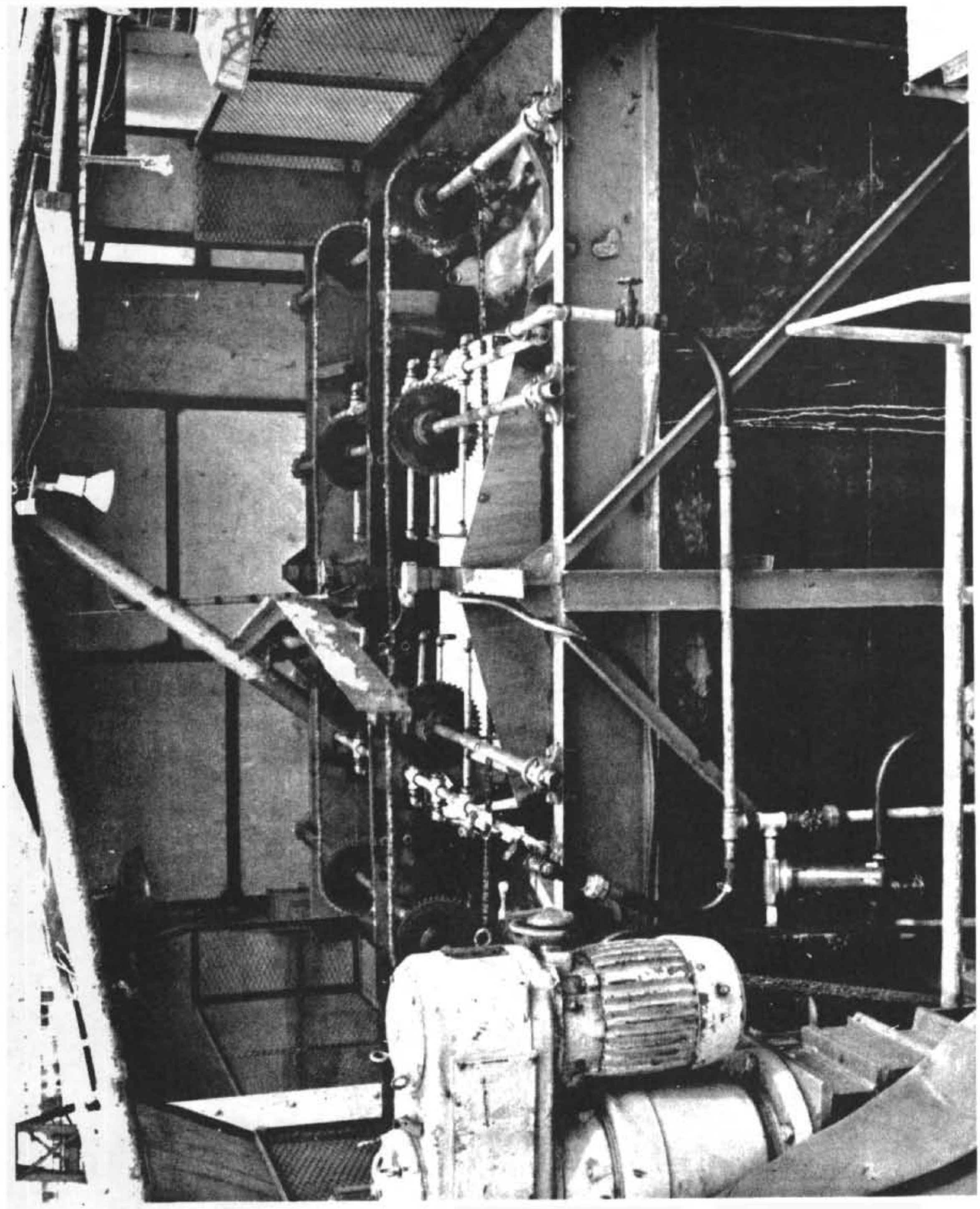


FIGURE 2: CHAIN DRIVE CONVEYOR TYPE SCRAPER

Approximately one half the water content of the original sea water is converted into ice and water vapor at this point. As one pound of vapor flashes from the falling slurry droplets roughly six pounds of ice are formed.

The ice slurry is pumped from the bottom of the freezer and most of it is recycled back to the standpipes, but a flow-controlled stream is introduced into the separation column which is rectangular in cross-section. The main walls are perforated to permit the drainage of brine as it leaves the piston of ice that has been formed from the slurry.

The ice continues to rise, being pushed from below, through a zone where it is washed by a small portion (0 to 10%) of the product water. At the column top the ice, washed free of adhering liquid brine, is harvested by a conveyor type scraper into the melt tank.

The melt tank is a square vessel surrounding the column where recirculated product water, which has been used as a coolant stream, melts the freshly harvested ice. The melt water is pumped from the bottom of the tank and splits three ways, the wash water portion going to the column top, the bulk of it recirculating through the absorption refrigeration machine as a coolant and the third part becoming the product stream. The salinity of the melt water circuit is continuously monitored.

The melt product water exchanges heat with incoming sea water and is mixed with the condensate product, metered, and checked for salinity before consumption or disposal.

The vapor stream which passes from the freezer top is conducted to the absorber where a lithium bromide solution contacts it and absorbs its water content. The diluted solution (weak liquor) passes through a surge tank from which it is pumped through a heat exchanger bank (H-7) to the generator-condenser vessel. Here the recirculated lithium bromide stream is reconcentrated by distilling off under vacuum an amount of fresh water equal to the original water vapor from the freezer.

This distillate, customarily called "process condensate" or simply "condensate" passes through a barometric leg into a float tank-condensate pump unit, through a rotameter, joins the melt product downstream of the H-3 exchanger but ahead of the product meter. Thus approximately one-sixth of the product is distilled water, the remaining five-sixths is melted ice.

Other streams deserving mention are:

(a) the underflow brine stream which may be withdrawn from the separation column bottom through manual valves to join:

(b) the brine stream from the sidewalls of the column that passes through swirl nozzles and a distributor ring to sluice the sidewalls of the freezer, provide additional mass transfer surface and prevent buildup of ice on the walls.

(c) the byproduct brine stream taken primarily from the column skirt by P-4, the brine recirculating pump, through the second feed pre-cooler (H-4) to be metered and discarded.

(d) the steam generated by the boiler which supplies high pressure steam (100 psig) to the ejectors and low pressure steam (7 psig) to the absorption system generator.

(e) the cooling sea water stream, only coarsely strained, that cools the absorption machine condenser, the auxiliary refrigeration condensers, the barometric condensers in the ejector systems.

(f) the auxiliary refrigeration circuit (R-22) which removes heat from the recirculated melt water to offset system heat gains.

(g) the regenerated lithium bromide absorbent which flows from the generator back through heat exchanger bank H-7 to the absorber where it is recirculated to maintain a low vapor pressure, the driving force for the freezing process.

3.0 PILOT PLANT OPERATION

The smooth, dependable operation of the pilot plant which was achieved during this contract period, facilitated more detailed investigation of various operating parameters and their effect on process efficiency. In many cases the tests were designed to determine the maximum product rate obtainable or the optimum value of some variable. To this extent then, the average performance for any given test may not be meaningfully compared with previous testing. Several tests were performed to demonstrate the value of the knowledge gained in these exploratory tests. Normally, the first day of each week was used to accomplish equipment modifications and maintenance. A test run normally comprised four days of continuous operation.

All testing from June 13 to October 5, 1962 was performed with the bolted panel rectangular column divided into two cells. This design was useful in the investigation of construction materials and coatings, and for study of modular control schemes. An extensive parametric study of the effect of slurry temperature and recycle ratio (ratio of slurry flowing through the freezer to slurry flowing to the separation column) on column performance was undertaken. An important aid in evaluating these results was the development of test equipment for detecting changes in the characteristics of the ice produced under the various operating conditions. The correlations obtained provided a basis for predicting column performance and greatly reduced the amount of pilot plant operation required for evaluation of these process variables.

The center partition was removed from the separation column during the week of October 8 - 12, 1962 to investigate the effect of increasing the horizontal distance between drain panels. All remaining tests during the contract period were performed with this single cell configuration.

In January 1963 an extensive modification of the separation column top was completed. The new installation featured a conveyor type scraper, with a design suitable for extrapolation to larger rectangular columns, and a melt tank surrounding, and integral with the separation column. Evaluation and development of this new arrangement proceeded to the end of the contract period.

A performance demonstration test was conducted during the last week of pilot plant operation. The operating parameters, component arrangement, and control scheme used in this test were chosen to demonstrate as many cost saving features investigated during the period as possible and excellent results were obtained.

Actual operation of the pilot plant was performed by two operator-technicians working as a team on each shift throughout the previous contracts and up until October, 1963. A total of six operator-technicians were required, therefore, to operate the plant on a 24 hour basis.

In October the operating crew was reduced to one operator-technician per shift or a total of three for 24 hours operation. This was made possible by the smoother, more dependable performance of the plant. This single operator-technician recorded all the necessary test data as well as operating the plant physically. Additional technical staff was used to calculate and analyse test results, and to plan the continuing program, etc.

4.0 PILOT PLANT PERFORMANCE

Although dependable operation of the pilot plant was demonstrated earlier in the program, optimization of the system and component design had not yet been attained. The pilot plant was specifically designed with separate components for individual steps in the process to facilitate evaluation. One obvious approach to economic improvement of the system design was through the integration of components into functional units. This design task was undertaken in Syracuse, New York, along with the evaluation of appropriate plant capacities for the vapor absorption process. This study was also closely coordinated with current developments at the pilot plant, and in many cases, the need for specific test information was pointed up by the design study.

The two main components in the system, the freezer and the separation column, still cannot be considered fully developed. Primary emphasis in the pilot plant program therefore was placed on testing concepts that might result in more efficient performance and/or reduction of capital costs for these components.

4.1 CONCEPTS CONTRIBUTING TO ECONOMIC IMPROVEMENT

In general, economic improvement may be obtained by increasing the cycle efficiency (energy costs), increasing component unit capacities (capital costs), decreasing component and control complexity (capital costs), and increasing automation (labor costs). It can readily be seen that these areas of improvement may overlap and that an appropriate overall balance must be determined.

The cycle efficiency is largely determined by the efficiency of heat transfer and the effective arrangement of equipment. To a great extent, these factors could be evaluated in design studies based on previous experience with the more conventional portions of the system. There were at least two important areas, however, that could only be investigated at the pilot plant. First, the cycle efficiency is directly affected by the net amount of wash water required to obtain an acceptable product. Good results at the end of the previous contract indicated the feasibility of limiting the net wash to no more than 5% of the gross product. Minimization of average net wash requirements was a continuing effort throughout the test program and a guiding parameter in evaluating changes in operating conditions. A second important factor in cycle efficiency is the concentration ratio through the process, or the per cent conversion of sea water to fresh water. As the per cent conversion increases, sea water feed and, therefore, theoretical steam and heat transfer requirements decrease. A consequence of greater concentration of the brine is the lowering of the operating slurry temperature. The effect of varying slurry temperatures on overall performance was investigated in a series of several tests.

The separation column is the most developmental component in the process and, as such, has the greatest potential for improvement and simplification. The evaluation of most test results ultimately revolves around the performance of this component. Specifically undertaken during this period was investigation of production capacity as a function of column height, width, drain plate design and ice bed characteristics such as porosity and permeability. Concurrently, simplification of the column design and control scheme was successfully undertaken.

The mechanical energy input to the system was reduced through the demonstrated ability to eliminate at least one pump in the brine circuit, and the mechanical agitation in the melt tank. Also a decrease in the power requirements for harvesting ice was demonstrated through use of a chain driven conveyor type scraper. More important in the case of the scraper modification was the development of a design for a rectangular column and particularly suited to larger sizes.

4.1.1 Final Plant Performance Demonstration Test Run

The final test run was conducted primarily to demonstrate the performance and cost saving features used in the design and cost evaluation of a 165,000 GPD plant. Specifically, the objectives were to attain an average production rate of at least 8.8 GPM (390 GPD/sq.ft.) at a net wash of less than 5% of the gross product. The test was performed at what has been judged to be the economic optimum slurry temperature (25^{o}F), a recycle ratio of 6.2:1, and without the use of brine underflow from the separation column. The stationary wash nozzle header was used and the harvested ice was melted without the aid of mechanical agitation. The column design features are discussed in Section 4.3.

The overall performance was excellent. The test was marred only by a 24 hour interruption caused by failure of a weld on one of the scraper blade support brackets. The average product rate goal was exceeded, salinity was maintained below 400 ppm at all times and the net wash was the lowest ever attained at the pilot plant over an extended test period. Column performance and the ice rise rate were exceptionally smooth, demonstrating beyond doubt the ability to operate the column without brine underflow. Summary data for the test are given in the table below and the half hourly production chart for the first 24 hours is shown in Figure 3.

TEST NO.	30403
Duration of Test - Hrs.	81
Ave. Slurry Temperature - °F	25.2
Recycle Ratio	6.2:1
Average Production - GPM	9.43
GPD/sq.ft.	419
Average Salinity	236
Per Cent Time Above 400 ppm	0
Average Net Wash as % of Gross Product	2.36
Average % Ice in Slurry	9.9
Highest 8 Hour Production - GPM	10.1
Highest 24 Hour Production - GPM	9.76
Production Stability:	
Per Cent Time, 1/2 Hr. Product Rate = Ave. ± 1 GPM	65
= Ave. ± 2 GPM	88

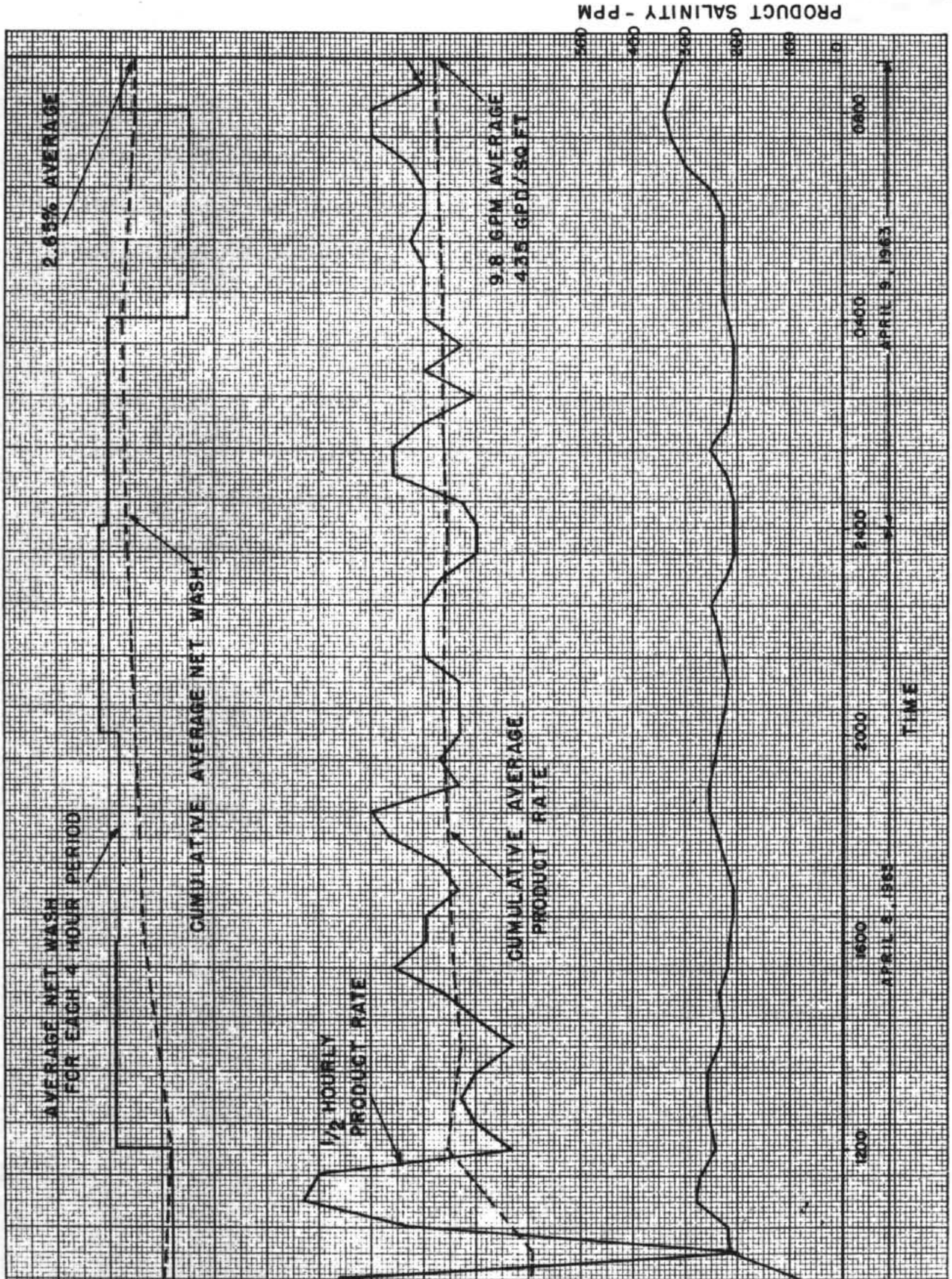

PRODUCT SALINITY - PPM
PER CENT NET WASH
PRODUCT RATE - GPM
AVERAGE NET WASH
FOR EACH 4 HOUR PERIOD
CUMULATIVE AVERAGE NET WASH
2.65% AVERAGE
1/2 HOURLY
PRODUCT RATE
CUMULATIVE AVERAGE
PRODUCT RATE
9.8 GPM AVERAGE
435 GPD/SQ.FT.
0800
1200
1600
2000
2400
0400
0800
APRIL 8, 1963
APRIL 9, 1963
TIME

FIGURE 3

4.2 FREEZER

The freezer configuration at the beginning of this contract was the same as that at the end of the previous contract and is shown in view "H" and drawing RD1039-9151. Slurry was recirculated in the freezer through twelve 1-1/2" standpipes each 48" high.

4.2.1 Effect of Slurry Temperature

The series of tests on varying the ice slurry temperature showed that this operating parameter has a very significant effect on column performance. The product rate at which the salinity could be controlled was reduced as the slurry temperature was reduced. Also, the net wash requirements increased at the lower temperatures. The method of operating the plant and all operating parameters other than slurry temperature were the same for Test Runs 29321 and 20827. No adverse affects were noted in actual freezer performance.

In order to make a meaningful comparison between maximum possible production rates, it has been found necessary to compare time periods considerably shorter than the complete test run. A period of 8 hours has been chosen as a standard since it represents several turnovers of the ice volume in the column and allows adequate time to determine a salinity trend. The maximum production as used in this report is defined as the highest 8 hour average rate obtained at a steady salinity at or below 400 ppm.

The maximum sustained production rate at each slurry temperature was estimated from these tests and is given in the table below.

TEST NO.	20827	20821	20827	20827
Slurry Temperature - ^{o}F	24	25	26	27
Recycle Ratio	4:1	4:1	4:1	4:1
Maximum 8 Hour Ave. Production Rate with Salinity Control: GPM	6.2	7.3	7.7	8
GPD/sq.ft.	318	375	395	410
Per Cent Net Wash at Maximum Product Rate	16	8.5	2	1
Product to Condensate Ratio:				
Corrected to Zero Net Wash	6.40	6.16	5.66	5.32
Actual - Test Run	5.16	5.70	5.35	5.16

The table also shows the gross product to condensate ratio for each temperature, which is a measure of the steam required to produce a given product. While column capacity increased at higher slurry temperatures, the energy required per unit production was also increased. Also, the size of the auxiliary equipment connected with the feed and brine systems must be increased as the freezing temperature is increased for any given product rate. Conversely, the expected advantages of lower slurry temperatures were partially offset by poorer column performance. The data does provide a basis, however, for arriving at the most appropriate design operating temperature.

These tests also served to emphasize the desirability of determining what process characteristics were changed with changes in the slurry temperature. One possibility was in the purely mechanical functioning of the separation column. The colder temperatures could aggravate some previously noted tendency for drain holes to freeze over causing poor wash water drainage. The problem was more likely due to a basic change in the character of the ice crystals produced. The investigation proceeded along both lines.

4.2.2 Porosity, Permeability and Relative Particle Size

During this report period techniques were developed for obtaining permeability and porosity data from an ice bed formed from a sidestream of slurry directly from the freezer. These data can be related to an equivalent diameter particle size as described in Appendix B.

The equivalent diameter concept has proved extremely useful in evaluating changes in ice characteristics although no basis has been found for relating this value to actual measured crystal size, or distribution of crystal sizes, in a given sample. Also, attempts to equate the equivalent diameter of these ice particles, directly from the freezer, with porosity and apparent permeability in the separation column, have been unsuccessful to date.

As previously noted the salinity controlled production rate was decreased as the slurry temperature was decreased. Subsequent tests showed a detectable decrease in the calculated value of the ice particle equivalent diameter "d_e" as the slurry temperature was decreased. This correlation prompted investigation of another operating parameter, recycle ratio, which was also expected to have an effect on "d_e".

Test runs No. 20910 and 20918 were scheduled specifically to obtain permeability and porosity data at various slurry temperatures and recycle ratios (ratio of slurry flow return to the freezer, to slurry flow to the separation column). The data from all these tests indicated that changes in the characteristics of the ice produced could be detected by permeability and porosity measurements and correlated with changes in slurry temperature and recycle ratio. Test data relating the equivalent diameter "d_e" to slurry temperature, recycle ratio and product rate (GPD/sq.ft.) are presented in the Appendix B.

4.2.3 Effect of Slurry Recycle Ratio

Test runs No. 20924 and 21002 were performed to demonstrate a predicted increase in salinity controlled production rate with a higher recycle ratio. The following table summarizes and compares the results with previous tests at lower recycle ratios.

TEST NO.	20924	20821	21002	20327
Slurry Temperature - °F	25	25	26	26
Recycle Ratio	5.2:1	4:1	6:1	4:1
Maximum 8 Hour Av. Production Rate with Salinity Control: GPM	8.0	7.3	9.5	7.7
GPD/sq.ft.	410	375	486	395
Per Cent Net Wash	4.6	8.5	5.3	2.0

Again the effect of slurry temperature is noted as higher production rates were possible at 26°F as compared to 25°F. Of greater significance in these tests was the demonstrated ability to increase production at any temperature by increasing the slurry recycle ratio, as predicted. At 25°F the product rate was increased from 375 GPD/sq.ft. to 410 GPD/sq.ft. by increasing the slurry return to the freezer from 4 times to 5.2 times the slurry flow to the column. Production at 26°F was similarly increased from 395 GPD/sq.ft. to 486 GPD/sq.ft. by increasing the recycle ratio.

4.2.4 Performance with Dense Slurries

As might be expected, increases in recycle ratio are subject to an economic limit. The ratio can be increased by increasing the return flow of slurry to the freezer, by decreasing slurry flow to the separation column or a combination of both. Horsepower requirements for a slurry recycle pump increase rapidly as the flow is increased and substantial recycle flow increases are required at a constant slurry flow to the column to obtain a significant increase in the ratio. Decreasing slurry flow to the column offered greater potential of an overall economic gain. Initial tests at low slurry flows resulted in freezer operating problems due to the increase in slurry density (higher weight per cent ice in slurry). As the freezer design evolved to the use of 12 overflowing standpipes (View H, Dwg. RD1039-9151) much of the original manifold was blanked off leaving unnecessary obstruction in the bottom of the freezer. The denser slurry tended to plug around this manifold.

The manifold was redesigned to distribute the slurry to the twelve standpipes with the simplest piping arrangement possible. The problem of balancing flow through the twelve standpipes was more critical than expected. Several modifications were made in December, 1962 before a successful design was achieved and tested in test No. 21226.

One of the arrangements tested is shown in View "I" on drawing RD1039-4197. This was not completely successful because it was necessary to operate with a high liquid level to keep the manifold piping submerged. This in turn prevented adequate agitation of the slurry with consequent blockage of flow.

The final arrangement tested is shown in View "J" of drawing RD1039-4197 in which the minifold and liquid level was lowered about 13". Slurries containing up to 15% ice by weight were handled without difficulty with this arrangement. The normal recycle ratio used in subsequent tests was about 6:1 giving a slurry density of about 9% to 12% ice.

4.2.5 Effect of Standpipe Height on Capacity

In developing a freezer design for larger capacity plants, a horizontal vessel was found to be most practical. Also, as discussed later in the report, the LiBr absorber was to be incorporated in the same freezer vessel. In order to avoid protrusions or extensions of the shell, the height of the standpipes in the freezer section would have to be limited to 30 inches above the slurry level. This would result in a decrease in mass transfer area and could possibly reduce ice production below the design value.

The original standpipes used in the pilot plant freezer extended approximately 42" above the liquid level. Just prior to test 21217, the standpipes were shortened by one foot and more than adequate production capacity was maintained. There was a noticeable depression in the freezer wet bulb temperature, 1.5 - 2°F below the slurry temperature, compared to previous test runs where the average depression was 1.0 - 1.5°F. This indicates that a slight increase in driving force was required due to the decrease in transfer area.

The increase in driving force in the freezer might be expected to have some effect on the ice particles produced. However, no detectable difference in the porosity and permeability of the ice could be found. It was concluded that this change in standpipe height would have no adverse effects either on production capacity or separation column performance.

4.2.6 Sidewall Wash Ring

The freezer sidewall wash ring was designed to sluice the sides of the freezer and keep them free of ice. Swirl jet nozzles were added prior to this contract period to improve the distribution of the sluice water. Other than periodic cleaning and resetting of the gap around the ring, little else has been done during this period.

The present arrangement adequately sluices the sidewall. However, the design should be improved to reduce the clogging problem and also to eliminate spurious sprays that seem to be a cause of carryover and icycle formation in the dome of the freezer and around the vapor baffles. The design problem should be simplified in a horizontal freezer, although the continuing problem at the pilot plant would be considered seriously in any new design. The majority of slurry plugs in the freezer and slurry piping were probably the result of solid ice falls from the dome of the freezer.

4.3 SEPARATION COLUMN

At the beginning of the contract period, the separation column was divided into two individual cells, with bolted panel construction as shown in view 11 on drawing RD1039-9149. At the present time, it is not considered feasible to increase separation column capacity indefinitely by simply scaling up the physical dimensions to provide the required cross section area. Two approaches to increased capacity for larger plants have been tried. The first involved installation of intermediate partial drain partitions within a single larger column. Some problems were encountered that, in the light of more recent information, may be eliminated with more appropriate materials of construction (eg. PVC). The problems were in fact resolved by complete separation of the column into cells and the development of a control scheme for operation of multiple separation cells in parallel from a single freezer.

The bolted panel column shown in view 11 has a wide center partition and a reduction in cross section due to the wall linings. The original round column had a cross sectional area of 44 sq. ft. The square column had an area of 36 sq. ft., and the column with divided cells had an area of only 28 sq. ft. After removing the center partition but retaining the wall liners, the area was about 32.4 sq. ft. All other factors being equal the capacity of the column should be approximately proportional to the area. The original design capacity was 340 GPD/sq.ft., and the performance of the modified columns is most appropriately judged in comparison with the figure even though the actual plant capacity may have been decreased.

With operational problems and a concept for design extrapolation largely resolved, more emphasis could be placed on increasing the production rate per sq. ft., decreasing control complexity, and investigating the extent to which an individual separation cell may be economically enlarged. In so far as practical, tests were performed at the pilot plant to correlate production rate with column height and width. The effect of various drain plate designs on washable product rate was also investigated. Concurrent with product rate testing, various construction materials and coatings were evaluated.

4.3.1 Bolted Panel Column with Drainage Partition

The two cell bolted panel construction is shown in view 11 on drawing RD1039-9149. This type of construction was chosen to facilitate testing of materials and coatings and proved to be one of the most significant advances made in the separation column for pilot plant work. The PVC plastic sheet material, originally installed in only the upper section of one cell, was shown to have superior friction and ice release properties that greatly improved column performance. By the beginning of this contract period, the east and west walls of both cells, including the drain plate areas were covered with 1/4" PVC sheet. Although not of bolted panel construction, the north and south walls were also covered with PVC sheet early in September, 1962. This completed covering the entire friction area of the ice piston with the exception of two 3' x 2' coated steel test panels. The much improved column performance led to simplification of the control scheme. A pulsing effect produced by solenoid valves in the column underflow, which was previously shown to be advantageous, became unnecessary. The control instrumentation required for operation of two separation cells in parallel was thus reduced to individual flow control of slurry to each cell with simple orifice control of the brine underflow. This is illustrated in the table below by comparing the product rate variation with the average product rate for each test.

TEST NO.	20821	20910	20924	21002
Column Cells	2	2	2	2
Hours on Stream	79.5	63	87.5	78
Slurry Temperature, $^{\circ}F$	25	26	25	26
Slurry Recycle Ratio	4:1	4:1	5.2:1	6:1
Underflow Control	Solenoid	Solenoid	No Flow	Orifice
Average Product Rate, GPM	6.37	6.83	7.44	8.51
Average Product Rate, GPD/sq.ft.	353	351	383	438
Average Salinity, PPM NaCl	217	220	365	171
Per Cent Time, ½ Hr. Product Rate - Ave. ± 1 GPM	57.1	71.3	66.3	70.2

PVC was added to the north and south sidewalls just prior to test 20910. Smoother performance is evident when compared to test 20821. Test 20924 was performed with no underflow primarily to obtain a higher recycle ratio. This test was very successful and, although production was somewhat more erratic, the possibility of complete elimination of underflow was demonstrated. With orifice controlled underflow in test 21002, production was as smooth as that of any previous test using solenoid pulsed underflow.

In addition to smoothing production through the column, brine underflow can be used to facilitate control of the per cent ice in the slurry feed to the column (slurry density). Thus, if product rate is limited by the density of the slurry that can be handled in the freezer or connecting piping, the density may be reduced by increasing slurry flow and adding brine underflow without affecting the performance of sidewall drains. This flexibility is particularly desirable in a test plant, but additional testing without underflow has indicated that this piping may be eliminated from a production plant with significant savings in capital costs.

4.3.2 Undivided Bolted Panel Column

During the week starting October 8, 1962, the center partition was removed from the separation column as shown in view 12 on drawing RD1039-4196. This resulted in a single larger cell approximately 5'-7" by 5'-11" in cross section with an ice area of 32.4 sq.ft. The distance between sidewall drains was increased from 28" to 67". The increased length of the wash water drain path would be expected to result in a decrease in washable product rate. It was expected that the extent of the decrease in product rate per square foot would facilitate evaluation of the economics of larger size separation cells.

4.3.2.1 Effect of Increased Distance Between Sidewall Drains

The initial tests with the larger cell showed an almost negligible decrease in allowable product rate. The table below compares maximum production rates for the two column configurations.

TEST NO.	20924	21002	21022	21029
Column Area - sq.ft.	28.0	28.0	32.4	32.4
Column Configuration	2 cells	2 cells	1 cell	1 cell
Distance Between Drain Plates - in.	28	28	67	67
Slurry Temperature - ^{O}F	25	26	25	26
Recycle Ratio	5.2:1	6:1	5.5:1	5.5:1
Maximum Sustained Product Rate with Controlled Salinity				
GPM	8	9.5	9	10.5
GPD/sq.ft.	410	486	400	465
Net Wash - % of Gross Product	4.6	5.3	7.0	5.6

It can be seen that although the drain plate distance was more than doubled, the production capacity per sq.ft. was decreased by less than 4 per cent. Also, with both configurations, there was roughly a 16 per cent increase in production at the higher slurry temperature with similar recycle ratios. These results indicate that additional savings in column construction costs may be obtained through even greater increases in the cell size. Unfortunately, it was not practical to increase the width of the pilot plant column beyond 67 inches.

4.3.2.2 Relationship of Effective Washing Height to Production Rate

Several tests were performed in which the effective washing height of the ice column was varied by changing the location of the sidewall drain area and the elevation of the brine level in the column skirt. The results are shown in the table below.

TEST NO.	21113	21108	21022	21126
Effective Washing Height - ft.	6.5	6.5	5.5	4.5
Recycle Ratio	7.5	5.6	5.5	6.0
Maximum Sustained Product Rate with Controlled Salinity				
GPM	9.3	9.05	9	8.0
GPD/sq.ft.	414	403	400	355

The recycle ratio is included in this table only to avoid drawing an erroneous conclusion from the higher product rate in test 21113. In this test, underflow was not used and the slurry flow was reduced accordingly. With this amount of increase in the recycle ratio, the production increase is actually less than would be expected. Therefore, the increase in column height is considered to have had little or no effect on the maximum allowable production rate. Decreasing the column height had a rather pronounced effect as a decrease in product rate of about 11 per cent was noted. When all factors are considered, it is reasonable conclusive that an increase in column height for the present drain plate spacing would not allow an appreciable increase in production but a decrease in height would decrease production. There appears to be a point at which the product rate approaches a maximum, unaffected by further increases in column height.

The effect of increasing the distance between drain plates beyond 67" is not expected to be limiting. The tests above indicate that the column height was greater than required for the 28" drain plate spacing, and that most of this excess height had been used by moving to the 67" spacing. The column height may be considered fully utilized when the resistance to wash water flow from the center to the sidewall drain increases to the point where additional head (column height) is required to obtain a positive net wash flow, in the center of the column. From this point on an increase in drain plate distance will require an increase in column height to obtain maximum production. However, it is expected that this maximum production rate can be obtained for any reasonable increase in drain plate spacing by increasing the column height.

The difference in the effects of column height and width on production rate can be visualized by considering the controlling factors in vertical and horizontal flow of wash water. Vertical or downward flow is controlled by the head available, the flow resistance of the ice bed, and the ice velocity up the column (production rate). Assuming a constant flow resistance, the ice velocity can be increased until the head loss per foot length of bed is equal to one foot. Any further increase in ice velocity will result in a negative wash rate and loss of salinity control. Additional column height, while increasing the total head available, increases the head loss proportionally. For a given flow resistance in the ice bed (governed by permeability), the production rate must approach a maximum unaffected by further increases in column height. This effect is shown graphically in Figure 4 using the data presented above for the 67" drain plate spacing.

Horizontal flow of wash water is controlled by essentially the same ice bed resistance and the length of the flow path, but would not be expected to vary with the vertical ice velocity. Theoretically, then, an increase in horizontal flow resistance caused by increasing the length of the flow path (increasing distance between sidewall drains) can always be compensated by increasing the available head (column height). The small amount of data available tends to support this analysis but is not sufficient to establish an experimental height vs width relationship that can be extrapolated with confidence.

Some subsequent tests have been performed which indicate that higher product rates are possible under certain conditions that probably affect the permeability of the ice bed by increasing the porosity. Ice bed porosity is discussed in the following sections of the report. Also, this concept, while useful, cannot take the place of more detailed knowledge of the mass transfer and flow phenomena occuring in the separation column. At this writing there does not appear to be a model available which appropriately describes the test results.

4.3.3 Drain Plate Design

The bolted panel construction of the column provided an extremely flexible unit for experimenting with drain plate designs. At the beginning of this period, there were eight drain plates, 46" high by 34½" wide, two

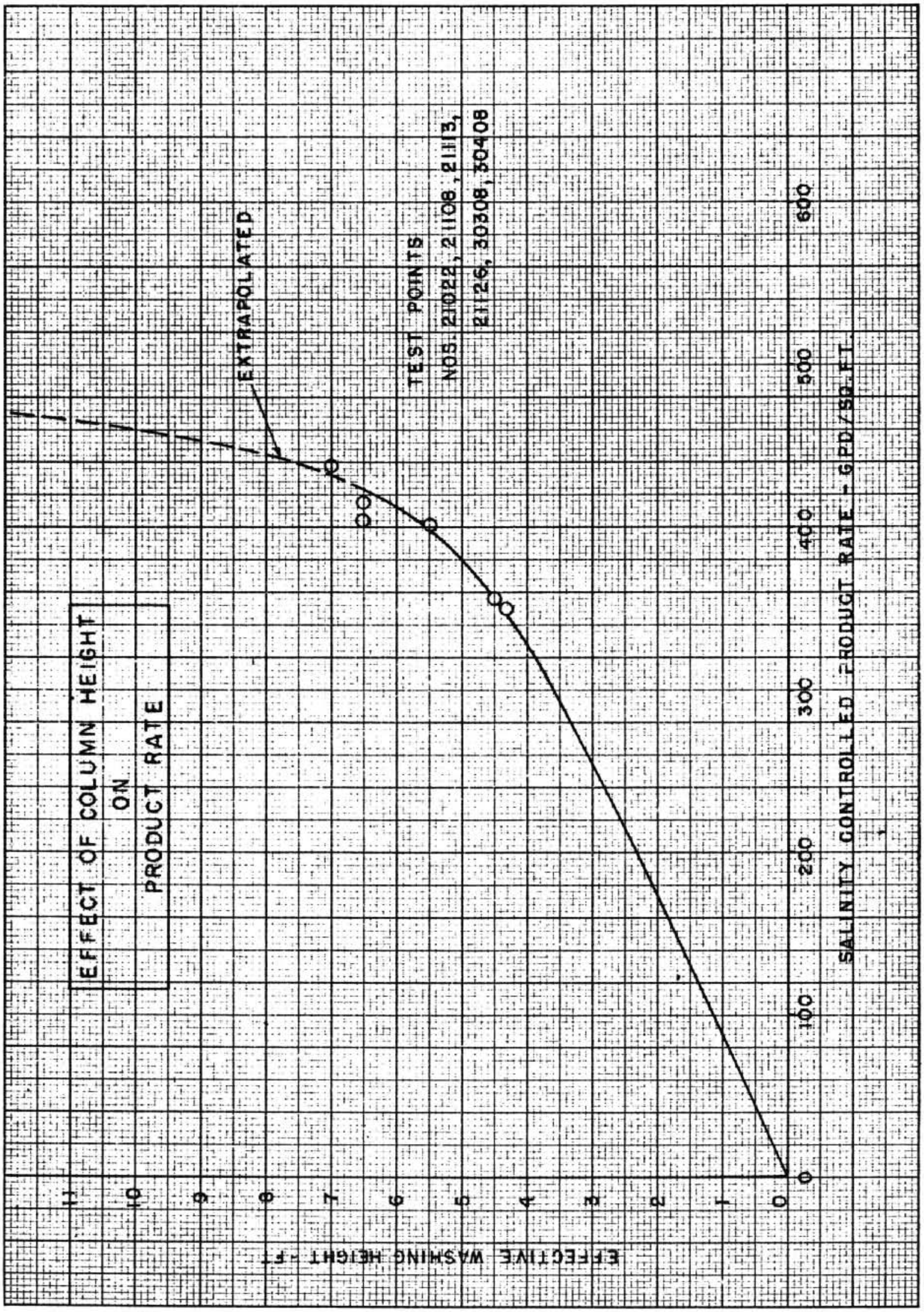

EFFECT OF COLUMN HEIGHT
ON
PRODUCT RATE
EXTRAPOLATED
TEST POINTS
NOS. 21022, 21108, 21113,
21126, 30308, 30408
EFFECTIVE WASHING HEIGHT - FT
0
1
2
3
4
5
6
7
8
9
10
11
SALINITY CONTROLLED PRODUCT RATE - GPD/SQ.FT.
0
100
200
300
400
500
600

FIGURE 4

installed in the east and two in the west sidewalls of each cell. The total number of drain plates was reduced to four with the removal of the center partition. Depending on the drain hole configuration desired, many changes could be made with the plates installed, with the assurance that an inappropriate change could be easily rectified by installing new plates. Probably the most significant innovation was the replacement of steel plates with PVC plates just prior to this contract period. Not only did these plates reduce friction on the ice piston, but the problem of drain holes freezing over was virtually eliminated.

The primary function of sidewall drains is to provide a passage for brine and wash water out of the column and to promote proper formation of the ice bed. The original emphasis in design was to eliminate ice slurry breakthroughs and to determine the gross flow area required to assure that the resistance through the drain plates would not be a limiting factor in wash water flow.

4.3.3.1 Drain Plate Flow Area

Insufficient flow area in the drain plate design will increase the flow resistance and the head required to accomplish brine and wash water removal from the ice bed. The resulting higher level of the brine crown in the ice bed reduces the effective washing height in the column and limits the salinity controlled product rate. At the beginning of this period the total drain area was 42.8 sq.in., 5.36 sq.in, in each of eight plates. With the center partition removed, the total drain area was reduced slightly to 37.3 sq.in., 9.33 sq.in. in each of four plates. As previously noted, there was no appreciable decrease in product rate after this change. In prior testing, an increase in product rate was noted as the total drain area was increased from 26.8 sq.in. to 58 sq.in. However, some of these tests were conducted with steel drain plates and the results may have been distorted by partial freezing of drain holes. Drain areas up to 19.0 sq.in. per plate (76 sq.in. total area) were used during this period with no effect on product rate. It is therefore concluded that a total drain area of only 43 sq.in. is entirely adequate for this column design.

4.3.3.2 Drain Plate Hole Size

An appropriate choice of drain hole side is an extremely important factor in column performance. Normally, hole sizes in the range of 3/16" diameter had been used in the column to avoid plugging and freezing over. In the bolted panel column, the original drain holes in the steel plate were 0.104 in. diameter. The total drain area of the plates was increased by enlarging holes near the top of the plates in an attempt to increase the production rate. Although production was increased, there were still some indications of progressive freezing of the small holes in the plates. The original PVC drain plates had mostly 0.104 in. diameter holes and actually a decreased total flow area. Column performance was greatly improved, both in regard to washable product rate and to the almost complete elimination of the ice slurry break-through problem. Apparently, freezing of the small holes in the steel plates also led to slurry break-through at the larger holes. The 0.104 in. diameter hole size was retained in the lower section of the PVC drain plates in all subsequent designs. In some cases, hole sizes as large as 3/8" diameter were used in the upper portion of the drain plates to obtain large flow area in a shorter vertical section.

4.3.3.3 Drain Plate Design and Column Porosity

The packing effect of flow through an ice bed was studied using the permeability porosity test equipment. The results in some flow ranges were somewhat erratic, probably due to limitations in facilities at the pilot plant. However, a more or less stable packing (porosity 0,5 cc/cc) was considered to occur at flows above 7 gpm/sq.ft (Figure 5). Higher porosities were obtained at flowrates below 7 gpm/sq.ft. of ice bed. The packing obtained with flow between 4 -7gpm/sq.ft. appear to be less stable. It will be noted on Figure 5 that several points have been plotted indicating actual porosities found at the top of the separation column during test runs. The actual flow area involved in these tests is not precisely known, and, of course, it varies throughout the bed cross section. The estimates used here are based on the total flow of concentrated brine and the area of the drain plate actually covered with drain holes. When the drain hole flow area is large, the upper portions are probably ineffective. Perhaps, the most important conclusion to be drawn is that the porosity can be affected by drain plate design and brine flow rate, and that effects other than flow density change the porosity in the separation column. The gravitational force due to the weight of the ice column is probably the most important factor. Consideration of porosity is very important in optimizing column capacity. The theoretical harvest rate vs. ice partical size at a constant zero net wash (Figure 6) illustrates the relatively great effect porosity can have on allowable production.

4.3.4 Total Functional Column Height

The previous sections have discussed the effective washing height required and the considerations involved in establishing the drainage height for the separation column. The third function of the column is ice bed formation, involving inlet slurry distribution and brine underflow.

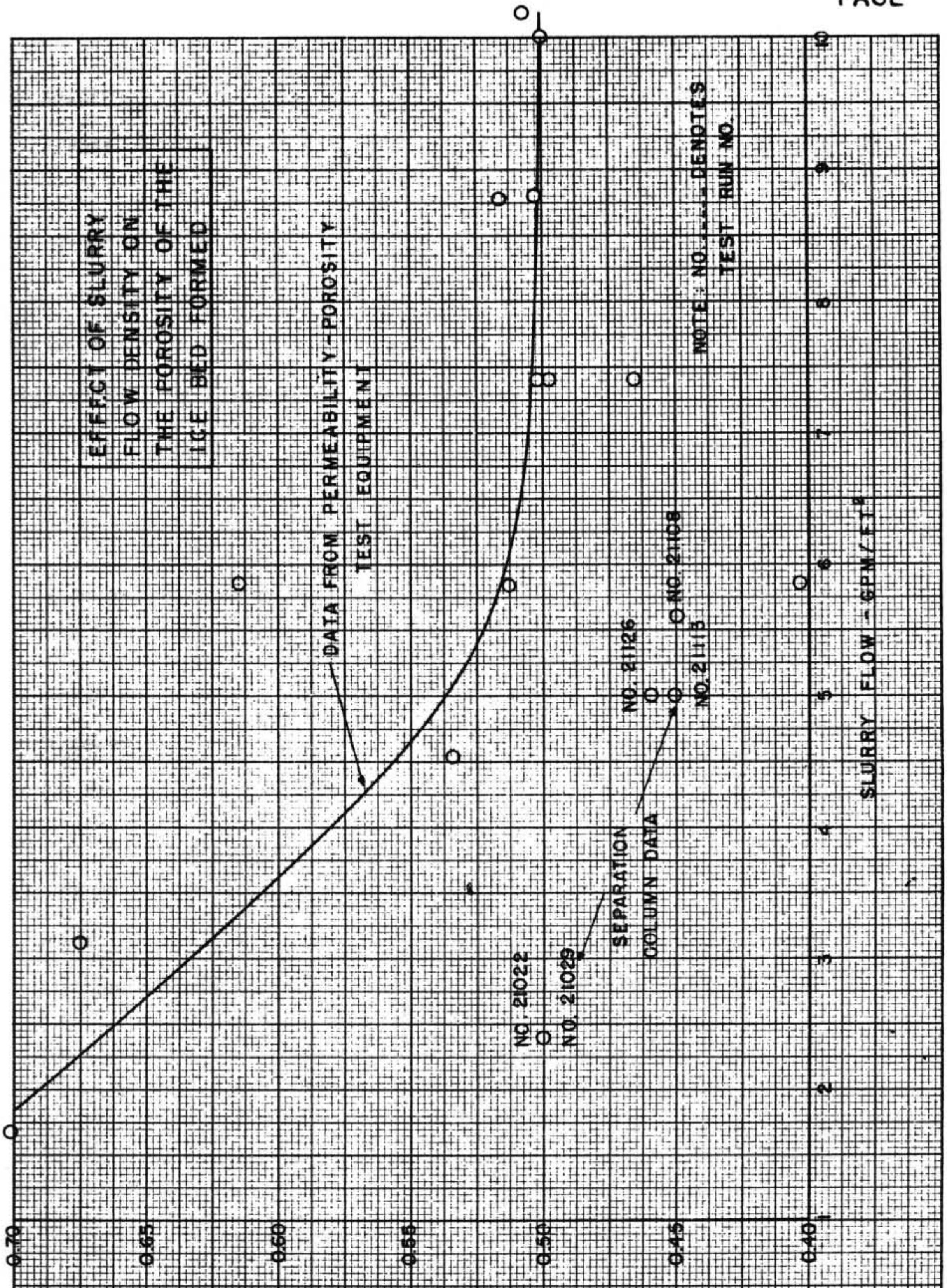

EFFECT OF SLURRY
FLOW DENSITY ON
THE POROSITY OF THE
ICE BED FORMED
DATA FROM PERMEABILITY-POROSITY
TEST EQUIPMENT
SEPARATION
COLUMN DATA
NOTE: NO. ----- DENOTES
TEST RUN NO.
NO. 21022
NO. 21029
NO. 21126
NO. 21108
NO. 21113
0.70
0.65
0.60
0.55
0.50
0.45
0.40
POROSITY - CC/CC
1
2
3
4
5
6
7
8
9
10
SLURRY FLOW - GPM/FT²

FIGURE 5

Normally 5 to 6 feet of column height was used for bed formation below the sidewall drain area. In February, 1963, the slurry inlet arrangement was modified as shown in view 13 on drawing RD1039-4196 so that the discharge was raised to within 26 ins. of the bottom row of sidewall drain holes. Also, the length of the expanded pipe section used to reduce the entrance velocity was cut from 20 ins. to 8 ins. This resulted in a simulated reduction in column height of about 24 ins. with the slurry distributor inside the column, or 36 ins. if the distributors were located below the bottom of the column. Very significantly, no adverse effects on bed formation or column performance were noted. At least during all subsequent tests where brine underflow was not used, the results are indicative of the performance of a column with a total height of only 11 feet.

In all arrangements investigated to date, the use of brine underflow has occasioned the formation of ice on the column bottom. In some cases the buildup was somewhat self limiting and no serious effects on operations occurred during the four day test periods. In other cases, the buildup of ice eventually choked the slurry supply nozzles and necessitated at least a temporary shutdown. This problem can probably be eliminated by some combination of piping arrangement and brine flow volume. However, this presents another good reason for eliminating underflow entirely.

The total column height required in designs using underflow may be greater. An optimum arrangement for brine underflow nozzles had not yet been developed by the end of this period. The most recent tests did indicate that the underflow outlets could be placed below the slurry inlets without carrying ice particles out of the column. Considering this relationship between the height of the slurry inlets and the underflow piping, at least 2 feet can be cut off the bottom of the column without affecting overall performance.

LIMITING ICE PRODUCT RATE
GRAVITY FLOW COUNTERCURRENT
WASH SEPARATION COLUMN

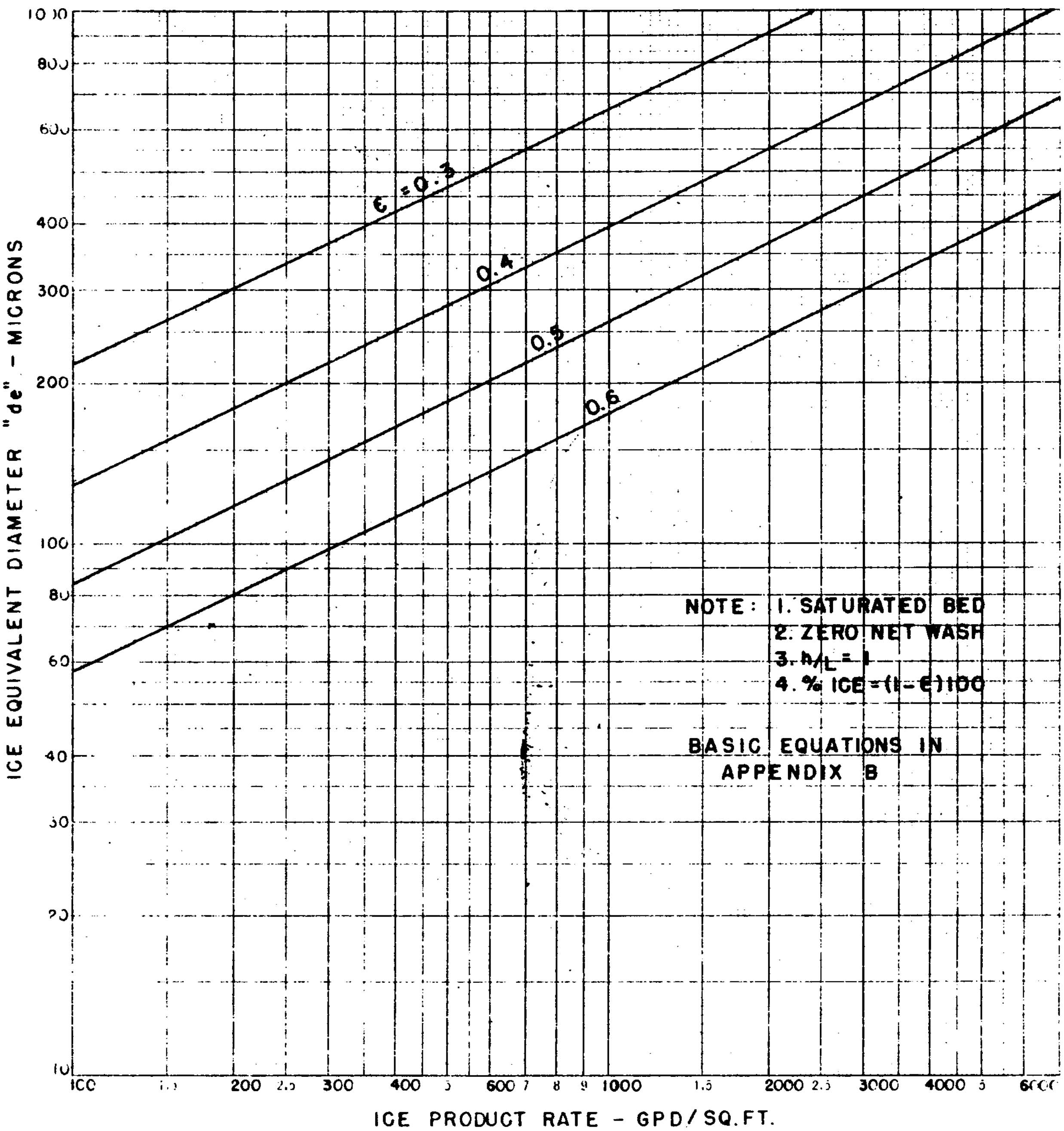


FIGURE 6

4.4 SCRAPER ASSEMBLY

Throughout most of the pilot plant program a two-bladed rotary scraper was used to harvest the ice product. This scraper assembly, originally installed in Syracuse, proved highly dependable and facilitated investigation of the more highly developmental features of the plant. With the development of the rectangular separation column, the possible advantages of using an endless chain conveyor type harvester were recognized. The inefficiencies of the rotary type scraper became particularly evident in design studies on larger rectangular separation columns.

4.4.1 Chain Driven Conveyor Type Scraper

In January, 1963, a chain driven conveyor type scraper was installed at the pilot plant. The general arrangement is shown in view 13 on drawing RD1039-4196, and a photograph is included as Figure 2. The construction details are shown on drawing R1039-S14-180. The first mechanical checks of the new scraper were started on January 17, 1963. Several modifications were found necessary in subsequent tests, but in general, satisfactory performance was obtained by January 28, 1963.

Initially the scraper blades were hinged in the center to facilitate changes in the cutting angle. It was expected that there would be advantages to accomplishing some ice removal at the sides of the column. In practice this was difficult to achieve, and no problems were encountered even though most of the ice was deposited at one end of the column. The blade angle was fixed rigidly in a later modification.

The most significant change required was the addition of a center and two side roller tracks to establish a rigid cutting position for the blades as they passed across the ice bed. The experience gained in this area should facilitate significant improvements in any future design.

The chain drive proved to be entirely adequate and trouble free. Also, the elimination of a portion of the reduction gear train used in the rotary scraper resulted in a reduction in load on the drive motor. The motor used had a nominal capacity of 5 horsepower. The normal load with the rotary scraper was about 2.5 - 3 horsepower. This was reduced to about 1.7 horsepower with the chain driven scraper.

4.5 MELT TANK

Along with the scraper modification in January, 1963, a square melt tank was constructed around the top of the separation column. This type of design incorporates the separation column, the brine surge tank, and the melt tank into a single compact unit, which could decrease overall construction costs. Drawing R1039-S14-180, view 13 on drawing RD1039-4196 and Figure 2 also show this new melt tank.

Prior to this change, testing was underway to prove the feasibility of providing sufficient agitation in the melt tank to melt the ice product by directed flow of the relatively warm melt water returning from the absorber. The success of the square melt tank arrangement depended to a great extent on the practicality of eliminating mechanical agitation, which was definitely proven in subsequent tests. The primary requirement was found to be maintenance of sufficient horizontal motion on the surface of the melt water to prevent agglomeration of the harvested ice. The minor amount of restriction used in the melt water lines to increase the discharge velocity, should not result in a significant increase in the pump power requirements. Since melt water flow at the pilot plant was already throttled, the increase in pump discharge head, if in fact there was an increase, was not detected.

4.6 Wash Water Distribution System

The installation of the new scraper required development of a new system for the distribution of wash water on the ice bed. Wash nozzles had been incorporated in the arms behind the blades of the rotary scraper. Some of the advantages of this system had not been fully recognized prior to the modification. The nozzles were spaced approximately two inches apart and the wash water flow could be varied over a wide range without appreciably affecting the pattern of distribution. Early attempts to duplicate this distribution with a stationary header design failed primarily due to the obstruction presented by the large bull gear of the rotary scraper, over the center of the column.

4.6.1 Rotary Wash Headers

The original design of the conveyor type scraper provided for introduction of wash water through nozzles in the hollow rotating idler shafts in the scraper drive train over the column top. Flat spray nozzles were used in each of the two idler shafts to effect wash distribution. As the shafts rotated, wash water was sprayed through an arc of about 100° controlled by a cam actuated quick opening and closing valve. The initial setting and synchronization of this system was difficult. Performance data from test runs using this system are given in the table below.

TEST NO.	30128	30205
Duration - Hrs.	31	80
Wash Distribution Method	Rotating Header	Rotating Header
Ave. Product Rate -		
GPM	8.3	8.03
GPD/sq.ft.	369	357
Ave. Salinity - PPM	1660	770
Ave. % Net Wash	-2.0	5.5
% Time Over 400 PPM	65	44
Highest 8 Hr. Salinity Controlled Product Rate		
GPM	7.8	8.37
GPD/sq.ft.	346	371

In test 30128 wash water was applied through 12 flat spray nozzles, 6 in each of the two hollow idler shafts of the scraper drive assembly. A green dye was injected into the freezer slurry system as an aid in evaluating the effectiveness of wash distribution. The dye, being dissolved in the concentrated brine, gave visual evidence of insufficient washing and pointed out areas of poor wash distribution. Localized areas of high salinity were evident during this test at relatively low product rates (6.5 - 7 GPM). Ineffective washing resulted in the high average salinity of the product noted in the summary data. Near the end of the test, several wash nozzles were replaced with a larger size and an 8 hour run at a product rate of 7.8 GPM with acceptable salinity was obtained.

Prior to test 30205, the total number of nozzles was increased to 16 and larger nozzles were used near the center section of the column. The product rate was then increased to about 8.4 GPM with salinity control.

Further improvements in the cam actuated rotating wash distribution system are considered possible. However, the quick opening and closing valve, and the cam operating mechanism present a hazard from the standpoint of reliability and continuous operation. Therefore, a more reliable stationary type wash header was tried since this became feasible with the conveyor type scraper.

4.6.2 Stationary Wash Nozzle Header

The stationary header was designed to provide approximately one nozzle per square foot of column cross section. A total of 36 nozzles was used. The nozzles were of the hollow cone type and necessarily smaller than those used in the rotating header to assure full development of the spray pattern at low flow rates. The smaller nozzles tended to plug more easily and seriously affected wash distribution in the first test. This situation was alleviated by changing the strainer basket in the wash line and the problem was completely eliminated in later tests by installing a new basket strainer with a 100 mesh screen.

Steady progress was made during the tests with a stationary system. The most significant advance was made after the installation of new, fine spray, solid cone pattern nozzles. The test data is summarized below.

TEST NO.	30212	30218	30225	30305	30312
Duration - Hrs.	54	60	38	70	70
Method, Wash Distribution	Stationary Header Hollow Cone	Stationary Header Hollow Cone	Stationary Header Hollow Cone	Stationary Header Solid Cone	Stationary Header Solid Cone
Ave. Product Rate - GPM	8.06	7.61	8.0	8.1	8.7
GPD/sq.ft.	358	327	356	360	386
Ave. Salinity - PPM	489	268	300	200	150
Ave. % Net Wash	11.1	7.9	10.2	8.0	12.0
% Time Over 400 PPM	55	3.9	0	2.0	0
Highest 8 Hr. Salinity Controlled Product Rate GPM	7.25	8.26	8.33	8.8	9.71
GPD/sq.ft.	322	367	369	391	432

In test 30312, with nozzle plugging problems eliminated, column performance was at least as good as that obtained in any previous test with the original rotary scraper and wash distribution system. The 9.71 gpm (432 GPD/sq.ft.) product rate noted was actually held for the last sixteen hours of the test with a net wash of only 5.4% of the gross product.

4.6.3 Flooded Column

Several tests were undertaken to investigate operation of the separation column with greatly increased wash water quantities sufficient to flood the top of the column completely. Successful operation under these conditions would be particularly advantageous with a vapor compression direct freeze process where it is desirable to condense the vapor directly on the ice surface. Flooding of the top surface of the ice should result in essentially perfect wash distribution. The tests were promising in at least two respects.

The integrity of the ice piston was retained and, in general, operating parameters required little change. It was found also that an ice rate significantly greater than that possible in the so-called "dry" column tests, could be washed effectively. The primary difficulty involved was the control of <u>net</u> wash, which is, of course, an important economic factor.

Actual determination of the net wash was a problem also encountered in these tests. Since the top of the column was completely flooded, it was not possible to obtain representative ice samples, which were the basis of the standard calculations. New procedures were developed based on heat and mass balances around the melt water system and the freezer. The new methods offer the side benefits of eliminating the work of obtaining ice samples and of providing the data required in a form suitable for automatic computation of net wash. The calculated net wash could thus be used in control circuitry if this proves desirable. These methods are discussed in Appendix A. The test data is summarized below.

TEST NO.	21210	21217	30325	30401
Duration of Test, Hrs.	34	55	76	71
Duration of Attempted Flooding, Hrs.	18	30	76	60
Average Product Rate, GPM	9.27	10.68	7.20	7.26
GPD/sq.ft.	412	475	320	323
Average Applied Wash, GPM	11.1	17.3	11.45	13.3
Average Net Wash as % of Gross Product	13.9-17.9	10.6-14.6	28.1	28.5
Average Salinity, PPM	289	330	490	182

During the first two tests, the new methods of determining net wash had not been fully developed and a range is given to indicate possible errors. The device used for control of the net wash in tests #21210, #21217, and #30401 was the rate of ice production. As pointed our earlier in this report, the velocity of the ice rising in the column has a direct effect on the downward flow of wash water. Therefore the downward flow, or net wash, can be controlled by controlling the production rate and ice velocity. During test 30325, a brine back pressure control scheme was used in an attempt to decrease the net wash at lower ice production rates.

The equipment arrangement for column skirt back pressure control is shown in view 13A on drawing RD1039-4196. The brine level in the skirt was increased to partially fill the small surge drum and then the top vent valve was closed. By controlling the by-product brine flow from the column skirt the surge drum level and back pressure on the column were controlled. The higher skirt back pressures required due to the physical location of the surge drum were equivalent to an actual reduction in the washing height of the column and salinity control was lost at an excessive net wash rate. However, the net wash and product rates were stabilized to some extent with this control scheme and it could prove useful if incorporated to control back pressure in a range more nearly equal to that resulting from a normal brine level in the skirt. This was not practical at the pilot plant because of physical limitations in the column and brine skirt design.

To date neither method of net wash control has proved entirely satisfactory. Acceptable net wash values were obtained at high ice rates for brief periods, but salinity control was lost on several occasions and the average net wash was excessive. Flooded column operation has been proven technically feasible, but additional testing would be required to develop a suitable method of net wash control and to prove economic feasibility.

4.7 ABSORPTION MACHINE

The absorption equipment was modified in minor details prior to the period covered in this report. The absorption system in this process was subjected to two conditions completely foreign to commercial refrigeration duty, namely gradual sodium chloride contamination from the freezer and severe exposure to non-condensables from the recycle brine circuit. The purge system was successfully modified to handle large amounts of non-condensables, up to 0.062 lbs. per hour in some instances. Batch precipitation of contaminating salts had been conducted successfully. An unexpected feature of this fractional crystallization was that the precipitate analyzed consistently higher than 95% sodium bromide. Continued entrainment contamination by sodium chloride and consequent sodium bromide precipitation could convert an appreciable portion of the lithium bromide absorbent to lithium chloride, which fortunately is also an excellent absorbent. Tests of a scheme for continuous purification of the lithium bromide system were initiated under an earlier contract.

Entrainment contamination of the LiBr absorbent was light during this period, probably due to the continued use of overflowing standpipes in the freezer. However, evaluation of the system for continuous cooling and filtration of an absorbent sidestream was continued. The sidestream circuit is shown on the process piping diagram RD1039-9193. Concentrated LiBr from the generator-condenser was directed through a cooling coil, a filter, and back to the main stream entering the absorber. Cold brine from the column skirt was circulated around the cooling coil. Solution samples taken before and after the cooler-filter indicate that sodium salts were effectively removed. The small tubes in the cooler tended to plug over an extended period of operation. However, the basic principle appears to be practical, since the bromide concentration and brine temperature normally available is suitable for maintaining an acceptable absorbent purity.

The absorption machine performed effectively throughout the pilot plant program and was essentially trouble free. The system performed according to design and was not a limiting factor on plant capacity.

4.8 CONTROL INSTRUMENTATION

Continued emphasis was placed on simplification of control instrumentation and on establishing operating principles compatible with the application of automatic control systems. The majority of the pilot plant instruments were used for automatic control of various flow streams and are not considered excessive. Some simplification may be obtained in future designs by advantageous equipment arrangements. The primary exception to the above conclusion is in the area of separation column control.

4.8.1 Simplification of Column Controls

At the beginning of this period the operation of the separation column involved automatic slurry flow control to each cell, automatic control of recycle brine flow, automatic control of reject brine flow, solenoid controlled pulsed brine underflow, and manual control of wash water flow.

As described earlier in this report, lining the column with PVC sheet material reduced sidewall friction and eliminated the sticking problems observed with the painted steel column. The need for pulsed underflow was thereby eliminated. Considering the number of solenoid valves that might be required in a multi-cell production plant, and the expected service life and maintenance required for such valves and actuators, this represents a major simplification in control.

The progress made in increasing the potential size of individual separation cells will reduce the overall costs for automatic control of slurry flow to each cell. Also, much smoother ice rise through the column has resulted in a considerable degree of stabilization of column bottom pressure. Since the only function of the automatic slurry flow control valve is to maintain a constant flow rate, the possibility of using a manual flow control valve or fixed nozzle for controlling the flow to multiple cells of a production plant is suggested. Further work would be necessary to confirm this possibility.

Automatic control of reject and recycle brine flow is occasioned by the requirement of maintaining a predetermined level in the freezer and the brine surge tank (column skirt). It is anticipated that some form of level control will always be required and therefore, there is little chance of significant simplification in these areas.

4.8.2 Automation in Plant Control

Particularly in the smaller plant sizes, completely automatic operation is necessary if reasonable operating costs are to be attained. Experience with pilot plant operation during this period indicates that several major obstacles to complete automation of the vapor absorption process have been removed.

Salinity control has been obtained principally by manual control of the wash water flow. During periods of variable production rates, there appeared to be no rational criteria for determination of the proper amount of gross wash to be applied that would both minimize the net wash and maintain salinity control. At least two possible solutions to this problem have recently been developed.

A continuous measurement of the ice harvest rate has been achieved at the pilot plant by sensing the change in melt water temperature resulting from the melting of the ice product. Basically the net wash is determined by comparing the ice harvest rate to the melt product rate (Appendix A). It should be practical to automatically control the gross wash water flow to maintain the desired net wash based on these measurements.

During periods of essentially constant production, it has been found possible to minimize the net wash by varying the gross wash applied to maintain the salinity of the melt product close to the maximum allowable. The ability to operate close to the salinity limit is entirely due to improve stability in ice harvest rate which provides sufficient time to correct a salinity trend by increasing gross wash flow. Under such conditions then, it should be practical to control wash flow from a continuous conductivity signal of salinity.

Another obstacle to automatic operation was the more or less severe consequences of a loss of salinity control or some other system upset. However, it is now possible to completely stop and start production without melting the ice in the column or making other abnormal changes in operating conditions. Again, this is directly connected with the elimination of sticking in the separation column and improvements in drain plate design. There appears to be nothing inherent in the process now that would prevent designing for automatic shutdown on an alarm signal and automatic restarting when the signal is cleared.

5.0 ECONOMIC EVALUATION

The vapor absorption type of freezing process has been shown to be technically feasible by 2 1/2 years experience in the successful operation of the nominal 15,000 GPD pilot plant at Wrightsville Beach. Based on this experience a design and cost study was made for a production plant having a capacity of 165,000 GPD in order to evaluate the potential economics of this process.

5.1 Plant Design

Drawing R1039-9177 is a flow scheme of the process chosen for evaluation. The principle of operation is identical with that of the pilot plant but certain refinements and modifications have been made to reduce overall capital and operating costs.

A multi-stage flash type evaporator is used for precooling the sea water feed. The vapor condensate recovered from the precooler adds appreciably to the plant capacity. With 83°F sea water the addition amounts to about 10%.

The absorber is incorporated in the upper part of the freezer vessel to reduce capital cost. The auxiliary refrigeration is applied directly to cooling coils in the absorber rather than being applied to the melt water circulation. This reduces the total heat transfer surface required and also reduces the power required for auxiliary refrigeration.

The regeneration of the absorbent solution is done in three stages, equivalent to a triple effect evaporator, rather than a single stage, thereby reducing the steam usage required.

The freezer is a controlling component in establishing the size of the plant. A size of 10 ft. diameter by approximately 30 ft. long has been established as a practical size for factory assembly and transportation. Based on pilot plant experience, such a freezer would have a capacity of about 150,000 gallons per day.

This capacity may be supplemented somewhat by using a multi-stage flash type evaporator for precooling the sea water feed. When the sea water temperature is at 33°F the gain from this type cooler amounts to about 10% making the maximum capacity of such a plant 165,000 GPD. At lower temperatures the gain is much less and at an assumed minimum temperature of 40°F the gain is negligible and the plant capacity is only 150,000 GPD corresponding to the basic capacity of the freezer.

Larger size plants can be built by using multiples of the 165,000 GPD module or by using field fabricated parts for those components which are too large for factory fabrication or shipment in combination with factory fabricated parts. The largest single unit considered feasible with present knowledge is about 500,000 GPD.

A plant size of 165,000 GPD fits in well with what is believed to be the probable demand for conversion plants for the next several years. It is believed that in general the need for saline water conversion plants will be in the range from 150,000 GPD or even less, up to 1,000,000 GPD approximately. Such plants may be needed by small or expanding communities.

In line with these considerations, a size of 165,000 GPD (with flash type precooler) was selected for detailed study. A very complete design was prepared to permit a reliable cost estimate to be made. Drawing R1039-9171 shows a general arrangement of the plant and drawing R1039-9174 shows a prespective view. Figure 7 is a photograph of a scale model of the plant which was made to permit a better visualization of the general arrangement of the various items of equipment in the plant. It will be noted from the general arrangement drawing that the plant required a space of only 41'-3" by 67'-0" which certainly is very compact for a plant of this capacity.

5.2 Capital Cost

Based on the general design described above a detailed estimate of capital cost was made. Design layouts of the freezer, the separation column, and other special equipment items were made in sufficient detail to obtain reliable manufacturing cost estimates from Carrier Air Conditioning Company, a division of Carrier Corporation. Quotations were obtained from outside vendors covering the principal items of standard equipment. Erection and construction costs including foundations, piping, etc. were made based on the general arrangement drawing R1039-9171. Subcontract quotations were obtained covering the electrical wiring, structural steel work, insulation, etc.

The general philosophy followed in preparing these designs and estimates was to obtain a minimum realistic cost for a commercial quality plant. All unnecessary features and requirements were eliminated. Special code requirements were avoided except for normal building codes and regulations applying to ordinary commercial plants. Construction materials suitable

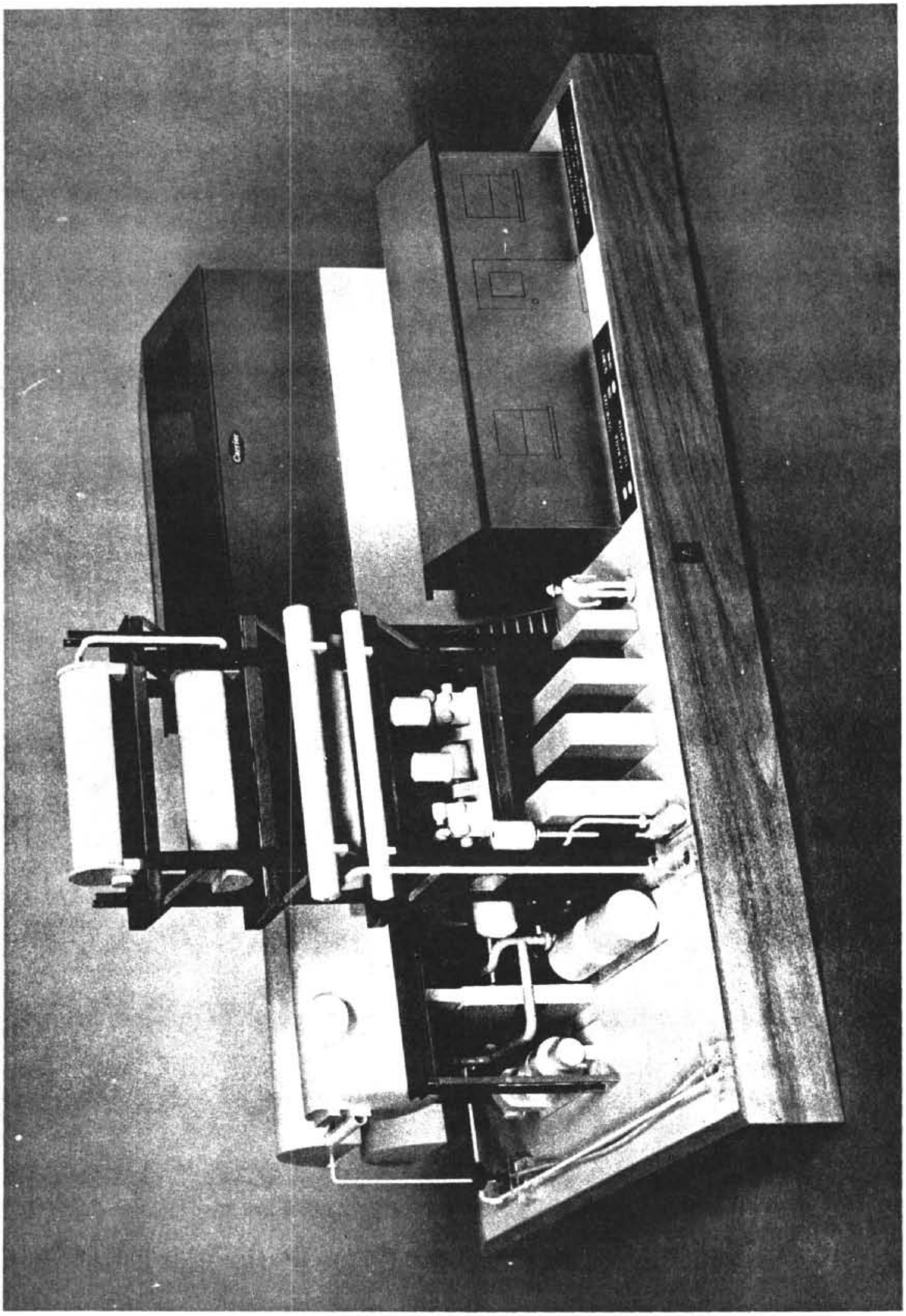


FIGURE 7 : SCALE MODEL OF 165,000 GPD SALINE WATER CONVERSION PLANT USING THE WATER VAPOR ABSORPTION TYPE FREEZING PROCESS

for the conditions in accordance with pilot plant experience were used but premium cost materials were avoided. Good industrial practice was followed throughout but an effort was made to keep all cost at a minimum consistent with good practice.

The following conditions were assumed as a basis of the cost estimate:

1. Plant is located near the sea coast in a mild climate similar to that at Wrightsville Beach, North Carolina.

2. Sea water intake can be taken from a protected bay or inlet at a point no more than 50 feet from shore with no dredging required.

3. Plant is located within 100 feet of the sea water intake.

4. Electric power is delivered to the plant site at 440 volt, 3 phase, 60 cycles. No substation or transformer has been included on the basis that such a facility would be furnished by the electric utility company.

5. No steam boiler is included in the plant cost. Steam is assumed to be delivered to the plant site from a separate facility.

6. No spare or standby equipment is included as a part of the plant cost.

7. No spare parts are included as a part of the plant cost.

8. A minimum building is provided as a control and operating room.

9. No maintenance shop facilities are included.

10. No provision is made for product water storage.

11. No provision is made for chlorination or other treatment of product water or sea water intake.

12. No access roads are included.

13. No landscaping or site improvement is included.

14. No fencing is included.

On the above basis the price of a 165,000 GPD production plant is estimated to be $557,000 including the contractors normal profit. Following is a breakdown of this cost.

ESTIMATED CAPITAL COST OF 165,000 GPD PLANT

Special Equipment	$120,970
Standard Equipment	103,300
Freight	5,050
Assembly and Erection	156,710
Instruments and Control	29,220
Buildings	4,130
Engineering and Supervision	61,700
Start Up and Test	8,520
Contingency	50,000
Sub-Total	544,600
Sea Water Intake Facility	12,400
Total Plant Investment	$557,000
Plant Investment per GPD Basis 165,000 GPD	$3.38

For comparison the cost was estimated by the OSW standard procedure using the same costs for special and standard equipment as used in the Carrier estimate. This estimate is as follows:

1.	Special Equipment - FOB	$ 96,700
	Installation - 30%	29,010
	Special Equipment - Installed	24,270
2.	Standard Equipment - FOB	100,100
	Installation - 30%	30,030
	Standard Equipment - Installed	8,200
	Total Principal Items of Equipment (PIE)	288,310
3.	Erection and Assembly - 30% of PIE	86,493
4.	Instruments - 4% of PIE	11,532
	Total Essential Plant Costs	$386,335
5.	Raw Water Supply at $5 per 1000 GPD (1,575 GPD = 2,270,000 GPD)	11,350
6.	Product Water Storage at $10 per 1000 GPD of Product	1,650
7.	Facilities and Buildings - 10% of PIE	28,830
8.	Contingencies - 10% of above 7 items	42,817
9.	Engineering - 10% of above 8 items	47,098
10.	Interest on Investment During Construction, 4% of above 9 items	20,725
11.	Site - $3 per 1000 GPD Product	495
	Total Plant Investment sum of above 11 items	$539,300
	Plant Investment Per GPD Basis 165,000 GPD	$3.27/GPD

A comparison of the two estimates indicates that the OSW procedure gives somewhat lower costs but is not grossly different in this case.

5.3 Operating Cost

The operating cost was calculated in accordance with the OSW standard procedure. For comparison, costs were calculated based on the actual estimated plant investment and on the investment calculated by the OSW procedure. These estimates are summarized below:

	Investment Calculated by OSW Procedure		Estimated Actual Investment	
Plant Capacity, GPD	165,000		165,000	
Total Plant Investment	$539,300		$557,000	
Plant Investment, $/GPD	$3.27		$3.38	
Electric Energy, Kwhr/1000 gal.	46.1		46.1	
Electric Rate, $/Kwhr	0.007		0.007	
Steam Usage - lbs/1000 gal.	557		557	
Steam Cost, $/1000 lbs	0.55		0.55	
	$/day	$/1000 gals	$/day	$/1000 gals
Electricity	53.30	.323	53.30	.323
Steam	50.55	.307	50.55	.307
Chemicals	2.25	.014	2.25	.014
Supplies	8.15	.050	8.45	.051
Operating Labor	23.50	.142	23.95	.145
Maintenance Labor	8.15	.050	8.45	.051
Payroll Extras	4.75	.029	4.85	.029
Overhead	10.90	.066	11.20	.068
Amortization	120.95	.733	124.90	.757
Insurance and Taxes	32.70	.198	33.75	.205
Interest on Working Capital	2.30	.014	2.35	.014
Total Owning and Operating Cost	317.50	1.926	324.00	1.964

The operating cost has been estimated in strict accordance with the OSW standardized estimating procedure. It should be pointed out that this procedure does not provide adequate operating labor for a plant of this size. The amount provided by the procedure is 10% of the electricity, steam, chemicals, supplies, and amortization which in this case is $0.142 or $0.145 per 1000 gallons depending on the initial plant investment used. Actually to provide one operator per shift at $2.50 per hour would amount to $0.364 per 1000 gallons. If allowance is made for a 10% wage premium on the second and third shift and for holidays, vacation, and sick leave the cost becomes $0.428 per 1000 gallons. This is considered a more realistic estimate of operating labor for a plant of this size.

The operating labor for a plant of 500,000 GPD or more should be more nearly in agreement with that provided by the OSW procedure.

It should be noted also that supplies and maintenance labor each are taken as 1/2% of the plant investment per year in accordance with the OSW procedure. Experience in the chemical and petroleum industry indicates that these two items together amount to from 2% to 10% with an average of about 4%. It is believed that the maintenance costs for the freezing process should be less than for other conversion processes because of the low operating temperatures. Corrosion and scaling should be minimized with this process. It is felt, however, in view of the experience of the chemical and petroleum industries, that the actual maintenance costs may be somewhat higher than provided in the OSW procedure.

Amortization has been taken over 20 years in accordance with the standard procedure.

5.4 Possibilities for Improvement

The evaluation made thus far has been based on the lowest cost reliable design it has been possible to prepare on the basis of present knowledge. Nevertheless it is possible that further reductions could be made with more experience in the construction and operation of a plant of this size.

A careful review of the equipment and construction costs indicates that with the elimination of all contingencies, and assuming some additional future improvements, the first cost might be reduced to about \$470,000 or \$2.85 per GPD. This is about a 15% reduction.

A review of the operating costs indicates that the major factors are energy costs for electricity and steam, operating labor, and amortization.

Amortization, of course, is directly related to the capital investment and so, in line with the above, it is unlikely to be reduced more than 15%. Of course, if a longer amortization period or lower interest rate could be justified, this would reduce this cost somewhat. A 30-year amortization instead of 20-year would reduce this item about 20%.

Operating labor may be reduced by increased automation or by operating a larger plant with the same labor. Also if the saline water conversion plant could be operated in conjunction with a power plant or other process plant such that only part of the total operating labor need be allocated to water conversion, then this could reduce the labor cost.

It may be possible ultimately to design small plants to operate completely automatically. This does not appear feasible in the foreseeable future, however, for a plant of this size. It is believed that a single operator could handle a larger plant or multiple plants of this size totalling up to about 600,000 GPD. This would reduce actual operating labor to about $0.12 per 1000 gallons assuming a base labor rate of $2.50 per hour and would be approximately in line with the OSW procedure.

Electric energy is estimated at 317.1 Kw for the 165,000 GPD plant amounting to 46.1 Kwhr per 1000 gallons. A review of this indicates the possibility that this may be reduced about 10% in the future to about 41.5 Kwhr per 1000 gallons.

Steam requirements for the process are 3,830 lbs per hour for 165,000 GPD or 557 lbs of steam per 1000 gallons (61.2 Btu per pound of water or about 15 lbs of water per pound of steam). No possibility of any appreciable reduction in this quantity can be foreseen.

A study was made of the possibility of using a steam turbine to furnish part of the power required for the auxiliary refrigeration and using the exhaust steam from the turbine to operate the absorption process.

Such an arrangement would reduce the total energy cost somewhat but would have a higher first cost for the turbine, and therefore, higher amortization and maintenance costs. Also a high pressure steam supply would be needed which would mean higher steam cost or boiler cost. A supply pressure of 250 psig was assumed for this study.

When these factors are considered, there is no appreciable saving by the use of a turbine in a plant of this size.

Another possibility for reducing steam cost occurs if the conversion plant can be located adjacent to a power plant so that low pressure bleed steam can be obtained from the power plant at low cost. The conversion plant with 3-stage absorption generator requires steam at the generator at 65 lbs/sq.in. absolute. To provide for line losses and controls the bleed pressure at the power plant turbine probably would have to be at least 75 psia. The cost estimates presented in this report have been based on steam costs of $0.55/1000 lbs. in accordance with the OSW procedure. If bleed steam at the necessary pressure can be obtained at lower costs, then there would be a corresponding saving in energy cost.

The 3-stage absorption process requires rather high steam pressure as noted above. It is perfectly possible to design a two-stage or single-stage process which would operate with much lower pressures but would require greater quantities of steam. The following table shows the steam pressure and quantity required and also the steam cost for equal energy costs for three-stage, two-stage, and single-stage absorption.

	3 Stage	2 Stage	1 Stage
Steam Pressure at Absorption Generator - psia	65.0	20.0	5.3
Steam Pressure at Turbine Bleed Port - est. - psia	75.0	30.0	15.3
Heat Required - Btu/1000 gals.	510,000	709,500	1,345,000
Steam Required - lbs/1000 gals.	557	733	1320
Equivalent Steam Cost - $/1000 lbs.	.55	.418	.232

The above table indicates, for example, that a single stage absorption process would require 1320 lbs of steam per 1000 gallons of product. The steam pressure required would be 5.3 psia at the absorption generator and 15.3 psia estimated at the turbine bleed port. This steam would have to be available at $0.232 per 1000 lbs to compete with a 3-stage process using steam at $0.55 per 1000 lbs.

It might be expected that a two stage or single stage process would be significantly lower in first cost than a three stage process. Although no detail cost study has been made it is estimated that actually there would be very little if any reduction. The total amount of water vapor to be evaporated remains the same and so the total heat transfer surface required remains about the same. Therefore, with fewer stages the individual stages must be larger. Also with fewer stages the amount of vapor to be condensed in the final condenser is increased and so the condenser becomes appreciably larger. This also would require increased sea water flow for condensing. Therefore it is estimated that the plant cost would not be affected appreciably by the number of stages in the absorption process.

Considering the possible future reductions discussed above, the following table shows the predicted operating costs that may be attainable in the future with this process.

Plant Capacity - GPD	165,000
Total Plant Investment	$470,000
Plant Investment - $/GPD	$2.85
Electric Energy - Kwhr/1000 gals.	41.5
Electric Rate - $/Kwhr	0.007
Steam Usage - lbs/1000 gals.	557
Steam Cost - $/1000 lbs.	0.55

	ESTIMATED FUTURE OPERATING COSTS-- $ per 1000 gallons
Electricity	.290
Steam	.307
Chemicals	.014
Supplies	.043
Operating Labor	.129
Maintenance Labor	.043
Payroll Extras	.026
Overhead	.060
Amortization	.639
Insurance and Taxes	.173
Interest on Working Capital	.013
TOTAL OWNING AND OPERATING COST $ per 1000 gallons	1.737

6.0 CHEMICAL AND BACTERIOLOGICAL TESTS

At the request of the Office of Saline Water, chemical and bacteriological tests were conducted at Wrightsville Beach by the Department of Health, Education, and Welfare; Public Health Service; Division of Water Supply and Pollution Control; Robert A. Taft Sanitary Engineering Center; Cincinnati, Ohio during June 1962. The results of these tests were reported directly to OSW by the Public Health Service in their report entitled "Chemical and Bacteriological Tests on Intake Sea Water and Product Water, Carrier Corporation Freezing Process Pilot Plant, Wrightsville Beach, North Carolina" and dated July 1962.

The following is summarized from that report. Statements in quotation marks are quoted verbatim from that report.

6.1 Chemical Tests

Two samples of the sea water feed and of the product water were collected for chemical analyses. Sample No. 1 was collected by adding 400 milliliters of each stream to polyethylene bottles every six hours from 10:00 a.m., June 15 until 4:00 a.m., June 17. Sample No. 2 was collected in a similar manner from 10:00 a.m., June 17 until 10:00 p.m., June 18. The sea water feed samples were taken after the deaerator. Product water samples were taken after the mixing of melted product ice and condensed water vapor from the lithium bromide water vapor absorption system.

"The chemicals specified in the Public Health Service Drinking Water Standards (1961) were determined. These chemicals, the concentrations listed in the Standards, and the concentrations found in the samples are shown in

Table 1. Not tested for were phenol, chromium, and cyanide. These materials would be unlikely to be in feed waters unless industrial pollution is present."

Several other chemical analyses were made as a matter of interest. The materials determined and the results are shown in Table 2.

"The water produced met the Public Health Service Drinking Water Standards in all respects except iron content which was 0.4 ppm and 0.64 ppm (standard 0.3 ppm) on the composite samples tested......Such a result would not be unexpected in a pilot plant operation where fresh metal surfaces are exposed to various forces.....Except for the slight excess of iron, the chemical quality of the product water was quite good. Considering the chemicals in the feed water, the freezing process has done a good job of producing potable water from the chemical standpoint."(1)

(1)Separate tests by Carrier showed that the excessive iron content was largely in suspended rather than dissolved form and therefore could be removed by settling or filtering. Samples which were allowed to set for 24 hours showed a very slight accumulation of sediment in the bottom of the sample bottle. The iron content of the decanted water was found to be less than the PHS standard of 0.3 ppm as compared with 0.6 ppm in the sample before settling.

A filter was installed in the product water line and it was found that a 5 micron filter reduced the iron content of the water from about 0.5 ppm to less than 0.2 ppm. A 75 micron filter was not as effective in removing the iron although the concentration was reduced to approximately 0.3 ppm.

It is concluded that the high iron content is not inherent in the process but represents pick up of rust and scale from rusty tanks and pipes in the pilot plant. This should not be a problem in a production plant. If it should be a problem, then these tests have shown that the iron can be reduced to a satisfactory level by suitable filtering.

TABLE 1. TESTS FOR COMPLIANCE WITH PUBLIC HEALTH SERVICE DRINKING WATER STANDARDS

CARRIER FREEZING PROCESS, WRIGHTSVILLE BEACH, NORTH CAROLINA (JUNE 1962)

COMPONENT	PHS DRINKING WATER STANDARD (1961)		FEED WATER		PRODUCT WATER		LITHIUM BROMIDE CONDENSATE[A]
	RECOMMENDED MAXIMUM	GROUNDS FOR REJECTION	NO. 1	NO. 2	NO. 1	NO. 2	
	mg/1	mg/1	mg/1	mg/1	mg/1	mg/1	mg/1
ALKYL BENZENE SULFONATE (DETERGENT)	0.5	--	--	--	<0.1	<0.1	0.00
ARSENIC	0.01	0.05	0.01	<0.01	<0.01	<0.01	<.01
BARIUM	--	1.0	1	1	0.005	0.005	N.D.
CADMIUM	--	0.01	N.D.	N.D.	N.D.	N.D.	N.D.
CARBON CHLOROFORM EXTRACT	0.2	--	.056	--	.074	--	--
CHLORIDE	250	--	18,425	19,112	173	127	--
CHROMIUM[B]	--	0.05	--	--	--	--	--
COPPER	1.0	--	0.09	0.07	0.1	0.9	--
CYANIDE[B]	0.01	0.2	--	--	--	--	--
FLOURIDE	1.7	2.2	1.9	2.1	0.08	0.08	.00
IRON	0.3	--	0.44	0.28	0.40	0.64	0.09
LEAD	--	0.05	N.D.	N.D.	N.D.	N.D.	N.D.
MANGANESE	0.05	--	N.D.	N.D.	N.D.	N.D.	0.0
NITRATE (NO_3)	45	--	N.D.	N.D.	N.D.	N.D.	--
PHENOLS[B]	.001	--	--	--	--	--	
SELENIUM	--	0.01	<0.01	<0.01	N.D.	N.D.	<0.01
SILVER	--	0.05	0.2	0.02	N.D.	N.D.	--
SULFATE	250	--	2,400	2,300	25	19	<1.0
TOTAL SOLIDS	500	--	35,560	36,230	346	245	--
ZINC	5		None	None	0.10	0.06	--

[A] THIS STREAM WAS SAMPLED TO DETERMINE CARRY-OVER
[B] NOT EXPECTED IN SAMPLES EXAMINED -- NO ANALYSIS

N.D. NOT DETECTABLE

TABLE 2. CHEMICAL TESTS OTHER THAN THOSE CALLED FOR IN PHS STANDARDS

CARRIER FREEZING PROCESS, WRIGHTSVILLE BEACH, NORTH CAROLINA (JUNE 1962)

Component	Feed Water No. 1 mg/l	Feed Water No. 2 mg/l	Product Water No. 1 mg/l	Product Water No. 2 mg/l	Lithium Bromide Condensate mg/l
Boron	1.9	1.6	0.05	0.02	0.00
Color	5	5	<5	<5	<5
Sodium	9,000	9,000	98	72	<1.0
Lithium	Trace	Trace	.03	.03	0.1
Potassium	360	360	4	3	<1.0
Total Phosphate (PO_4)	0.03	0.03	0.04	0.03	0.04
Total Alkalinity ($CaCO_3$)	102	108	5	4	4
Total Hardness ($CaCO_3$)	5,950	6,200	100	44	8
Dissolved Solids[a]	35,510	36,180	346	243	
Suspended Solids	46	50	0	3	
Volatile Solids	4,170	4,060	46	30	
Conductance ($Ohms^{-1}$)	$4.5x10^{-2}$	$4.3x10^{-2}$	$6.1x10^{-4}$	$4.5x10^{-4}$	
pH	7.9	7.8	6.6	6.4	
Chemical Oxygen Demand	--	--	N.D.	N.D.	
Odor	4[b]		1[b]		

a - By Difference b - Threshold Odors N.D. - Not Detectable

6.2 Odor Tests

"Threshold odor tests were run on the raw deaerated water and the product water on three days. The test is run by diluting the sample with odor-free water to the point where odor detection just occurs. Four or five people were used to judge each sample. The threshold number is the number of dilutions made to the point of detection. For example, a threshold of four means the original sample constitutes one-fourth of the diluted sample. The following results were obtained."

	Number of Times Observed	
Threshold Number	Feed Water	Product Water
1	---	7
2	2	3
4	5	3
8	5	---

"Expressing the total panel results as the median, the raw water had a threshold of four, the product a threshold of one."

"The odor quality is judged to be satisfactory. In practice, the consumers are the final judge of odor quality and while indications are that the odor is satisfactory, final judgement would await actual use of the water."

"The deaerated sea water also had relatively low odor (threshold four).....The production of good quality water from the odor standpoint in these tests would not assure that good quality water could be produced from a highly odorous source."

6.3 Bacteriological Tests

Beginning on June 15, 1962 and continuing for four consecutive days, samples were collected at 4:00 a.m., 10:00 a.m., and 4:00 p.m. from:

1. The sea water feed after the deaerator.
2. The finished ice product prior to melting and mixing with the condensate return from the lithium bromide unit.
3. The waste brine.

Each sample was tested for:

1. Total coliform enumeration.
2. Fecal coliform enumeration.
3. Fecal streptococcus enumeration.
4. Tests on one sample from each of the three sources once a day for total bacterial enumeration.
5. Tests to confirm fecal coliform colonies from one saline intake sample.

Bacterial densities were determined as count per 100 milliliters of sample. The results are summarized in the following table which is taken from Table 6 of the PHS report.

GEOMETRIC MEANS OF BACTERIAL COUNTS

	Sea Water (Raw)	Finished Water (Ice)	Brine Waste
Total Coliforms	30/100 ml.	2/100 ml.	22/100 ml.
Fecal Coliforms	16/100 ml.	2/100 ml.	13/100 ml.
Non-fecal Coliforms	14/100 ml.	2/100 ml.	9/100 ml.
Streptococci	20/100 ml.	1/100 ml.	17/100 ml.
Colony Count (MF)	14,000/100 ml.	1200/100 ml.	6700/100 ml.

"The coliform group has been the classical measure of the sanitary quality of water. The coliform group density in the raw sea water at the Wrightsville intake averaged 30 per 100 milliliters representing a relatively good quality sea water; however, sea water in the open ocean would probably be free from the coliform group. There are no bacterial standards for sea water to be used for processing to a potable fresh water.....The maximum coliform density for overlying water for approved shellfish beds is 70 per 100 milliliters."

"The coliform group may be divided in those organisms (fecal coliforms) derived from feces of warm blooded animals and a group (non-fecal) from unknown sources." Likewise the fecal streptococcal test is another index of micro-organisms derived from fecal (warm-blooded) pollution. "Both the fecal coliforms and the streptococci are a measure of fecal pollution in a stream, sea water, or wherever they may be found."

"The fecal coliform density in the untreated sea water averaged 16 per 100 milliliters and the fecal streptococcal density was 20 per 100 milliliters. These numbers represent the actual organisms present as fecal pollution and their essential removal, plus the elimination of the non-fecal coliform group, represents a good product from the bacteriological quality viewpoint.....The finished water on an average had no coliforms (less than 2 per 100) and one streptococcus per one hundred in the samples examined..... The Public Health Service Standards for potable waters using this technique and sampling frequency states that coliforms should not exceed four per hundred milliliters. The water thus meets potable Standards."

"The bacterial quality of the sea water being fed to the process at the time of these tests was good. Ability to meet the Standards with this feed does not assure that Standards can be met when a more polluted feed water is used.....It would be essential, of course, to chlorinate the finished water if used as a drinking water supply, which is in accordance with good public health practice for water supplies."

7.0 CONCLUSIONS

The following conclusions have been reached as a result of the work performed under this and the preceding contracts.

The feasibility of producing fresh water from sea water by the direct freezing process has been demonstrated. Sustained production of potable water has been demonstrated using less than 5% of total product for wash separation.

The chemical quality of water produced met the Public Health Service Drinking Water standards in all respects except iron content which was 0.4 ppm to 0.64 ppm compared with the standard requirement of 0.3 ppm. It was shown that the iron content could be reduced to less than 0.2 ppm by simple filtering of the final product.

The bacteriological quality of the product water also met the Public Health Service standards for potable water although this does not necessarily assure that the Standards could be met if a more polluted feed water were used.

Design requirements for a production plant of this type have been developed based on the improved component and process efficiency demonstrated during the pilot plant program. These cover the design of the freezer, the wash separation column, the scraper, the ice-brine slurry handling system, the process control system, and the necessary auxiliaries.

A practical size for the vapor absorption type process using prefabricated factory assembled components, and considering transportation limitations, is about 165,000 GPD (150,000 GPD from freezing and 15,000 GPD from flash evaporative precooling of feed).

Larger capacity plants can be built using multiples of the 165,000 GPD unit or by combining certain factory fabricated components with field fabrication or shipment. The largest single unit considered feasible with present knowledge is about 500,000 GPD.

The energy requirements of the vapor absorption process are estimated at 46.1 Kwhrs and 557 lbs of steam (at 65 lbs per sq.in. absolute) per 1000 gallons of product. It is predicted that the electrical energy may be reduces in the future but no appreciable reduction in steam requirements can be foreseen.

The estimated capital cost and operating cost for a 165,000 GPD plant based on the vapor absorption process are summarized as follows:

	Plant Capacity GPD	Total Plant Investment	Plant Investment $/GPD	Operating* Cost $/1000 gal
Estimated Acutal Cost Based on Present Knowledge	165,000	$557,000	$3.38	$1.96
Cost Computed by OSW Formula	165,000	$539,300	$3.27	$1.93
Predicted Future Cost	165,000	$470,000	$2.85	$1.74

*Operating costs are estimated according to OSW standard procedure in all cases.

The major items making up the total operating cost are energy costs, operating labor, and amortization of plant investment. All three of these elements must be reduced to very low levels if a total operating cost approaching \$1.00 per 1000 gallons is to be achieved. The fixed costs alone for amortization, and insurance and taxes amount to about \$.38 per 1000 gallons for each \$1.00 of plant investment per GPD of capacity.

Where low pressure steam is available at low cost, a two stage or single stage abosrption process can be used but will require greater quantities of steam. Relative economics will depend on the relative cost of the available steam at the required pressures.

8.0 RECOMMENDATIONS

1. The estimated costs for the vapor absorption process, based only on pilot plant experience, are sufficiently close to the publicly reported costs for other more fully developed processes that the vapor absorption process should be given continued consideration. It is recommended that a more direct cost comparison be made with comparable size plants of freezing and other known types using the same energy rates, labor rates, etc. The necessary detail information from actual operating pilot plants pertaining to other processes is not available to Carrier and, therefore, it is recommended that this comparison be made by the Office of Saline Water.

2. The vapor absorption process is believed to be the only freezing process that has demonstrated satisfactory operation for sustained periods of time in a pilot plant and produced potable water meeting the standards of the Public Health Service. Since other freezing processes may not reach this stage of development until some time in the future, it is recommended that a large capacity improved design plant as described herein should be constructed to carry forward the program of the Office of Saline Water for the following reasons:

a. To verify extrapolation of the design from the pilot plant size.

b. To verify the estimates of capital and operating costs.

c. To prove the performance of the design improvements which are proposed to reduce costs.

d. To develop and test improved instrumentation and automation of the process to minimize operating labor.

e. To obtain information needed to develop further design improvements and to extend the design to larger sizes.

f. To develop factual data on costs to enable a reliable comparison to be made with other processes which have already progressed through such a development program with a fully developed technical background.

APPENDIX A

A-1.0 DETERMINATION OF NET WASH

The net amount of product used or lost in washing the ice is an important factor in the efficiency of the freezing process for saline water conversion. The determination of the net wash is not always a simple matter, however. It is not simply the amount of wash water applied to the top of the column since a large amount of this water adheres to the ice and is returned to product with the ice.

Several methods for determining the net wash have been developed. These are described and the necessary equations are derived in the following sections.

A-1.1 Ice Sample Method

By taking a sample of the wet ice harvested from the top of the colum and melting it in a calorimeter it is possible to determine the percentage of ice and water in the wet ice. With this information and other data normally taken it then is possible to calculate the net wash.

Figure A-1 shows schematically the systems involved and the nomenclature used. The necessary formulas are derived as follows:

Mass Balance: System X

$$A + I = I' + D + N \tag{1}$$

$$I' + D = A + I - N \tag{1a}$$

Mass Balance: System Y

$$I' + D + V = P + A \tag{2}$$

$$I' + D = P + A - V \tag{2a}$$

$$I' + V - P = A - D \tag{2b}$$

Combining (1a) and (2a)

$$A + I - N = P + A - V$$

$$N = I + V - P \qquad (3)$$

Equation (3) is the basic formula for net wash and will be found to apply to all methods. V and P are measured directly but it is necessary to calculate I in order to determine N.

By definition

$$r = \frac{I'}{I' + D} \text{ or } I' = r(I' + D) \qquad (4)$$

Combining with (2a)

$$I' = r(P + A - V) \qquad (5)$$

Also

$$D = I' + D - I' \qquad (6)$$

Combining with (2a) and (5)

$$D = P + A - V - r(P + A - V)$$

$$D = (1-r)(P + A - V) \qquad (7)$$

<u>Heat Balance: System X</u>

$$I(-143.4) + H + A(Ta-32) = N(32-32) + D(32-32) + I'(-143.4) \qquad (8)$$

$$143.4\ I = H + A\ (Ta-32) + 143.4\ I'$$

$$I = I' + \frac{H + A\ (Ta-32)}{143.4} \qquad (9)$$

Referring to (3)

$$N = I + V - P$$

Combining with (9)

$$N = I' + V - P + \frac{H + A\ (Ta-32)}{143.4} \qquad (10)$$

Substituting from (2b)

$$N = A - D + \frac{H + A\ (Ta-32)}{143.4} \qquad (11)$$

Substituting again from (7)

$$N = A - (1 - r)(P + A - V) + \frac{H + A\ (Ta-32)}{143.4} \tag{12}$$

Per cent net wash with respect to gross product is

$$\%N = \frac{N \times 100}{P + N} \tag{13}$$

Equation (12) is the final relation for determining the net wash using the ice sampling procedure. It is dependent on obtaining a truly representative sample of the ice leaving the top of the column in order to determine the ratio, r, of the ice to the total of ice and adhering water leaving the column. Also it requires a determination of the heat gains, H, at the top of the column due to radiation and convection, condensation from the atmosphere, and heat input to the ice from the scraper.

It should be noted that to be consistent with the GPM units used throughout that the heat gain term, H, must be expressed in terms of gallon degrees per minute or Btu/min. divided by 8.34. The heat gains applying to this method are shown on Figure A-5.

Equations (12) and (13) above, or equivalent, were used throughout the pilot plant test program up until December 1962. After that date alternate methods were used as described in the following.

A-1.2 Heat Balance Method - Method C

The basic formula for net wash is N = I + V - P per equation (3) developed for the ice sample method. The terms V and P normally are available and it is only necessary to determine I. This can be done by taking a heat balance around the top of the column and the melt tank.

Figure A-2 shows schematically the system involved and the nomenclature used. The necessary formulas are derived as follows:

Mass Balance

$$I + M = N + M + P - V \quad (1)$$

$$N = I + V - P \quad (2)$$

This is the same basic equation as derived for the ice sample method at equation (3).

Per cent net wash with respect to gross project is

$$\%N = \frac{N \times 100}{I + V} \quad (3)$$

The quantity, I, may be determined from a heat balance as follows: Note that the boundary of the system is taken across the separation column below the top of the ice but near the top where both ice and associated water are in equilibrium at 32°F.

$$I(-143.4)+M(T_7-32)+H = N(32-32)+M(T_5-32)+(P-V)(T_5-32) \quad (4)$$

$$H - 143.4\ I + M(T_7-32-T_5+32) = (P-V)(T_5-32) \quad (5)$$

$$I = \frac{H + M(T_7-T_5) - (P-V)(T_5-32)}{143.4} \quad (6)$$

The heat gains indicated by H must include all the gains to the heat balance system indicated. This includes convection and radiation, and condensation from the atmosphere (if any) to the surface of the ice and to the water in the melt tank and also to all vessels and piping within the boundaries of the system. This also includes the heat equivalent of power input to the melt water circulating pump, P-5, and heat input to the ice from the scraper.

It should be noted again that for consistency with other units the heat gains must be expressed in gallon degree per minute or Btu/min $\div$ 8.34. The heat gains applying to this method as calculated for the pilot plant are shown by curve C on Figure A-5 as a function of the ambient temperature.

At the pilot plant the temperature difference, T_7-T_5 was measured directly by a thermo-pile containing 2 thermocouples in each element.

A-1.3 Heat Balance Method; Method C-1

Figure A-3 shows an alternate heat balance method for determining the net wash. Again the system boundary is taken across the top of the separating column just below the top of the ice where the ice and water are in equilibrium at 32°F.

The formulas applying to this method are derived as follows:

Mass Balance

$$I + M + A = N + M + A + P - V \qquad (1)$$

$$N = I + V - P \qquad (2)$$

This again is the same basic equation as derived for the other methods.

Per cent net wash with respect to gross product is:

$$\%N = \frac{N \times 100}{I + V} \tag{3}$$

Heat Balance

$$I(-143.4)+M(T_7-32)+A(T_5-32)+H = N(32-32)+(M+A+P-V)(T_x-32) \tag{4}$$

$$143.4\ I = H+M(T_7-32)+A(T_5-32)-M(T_x-32)-A(T_x-32)-(P-V)(T_x-32) \tag{5}$$

$$143.4\ I = H + M\ (T_7-T_x) + A(T_5-T_x) - (P-V)(T_x-32) \tag{6}$$

$$(T_x-32) = (T_5-32) - (T_5-T_x) \tag{7}$$

Substitution in (6)

$$143.4\ I = H+M(T_7-T_x)+A(T_5-T_x) - (P-V)(T_5-32) + (P-V)(T_5-T_x) \tag{8}$$

$$I = \frac{H + M(T_7-T_x) - (P-V)(T_5-32) + (A+P-V)(T_5-T_x)}{143.4} \tag{9}$$

This is the basic equation for method C-1. It will be noted that this is similar to the final equation (6) for method C except for the final term in the numerator. The heat gains, H, applying to this method are also shown on Figure A-5 by curve C-1. These gains are much less than for method C mainly because the melt water circulating pump is not included in the system boundaries.

At the pilot plant the term (T_7-T_x) was measured directly by a thermo-pile using five thermocouples in each element. The temperature, T_x, was not measured separately and therefore the term T_5-T_x was not available from direct readings. For ordinary operating conditions, however, the term $(A + P - V)(T_5-T_x)$ was found to have a value of about 4. This is very small in comparison with the total numerator which normally amounted to 1000 or more. Therefore the error due to using an average value for the term $(A+P-V)(T_5-T_x)$ is negligible and equation (9) may be simplified to the following:

$$I = \frac{H + M\ (T_7-T_x) - (P-V)(T_5-32) + 4}{143.4} \tag{10}$$

This is the working equation which was used for this method.

A-1.4 Vapor Condensate Method: Method D

Another method of determining the ice production rate, and hence the net wash, is by means of a heat balance around the freezer. Figure A-4 is a schematic diagram of this method showing the boundaries of the heat balance system and the nomenclature used. Again the boundary has been taken near the top of the separation column at a plane where the ice and wash water are in equilibrium at 32°F.

As in previous methods the basic equation for net wash is

$$N = I + V - P \tag{1}$$

Per cent net wash is

$$\%N = \frac{N \times 100}{I + V} \tag{2}$$

The equation for determining the ice rate, I, is derived as follows:

Heat Balance

$$FG_{PF}(T_F-32)+N(32-32)+H = I(-143.4)+ BG_{PB}(T_B-32)+Vh \tag{3}$$

$$143.4\ I = VH - BG_{PB}\ (32-T_B) - FG_{PF}\ (T_F-32) - H \tag{4}$$

This is the basic equation for method D. The heat gains, H, include the radiation and convection and atmospheric condensation (if any) on the freezer, the separation column enclosure, and all the piping within the system boundaries. The heat input from the pumps, P-3 and P-4, also must be included. Heat gains to the top of the ice are not included, however, since this surface is outside the boundary of the system being considered. The heat gains for this method are shown on Figure A-5 by curve D.

At the concentrations normally encountered at the pilot plant C_{PB} = .9624 and C_{PF} = .975. A value of .97 may be used satisfactorily for both of these coefficients. Also normally B will equal 1/2 F, T_B will be about 27° and h will be about 1072. Therefore, for the normal operating conditions equation (4) may be simplified by these approximations without affecting the accuracy appreciably.

$$143.4\ I = 1072\ V - .97\frac{F}{2}\ (32-27) - .97F(T_F-32) - H \qquad (5)$$

$$143.4\ I = 1072\ V - .97F(T_F-32+\frac{5}{2}) - H \qquad (6)$$

$$I = \frac{1072V - .97F(T_F-29.5) - H}{143.4} \qquad (7)$$

This is the working equation which was used for this method. The approximations used should not cause an error of more than 1% for the operating conditions at the pilot plant. Errors in the determination of the heat gain, H, would increase the possible overall error slightly.

A-1.5 Discussion of Methods

None of the methods described can be considered completely accurate since each involves one or more terms that are difficult to evaluate accurately.

The ice sample method requires a truly representative sample of the ice from the top of the column to obtain the percentage of ice and water entering the melt tank. There is no way of knowing whether such samples are truly representative or mot. This method becomes virtually impossible when the amount of gross wash water is high so that the top of the column is actually flooded or nearly so.

This method also requires determination of heat gains to the top of the ice which is difficult to do with accuracy. The effect of atmospheric condensation is especially difficult to determine and would require observation of the relative humidity as well as the ambient temperature. The relative humidity was not available except in certain isolated cases and therefore this was estimated on the basis of a constant dew point depression of $8^{o}F$ below the ambient temperature. This corresponds to a relative humidity of 75 to 80% which is in good agreement with the average conditions at the pilot plant location. The heat gains in this method are relatively small, however, in comparison with methods C and C-1.

Method C requires an accurate measurement of the small temperature difference, $T_7 - T_5$. This difference normally is about 3^o to 4^o so an error of only 0.1^o amounts to an error 2-1/2% to 3% in the calculated ice rate and a corresponding error in net wash. At the pilot plant this difference was measured by a thermo-pile using two thermocouples in each element but even with this arrangement the accuracy of this measurement is somewhat questionable.

In addition to this difficulty, method C has the highest heat gains to be considered of any of the methods. The heat gain by convection and atmospheric condensation must be considered on the top of the ice as in the ice sample method and also on the water in the melt tank and considerable interconnecting piping. The heat input from the melt water circulating pump, P-5, also must be considered. The uncertainty in determining the heat gains of course affects the accuracy of the results.

Method C-1 again requires an accurate measurement of the small temperature difference, $T_7 - T_x$, which also is about 3^o to 4^o. This difference was measured by a thermo-pile using five thermocouples in each element so that the error in this measurement was minimized. Heat gains involve the top of the ice and the melt tank, but the melt water pump and some of the piping are omitted so that the heat gain factor is less than for method C.

Method D depends on a measurement of the vapor flow rate leaving the freezer. This cannot be measured directly but is measured as the condensate from the absorption generator. This is true as an average over a long period of several hours but may not be true over a short period of only a few hours.

The heat gain surface to be considered is quite large but very little atmospheric condensation is involved since most of the surface is rather well insulated. Heat gains from the freezer slurry pump, P-3, and the brine return pump, P-4, must be considered. Heat gains are relatively low and since atmospheric condensation is not a large factor, the heat gain varies less with ambient temperature than for methods C and C-1.

Methods C, C-1, and D were compared on the basis of 4 hour averages for two 24-hour test periods. The following tabulation shows the number of periods for which each method gave highest, lowest, or intermediate results compared with the others.

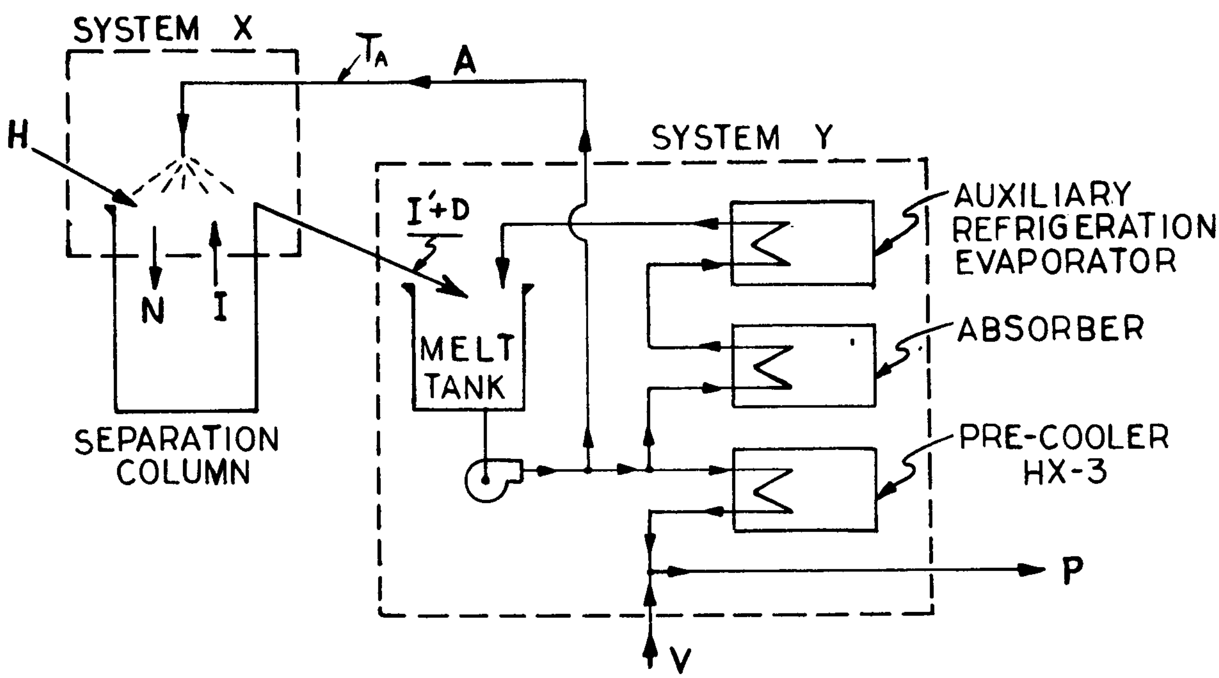
	Calculation Method		
	C	C-1	D
Highest Result	1	4	7
Intermediate Result	4	5	3
Lowest Result	7	3	2

The spread or difference between the highest and lowest result by the three methods varied from .2% (virtually exact agreement) to about 6% in general although one 4 hour period showed a maximum discrepancy of 10.2%. Over a 24 hour period the maximum spread amounted to only 2-1/2 to 3%. Over the total 48 hour period the averages for each method were as follows:

Method C	5.45%
Method C-1	6.70%
Method D	6.90%

The total spread here is only 1.45% and it will be noted that methods C-1 and D agree almost exactly. These comparisons demonstrate that all three methods are in good agreement when taken over a long period of time.

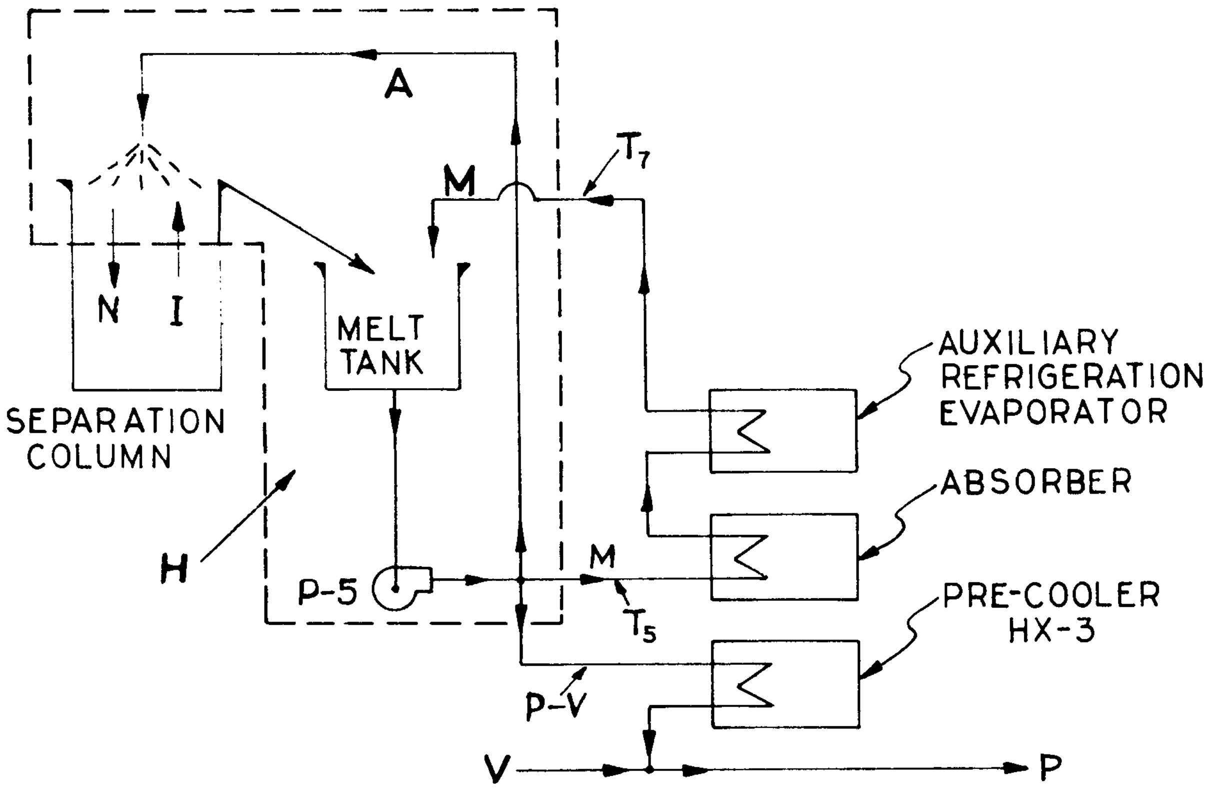
The ice sample method was used throughout most of the pilot plant program through December 1962. This was discarded and the alternative methods were developed when the scraper was changed to the chain driven conveyor type because of the difficulty of obtaining a representative ice sample with this scraper. Method C-1, employing the five element thermo-pile, then was selected for normal use as giving the most reliable results combined with ease of calculation.



A = GROSS WASH WATER TO COLUMN, GPM
D = WATER FROM COLUMN TO MELT TANK, GPM
H = HEAT GAINS, GALLON DEGREES/MIN = BTU/MIN ÷ 8.34
I = ICE RISING THROUGH COLUMN, GPM
I' = ICE TO MELT TANK, GPM
N = NET WASH, GPM
P = PRODUCT, GPM
T_A = TEMPERATURE OF WASH WATER, °F
V = VAPOR CONDENSATE, GPM
r = RATIO OF ICE TO TOTAL OF ICE PLUS WATER GOING TO MELT TANK, $r = I' \div (I' + D)$

NET WASH DETERMINATION BY ICE SAMPLE

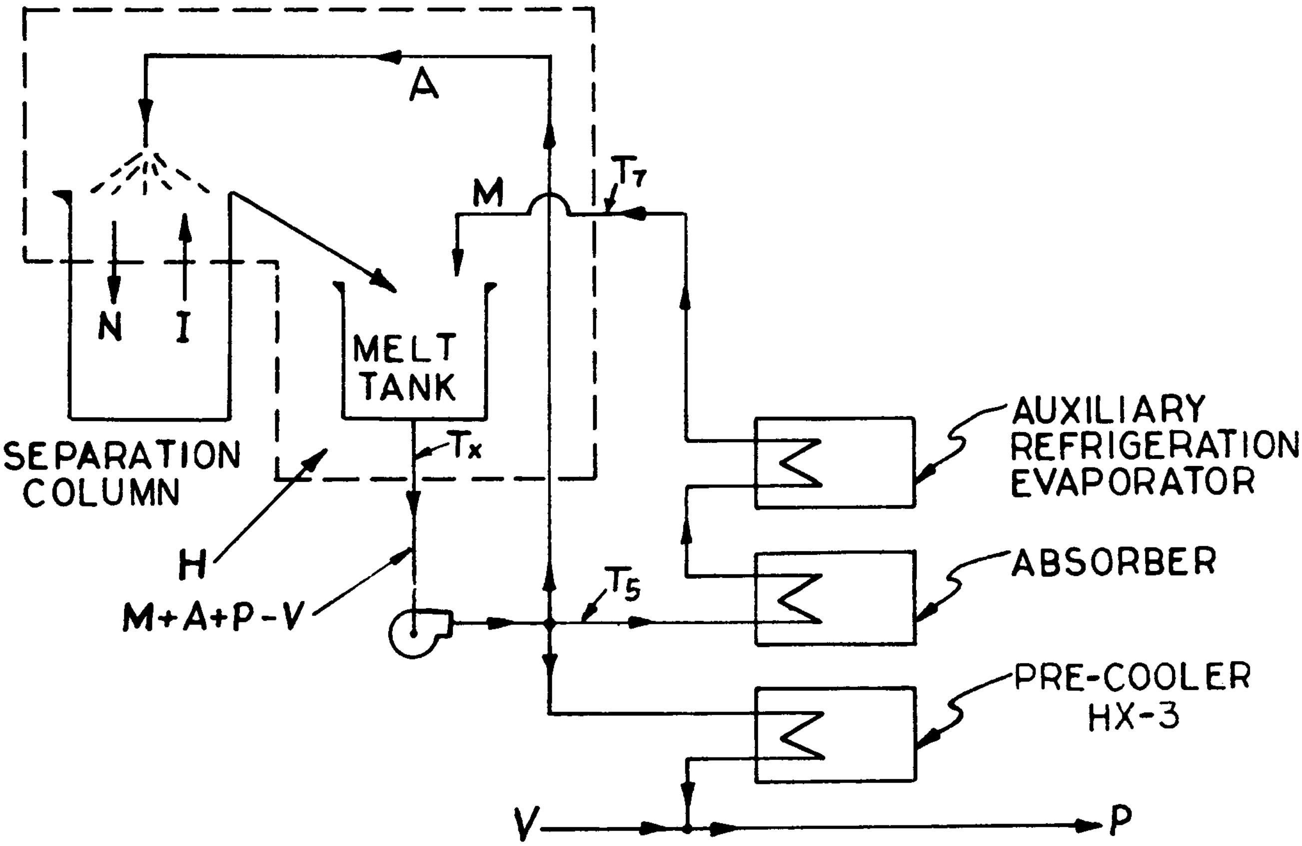
FIGURE A-1



H = HEAT GAINS, GALLON DEGREES/MIN = BTU/MIN ÷ 8.34
I = ICE RISING THROUGH COLUMN, GPM
M = FLOW OF WATER RETURNING TO MELT TANK, GPM
N = NET WASH, GPM
P = PRODUCT, GPM
T_5 = TEMPERATURE OF WATER LEAVING MELT TANK TAKEN AFTER THE PUMP, P-5, °F
T_7 = TEMPERATURE OF WATER RETURNING TO MELT TANK, °F
V = VAPOR CONDENSATE, GPM

NET WASH DETERMINATION
BY HEAT BALANCE
METHOD C

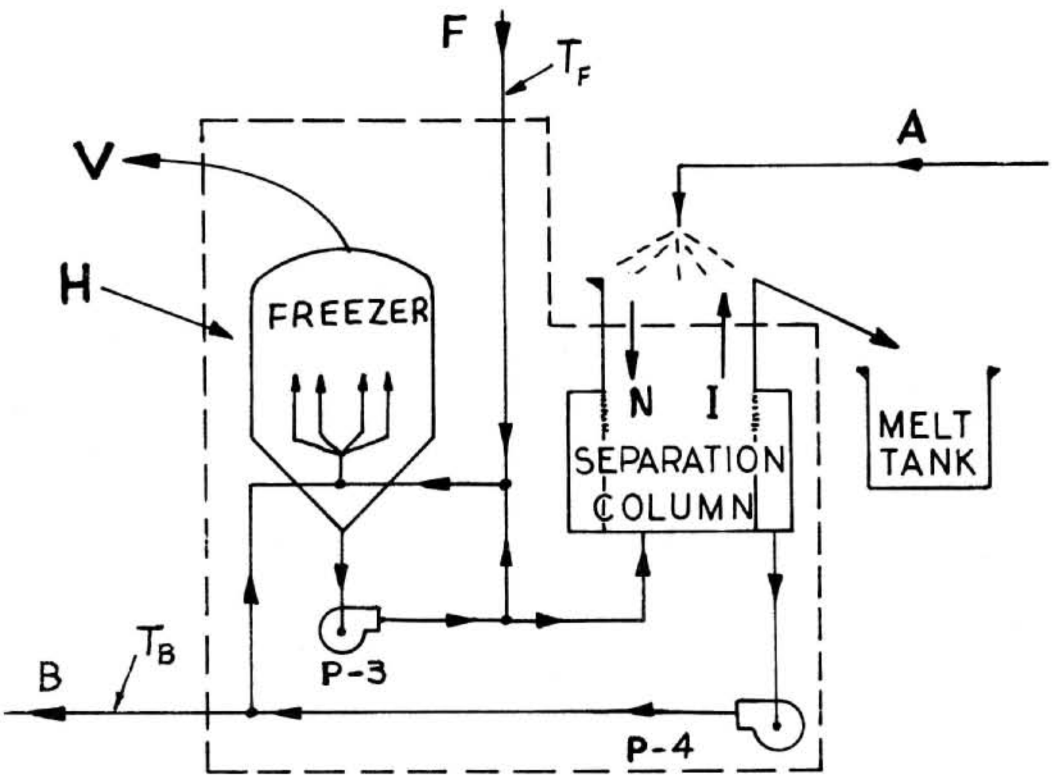
FIGURE A-2



A = GROSS WASH WATER TO COLUMN, GPM
H = HEAT GAINS, GALLON DEGREES/MIN = BTU/MIN ÷ 8.34
I = ICE RISING THROUGH COLUMN, GPM
M = FLOW OF WATER RETURNING TO MELT TANK, GPM
N = NET WASH, GPM
P = PRODUCT, GPM
T_5 = TEMPERATURE OF WATER LEAVING MELT TANK TAKEN AFTER THE PUMP, P-5, °F
T_7 = TEMPERATURE OF WATER RETURNING TO MELT TANK, °F
T_x = TEMPERATURE OF WATER LEAVING MELT TANK, °F
V = VAPOR CONDENSATE, GPM

NET WASH DETERMINATION
BY HEAT BALANCE
METHOD C-1

FIGURE A-3



B = WASTE BRINE, GPM
F = SEAWATER FEED, GPM
G_P = SPECIFIC HEAT, GALLON DEGREES PER GALLON PER °F WHICH IS NORMAL SPECIFIC HEAT X SPECIFIC GRAVITY
H = HEAT GAINS, GALLON DEGREES/MIN = BTU/MIN ÷ 8.34
I = ICE RISING THROUGH COLUMN, GPM
N = NET WASH, GPM
T_B = TEMPERATURE OF WASTE BRINE, °F
T_F = TEMPERATURE OF SEAWATER FEED, °F
V = VAPOR CONDENSATE, GPM
h = ENTHALPY OF VAPOR OUT OF FREEZER, GALLON DEGREES/GAL. = BTU/LB.

NET WASH DETERMINATION
BY VAPOR CONDENSATION
METHOD D

FIGURE A-4

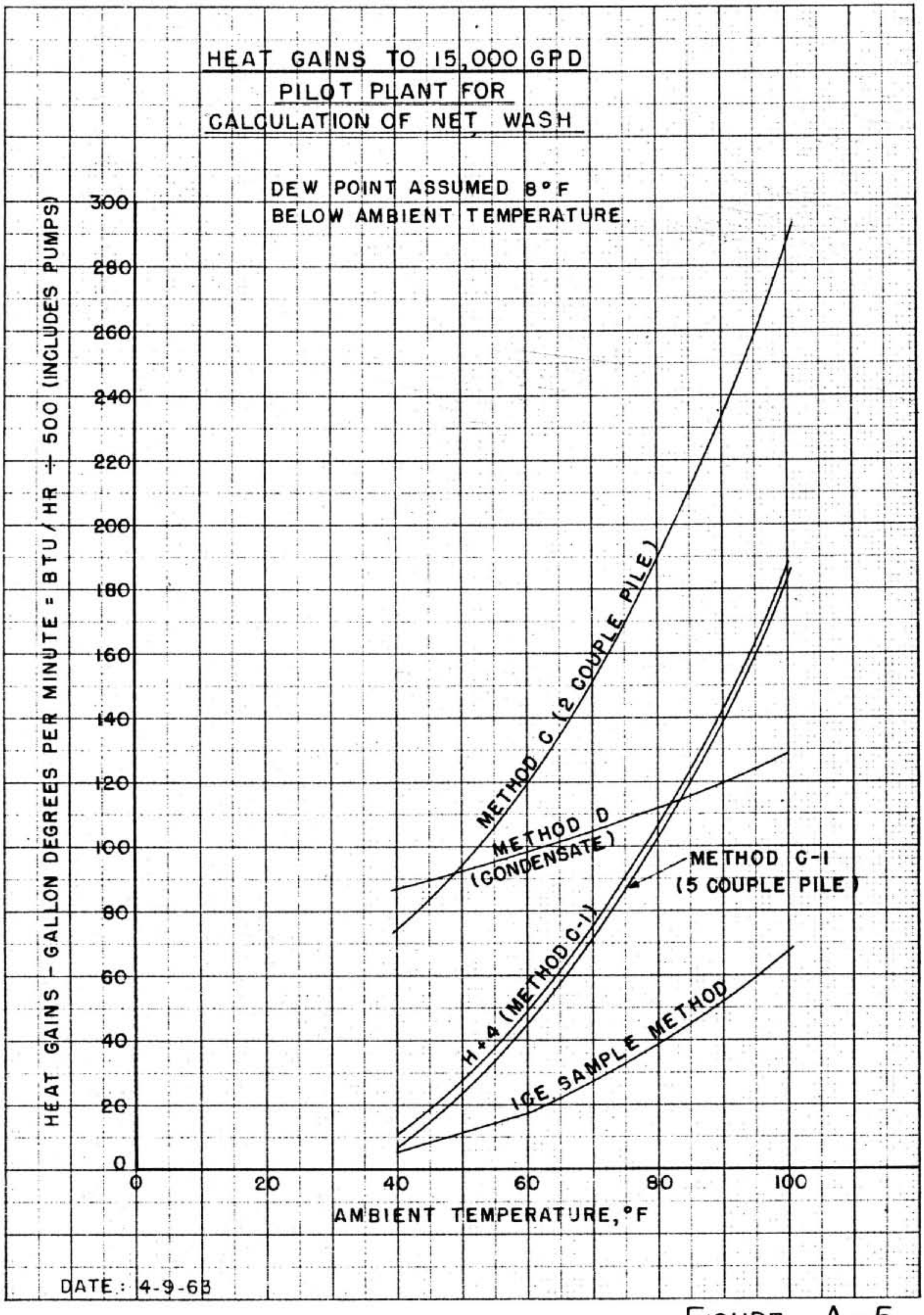

HEAT GAINS TO 15,000 GPD
PILOT PLANT FOR
CALCULATION OF NET WASH
DEW POINT ASSUMED 8°F
BELOW AMBIENT TEMPERATURE.
HEAT GAINS - GALLON DEGREES PER MINUTE = BTU/HR ÷ 500 (INCLUDES PUMPS)
300
280
260
240
220
200
180
160
140
120
100
80
60
40
20
0
METHOD C (2 COUPLE PILE)
METHOD D
(CONDENSATE)
METHOD C-I
(5 COUPLE PILE)
H+4 (METHOD C-I)
ICE SAMPLE METHOD
0
20
40
60
80
100
AMBIENT TEMPERATURE, °F
DATE: 4-9-63

FIGURE A-5

APPENDIX B

DETERMINATION OF EQUIVALENT ICE PARTICLE SIZE

Some basic concepts on the flow of fluids through porous media were investigated for application to the wash separation column under Contract No. 14-01-001-86 and are summarized in OSW Research and Development Progress Report No. 32. The flow characteristics are described by a permeability coefficient which can be related to the porosity of the bed and an equivalent diameter for the particles comprising the bed. During this report period techniques have been developed for obtaining permeability and porosity data which have proved useful in analyzing column performance.

Permeability is defined by the equation:

$$B_o = \frac{Q \mu L}{A \rho g h}$$

where:

B_o = permeability - cm^2

Q = flowrate - cc/sec.

μ = viscosity - gms/cm. sec.

L = bed length - cm

A = bed cross section - sq. cm.

ρ = fluid density - gm/cc

g = acceleration of gravity - cm/sec^2

h = head loss - cm

The equivalent particle diameter is defined by the equation:

$$d_e = \left[\frac{180\ B_o\ (1 - e)^2}{e^3}\right]^{\frac{1}{2}}$$

where:

d_e = diameter of equivalent spherical particle - cm

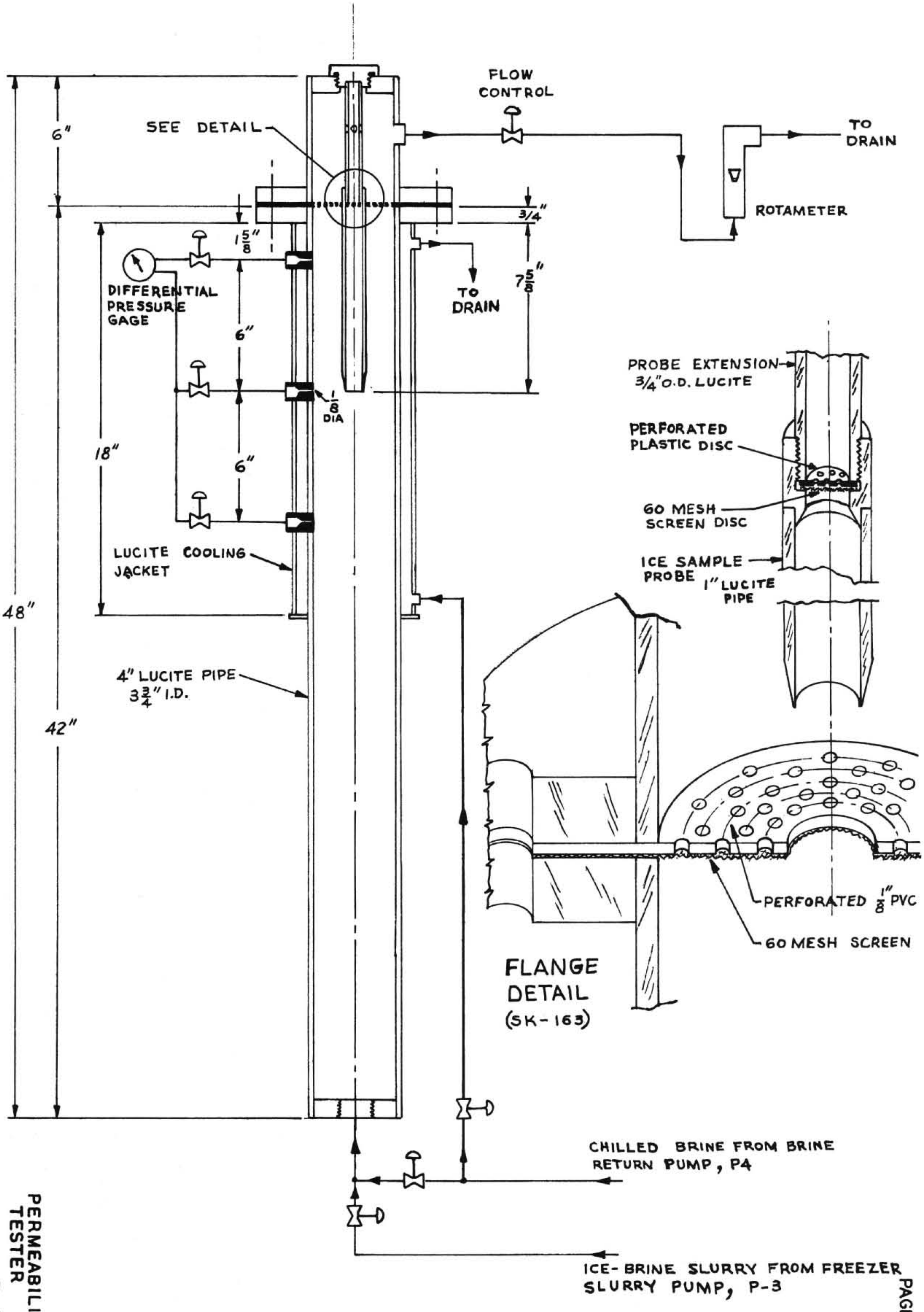
e = porosity - cc/cc

The equivalent diameter concept has proved extremely useful in evaluating changes in ice characteristics although no basis has been found for relating this value to actual measured crystal size, or distribution of crystal sizes, in a given sample. Also, attempts to relate the equivalent diameter of these ice particles taken directly from the freezer, with porosity and apparent permeability in the separation column, have been unsuccessful to date.

There are several reasons for believing that the calculated ice particle sizes are not completely accurate. It is considered important here to discuss the limitations of the apparatus and the factors which may affect the results. The apparatus is shown in Figure B-1. Permeability and porosity data are obtained by: 1) Passing slurry taken from the feed line to the column, through the test piece until a fairly solid bed, greater than 8 inches in length is established; 2) Measuring the pressure drop across 6 inches of the bed, and the flow through the bed; 3) Removing the center probe full of ice and running a calorimetric porosity determination. Brine from the column skirt is used to cool the jacket (approximately 3 - 5^{o}F above slurry temperature).

The actual plug of ice used for porosity measurement extends above the pressure taps used for permeability measurement. There appears to be no simple change that will correct this situation without introducing other errors. Therefore, a portion of the more highly compacted bed is used, and measured porosities are somewhat lower than the actual porosity in the applicable region of the ice bed. In the permeability measurement, the time required to build an adequate bed (dependent on slurry ice content) is important. Even though heat gains are small, the bed has a tendancy to erode. The erosion tends to occur primarily in the vicinity of the low pressure tap. This leads to obtaining lower pressures at this point, which results in erroneously high indicated differential pressures across the bed. This effect is considered to be small when the ice content in the slurry is high. Qualitatively, the possible error in porosity measurements should lead to larger calculated particle sizes and the errors in permeability measurements should lead to smaller calculated sizes. Of the two, the permeability measurements are considered to be most critical. No quantitative analysis is possible at this time.

The effect of both slurry temperature and recycle ratio is shown on Figure B-2. Most of the data was obtained during special test runs where the summary production data was irrevelent and was therefore not included in this report. However, spot checks have been made and compared with these curves on production runs throughout the report period. Finally, a graph relating the relative particle sizes determined at the pilot plant with the salinity controlled product rate obtained in various tests is shown on Figure B-3.



PERMEABILITY TESTER

PERMEABILITY TESTER
FIGURE B-1

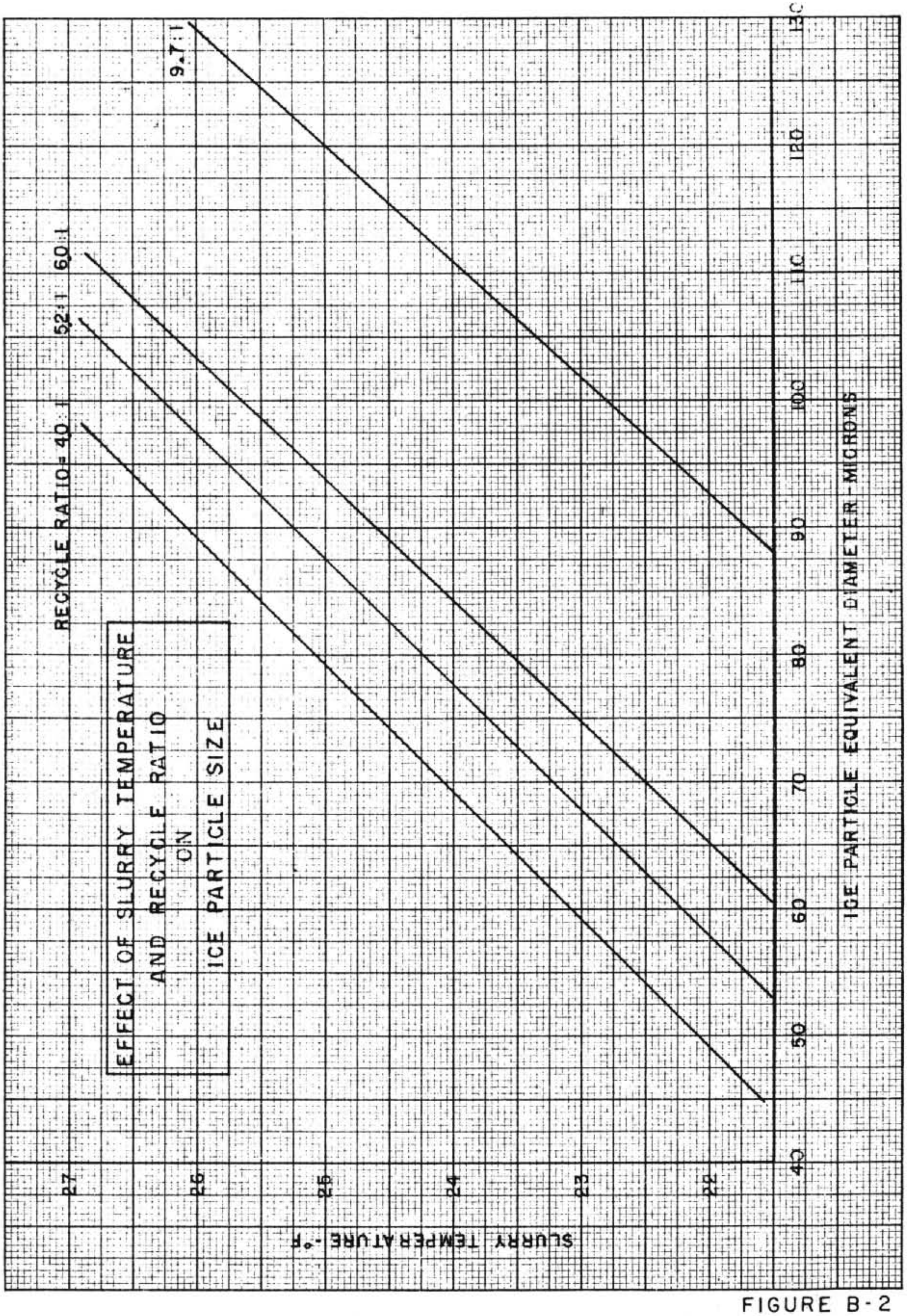

EFFECT OF SLURRY TEMPERATURE
AND RECYCLE RATIO
ON
ICE PARTICLE SIZE
RECYCLE RATIO= 40:1
52:1
60:1
9.7:1
SLURRY TEMPERATURE -°F
27
26
25
24
23
22
ICE PARTICLE EQUIVALENT DIAMETER-MICRONS
40
50
60
70
80
90
100
110
120
130

FIGURE B-2

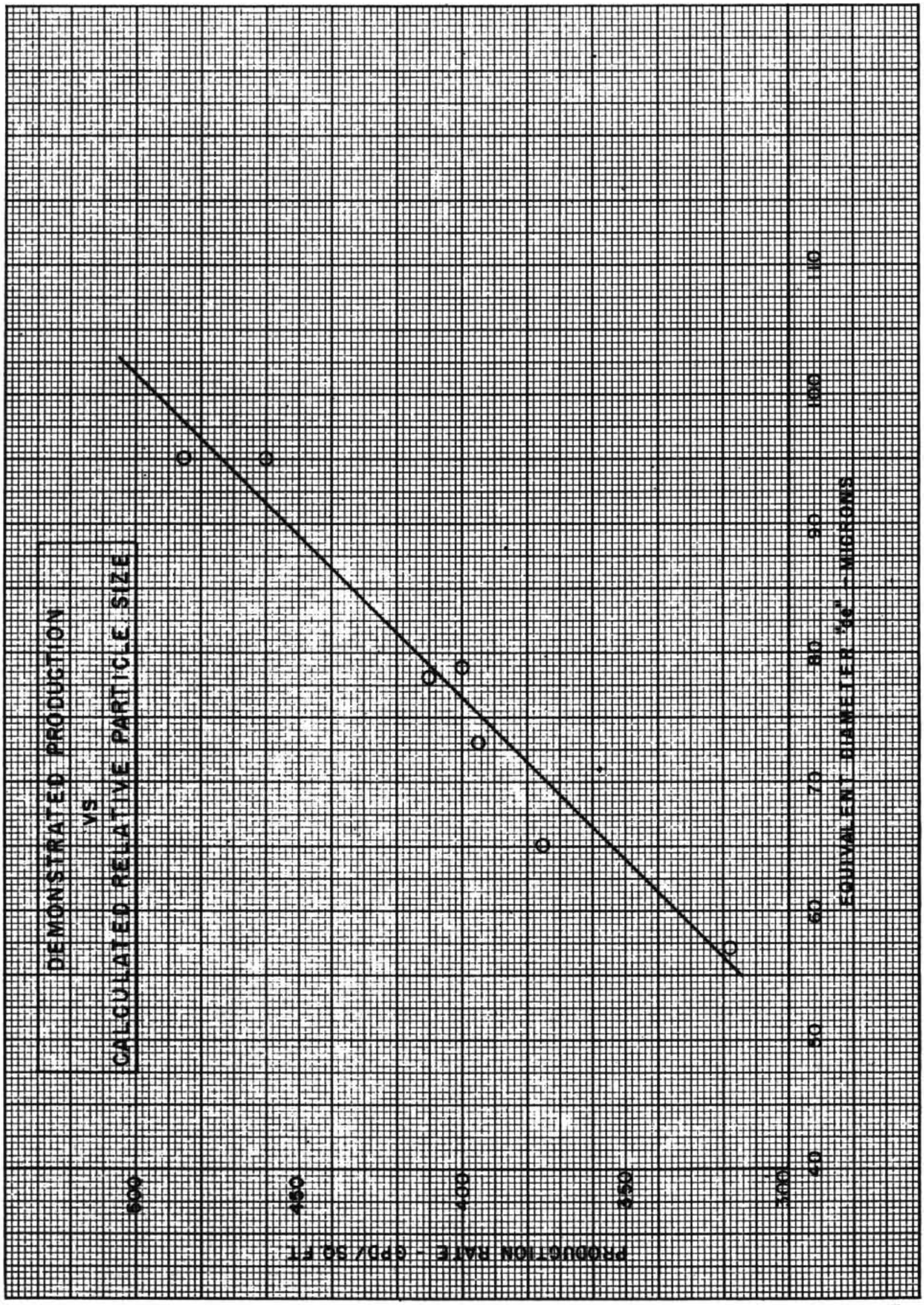

DEMONSTRATED PRODUCTION
VS
CALCULATED RELATIVE PARTICLE SIZE
EQUIVALENT DIAMETER "de" - MICRONS
40
50
60
70
80
90
100
110
PRODUCTION RATE - GPD/ SQ FT
300
350
400
450
500

FIGURE B-3

APPENDIX C

SEPARATION COLUMN MATERIALS OF CONSTRUCTION

Prior to this contract period most of the pilot plant testing was performed with a steel separation column. Rustoleum red primer was used to coat the internal surfaces, primarily as a corrosion inhibitor. As the overall performance of the plant was improved, it became evident that the performance of the separation column was being adversely affected by the tendency of ice to stick or freeze on the steel surfaces. The bolted panel column was specifically designed to investigate the effects of different materials on column performance and the abrasive effects of the ice on materials and coatings.

The first materials tested were 1/4" PVC sheet, steel plate painted with various primers, steel plate factory coated with a baked phenolic, and steel plate factory coated with a baked teflon phenolic. At this time the column contained two individual separation cells. One of these was fitted with the 1/4" PVC sheet and the other used painted steel. The performance of the cell containing primarily PVC sheet material was compared with the painted steel cell and showed conclusively better performance with PVC. The subsequent replacement of steel surfaces in contact with the ice piston is covered in the body of this report.

The two phenolic coatings were tested on specially prepared 2' x 3' steel plates. They were installed in the column center partition and were in service up until the removal of this partition, a period of about four months. Both coatings were found to have satisfactory abrasion resistance, and as far as could be determined from these relatively small surface areas, the friction characteristics were comparable to PVC. The primary disadvantage was the cost of these coatings and the necessity of factory application. The coating cost is about $1.60 per sq.ft. and adding the cost of the steel plate, a total cost of about $2.50 per sq.ft. is obtained. PVC sheet can be field applied at a material cost of roughly $2.50 per sq.ft. Since the cost of both materials is considered high, investigation of various field applied coatings was undertaken.

On the suggestion of one coating firm, two steel test panels spray coated with a five layer vinyl system were installed in the column in November, 1962. These coatings failed from abrasion after only two weeks of testing. The panels were returned to the manufacturer for evaluation. Poor surface preparation was postulated as a cause of the failure. However, the final coating was only 6 mils thick and this is probably not sufficient for abrasive service.

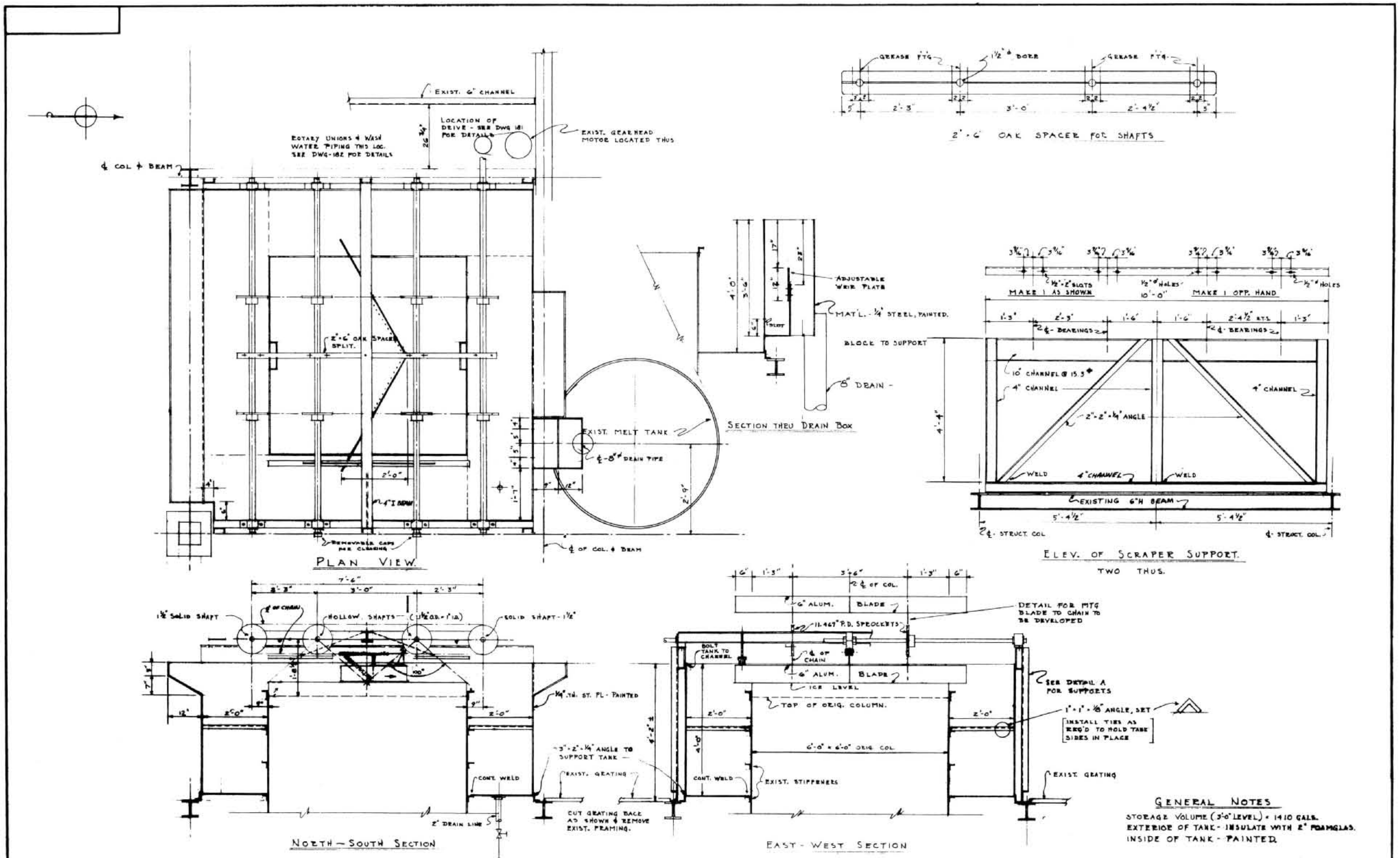
Epoxy coatings were next investigated for either a steel or concrete structure. A concrete structure seems to offer great advantages for field construction, considering the difficulty of proper field preparation of steel surfaces for any type of coating. Epoxy coatings in the range 10 - 30 mils thick have been successfully used in concrete tanks (notably in the nuclear power industry) for several years. This coating can be field applied for approximately $0.40 to $0.60 per sq.ft., representing a substanital savings in first cost.

Three sample slabs of concrete block were coated with different types of epoxy coatings. The surface characteristics appeared to be equal to PVC sheet in regard to surface friction and abrasion resistance. All samples withstood a submergence test in concentrated brine for about one month with no sign of deterioration. No tests simulating actual column wall conditions were possible.

An epoxy lined concrete column with PVC drain plates appears to have considerable potential. An experimental column of this type should be subjected to prolonged tests, however, before using this construction for a production plant.

DRAWING LIST

R1039-S14-180	Column Top Revision
RD1039-4196	Separation Column Evolution
RD1039-4197	Bottom Manifold Arrangements
RD1039-9113	Valving Diagram (Initial Design)
RD1039-9149	Separation Column Evolution (Prior to Present Contract)
RD1039-9150	Valving Diagram (Prior to Present Contract)
RD1039-9151	Bottom Manifold Arrangements (Prior to Present Contract)
RD1039-9153	Flow Scheme (Pilot Plant)
RD1039-9171	General Arrangement of 165,000 GPD Plant
RD1039-9174	Perspective of Proposed 165,000 GPD Plant
RD1039-9177	Flow Scheme (165,000 GPD Plant)
RD1039-9193	Valving Diagram (at Conclusion of Present Contract)



General Notes

STORAGE VOLUME (3'-0" LEVEL) = 1410 GALS.
EXTERIOR OF TANK - INSULATE WITH 2" FOAMGLAS.
INSIDE OF TANK - PAINTED.

NO.	REVISION	DATE	BY
2	REVISED PER CONST. CHANGES	1-25-65	J.C.
1	REVISED TO "AS BUILT"	1-8-65	E.S.F.

DO NOT SCALE THIS DRAWING

FOR: CHAIN DRIVE SCRAPER, RECTANGULAR MELT TROUGH.

BUYERS NO.

TITLE: ASSEMBLY DRAWING COLUMN TOP REVISION

DR. ESF 12-6-62 SCALE 3/4" = 1'-0"

CH.

Carrier

DWG. NO. R1039-514-180

DATE ISSUED -

SUPERSEDES ______ SUPERSEDED BY ______ REV. | DWG. NO. | SIMILAR TO ______

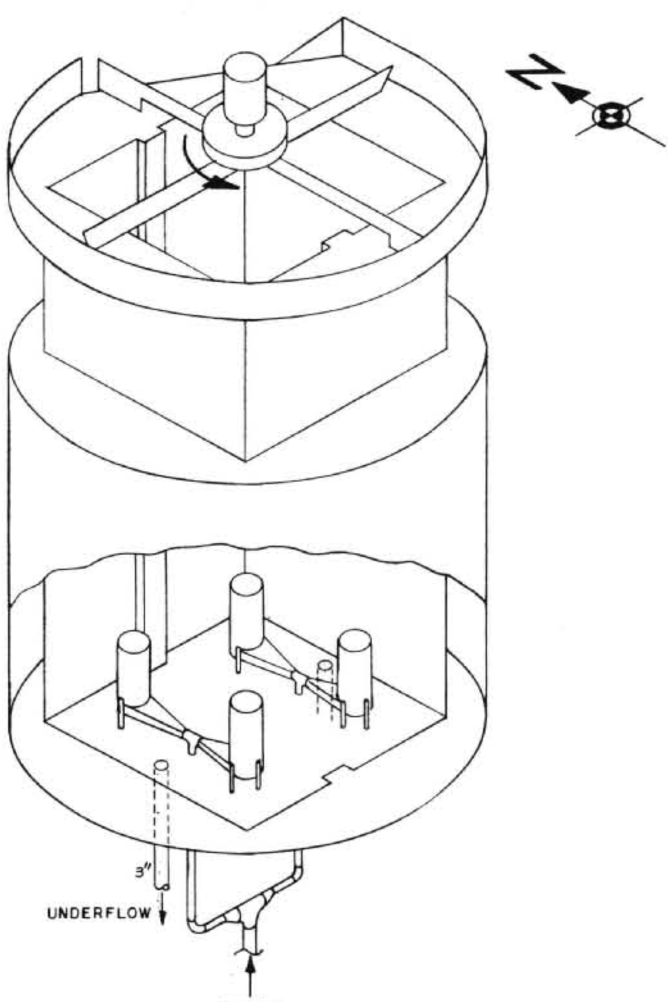
12

OCT 17, TO DEC 28, '62

BOLTED PANEL CONSTRUCTION
CENTER PARTITION REMOVED

SK-173 ; E.R.# :

FOR COLUMN ARRANGEMENT
FROM AUG 17, TO OCT 17, '62
SEE VIEW 11 DWG RD1039-9149

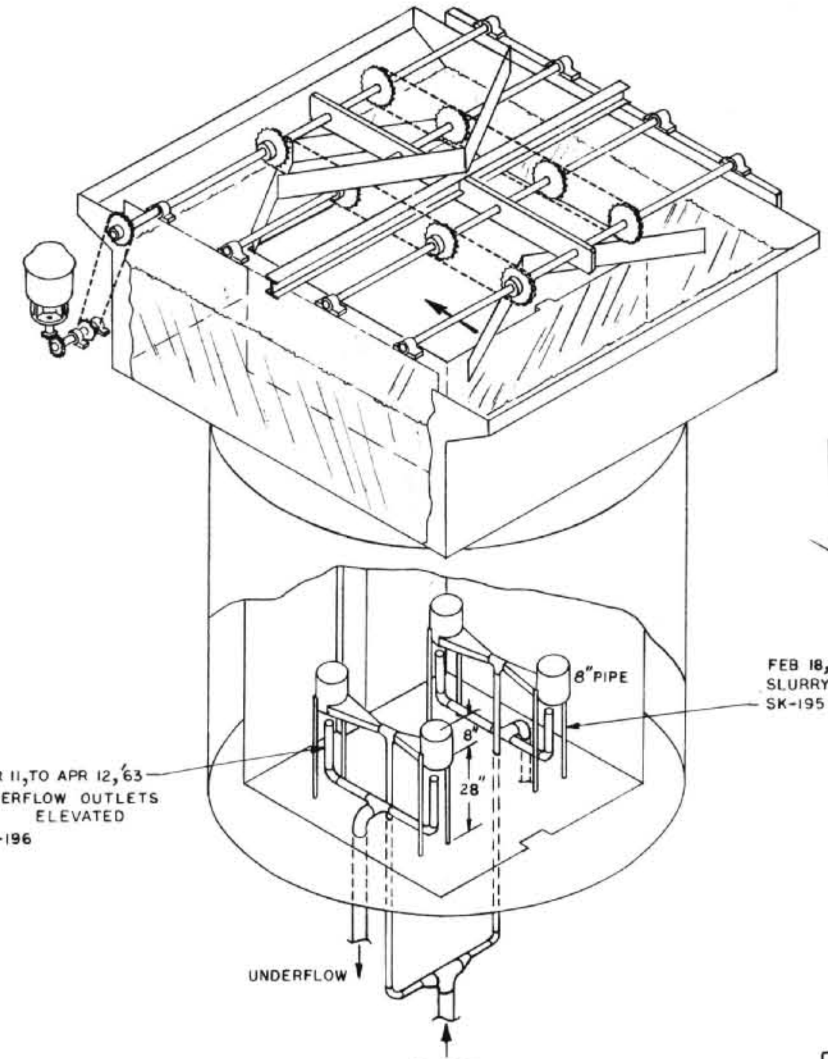


13

JAN 21, TO APR 12, '63

CHAIN DRIVE CONVEYOR TYPE SCRAPER
AND SURROUNDING SQUARE MELT TANK

DWG RD1039-S14-180 ; E.R.#3



MAR 11, TO APR 12, '63
UNDERFLOW OUTLETS
ELEVATED
SK-196

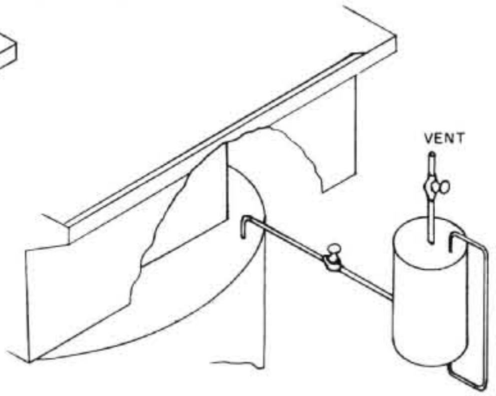
FEB 18, TO APR 12, '63
SLURRY INLETS ELEVATED
SK-195 ; E.R.# 3

13 A

MAR 18, TO APR 12, '63

COLUMN SKIRT BACK PRESSURE
CONTROL

SK-202



LEGEND:

SK-NO = ENGINEERING SKETCHES

E.R.# = ENGINEERING (PROGRESS) REPORTS BY CARRIER TO O.S.W.

SYM.	REVISION RECORD	DATE	BY	APPD. BY	CH'KD. BY	E.C.R. NO.

USED ON ASSY				
AUTHORIZATION NO. R1039-S14				
MANUFACTURED	PURCHASED			
APPROVED FOR	ENGRG.	DATE	MFG.	DATE
TRIAL PROD. RUN				
FINAL PROD.				

MATERIAL

APPLIED FINISH

TOLERANCES UNLESS OTHERWISE SPECIFIED

FRACTIONS	DECIMALS	ANGLES
±	±	±
SURFACES	HOLE SPACING	HOLE LOCATION
√ AIN.	±	±
	NON-CUMULATIVE	

DR. [signature] 26 APR 63 — CH'KD. [signature] May 2, 63 — SCALE (DO NOT SCALE DRAWING)

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SYRACUSE, NEW YORK, U.S.A.

SEPARATION COLUMN EVOLUTION
SALINE WATER CONVERSION
PILOT PLANT
WRIGHTSVILLE BEACH, N.C.

DWG. NO. RD1039 - 4196

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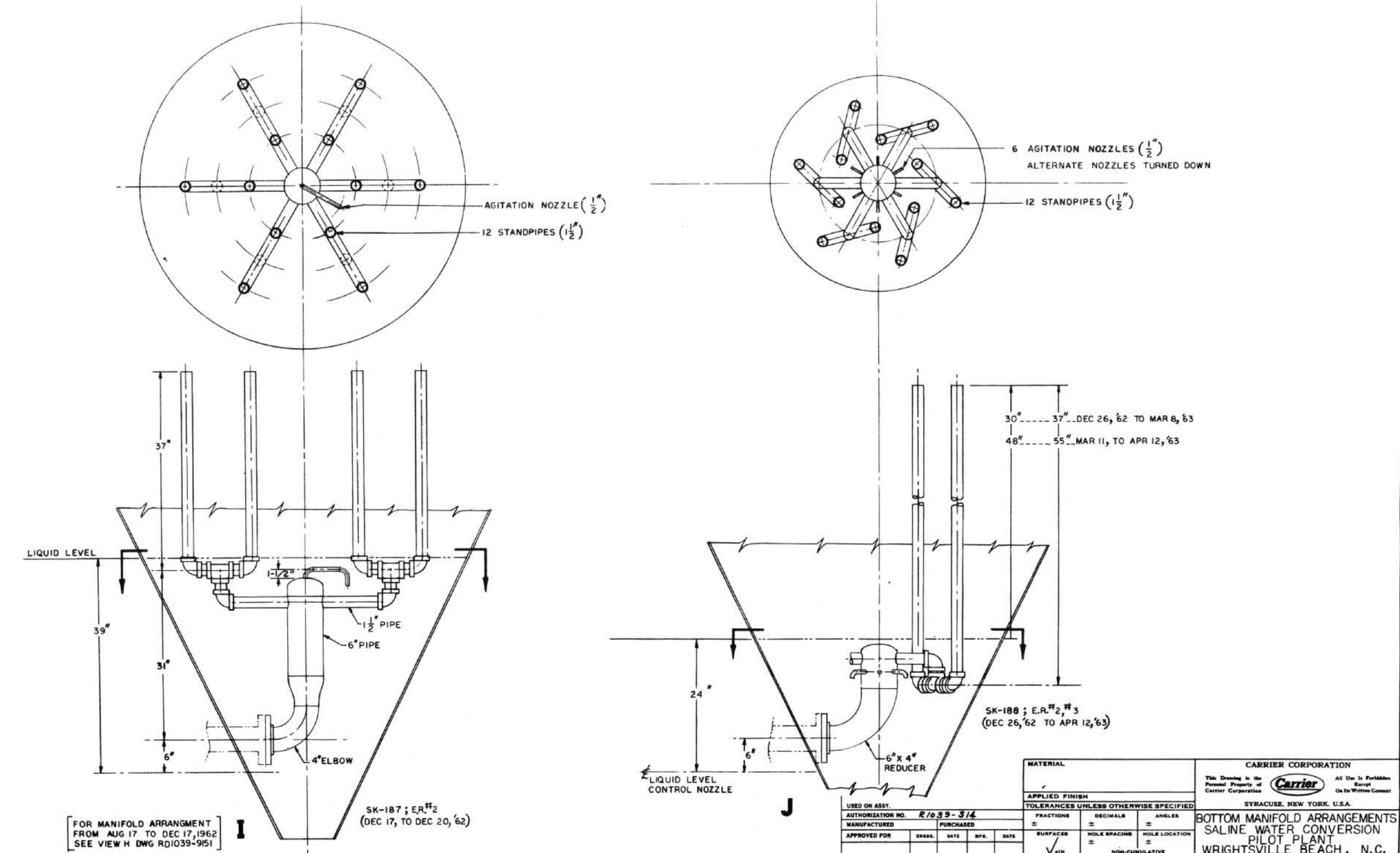

SUPERSEDES ____ SUPERSEDED BY ____ REV. | DWG. NO. | SIMILAR TO ____

SYM.	REVISION RECORD	DATE	BY	APPD. BY	CH'KD. BY	E.C.R. NO.

USED ON ASSY.
AUTHORIZATION NO. R1039-314
MANUFACTURED | PURCHASED
APPROVED FOR | ENGRG. | DATE | MFG. | DATE
TRIAL PROD. RUN
FINAL PROD.

MATERIAL
APPLIED FINISH
TOLERANCES UNLESS OTHERWISE SPECIFIED
FRACTIONS ± | DECIMALS ± | ANGLES ±
SURFACES √ AIN. | HOLE SPACING ± | HOLE LOCATION ±
NON-CUMULATIVE
DR. LaPointe 27 Apr 63 | SCALE (DO NOT SCALE DRAWING)
CH'KD. 2 May 63

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BOTTOM MANIFOLD ARRANGEMENTS
SALINE WATER CONVERSION
PILOT PLANT
WRIGHTSVILLE BEACH, N.C.

DWG. NO. RD1039 - 4197

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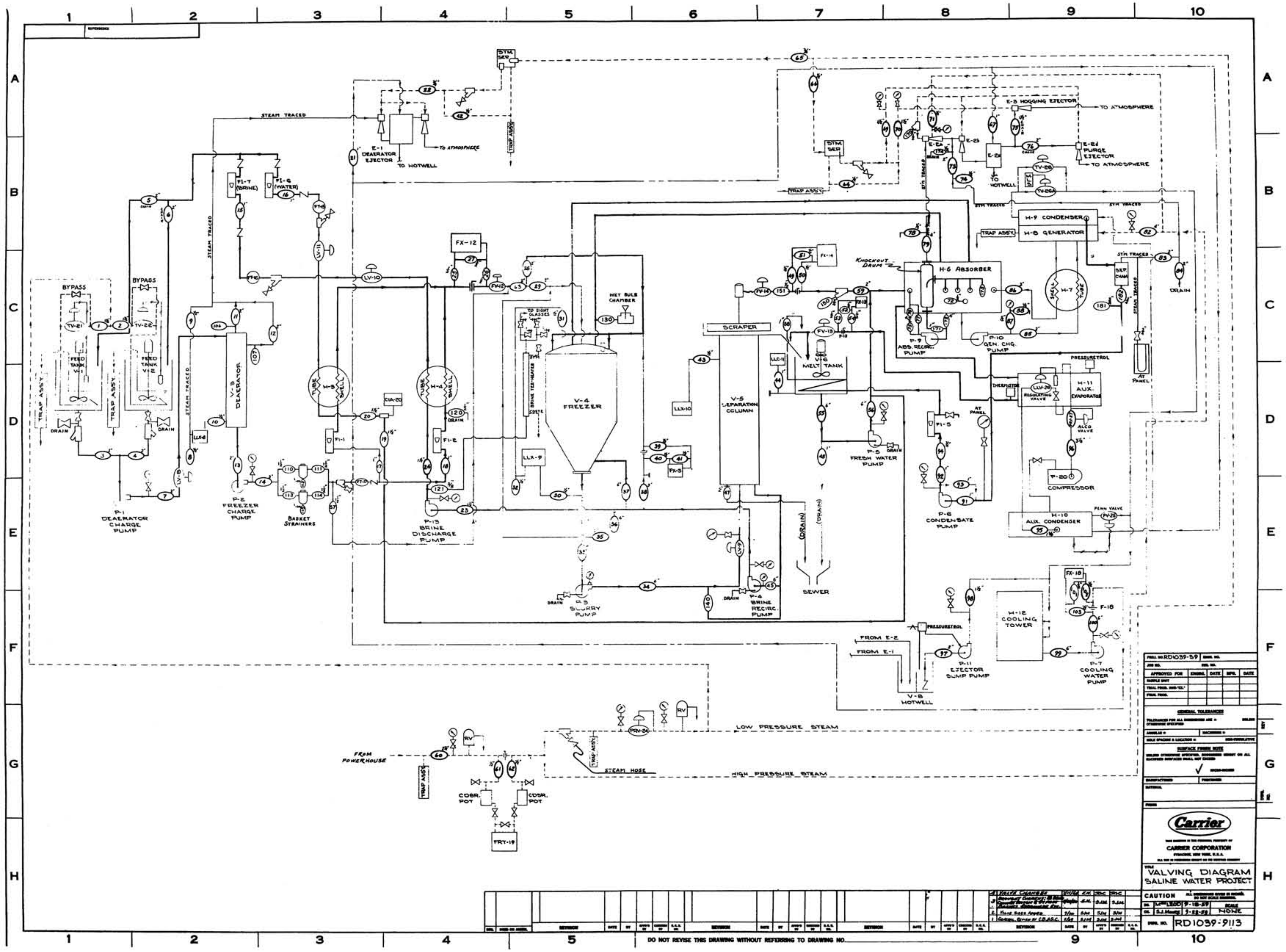
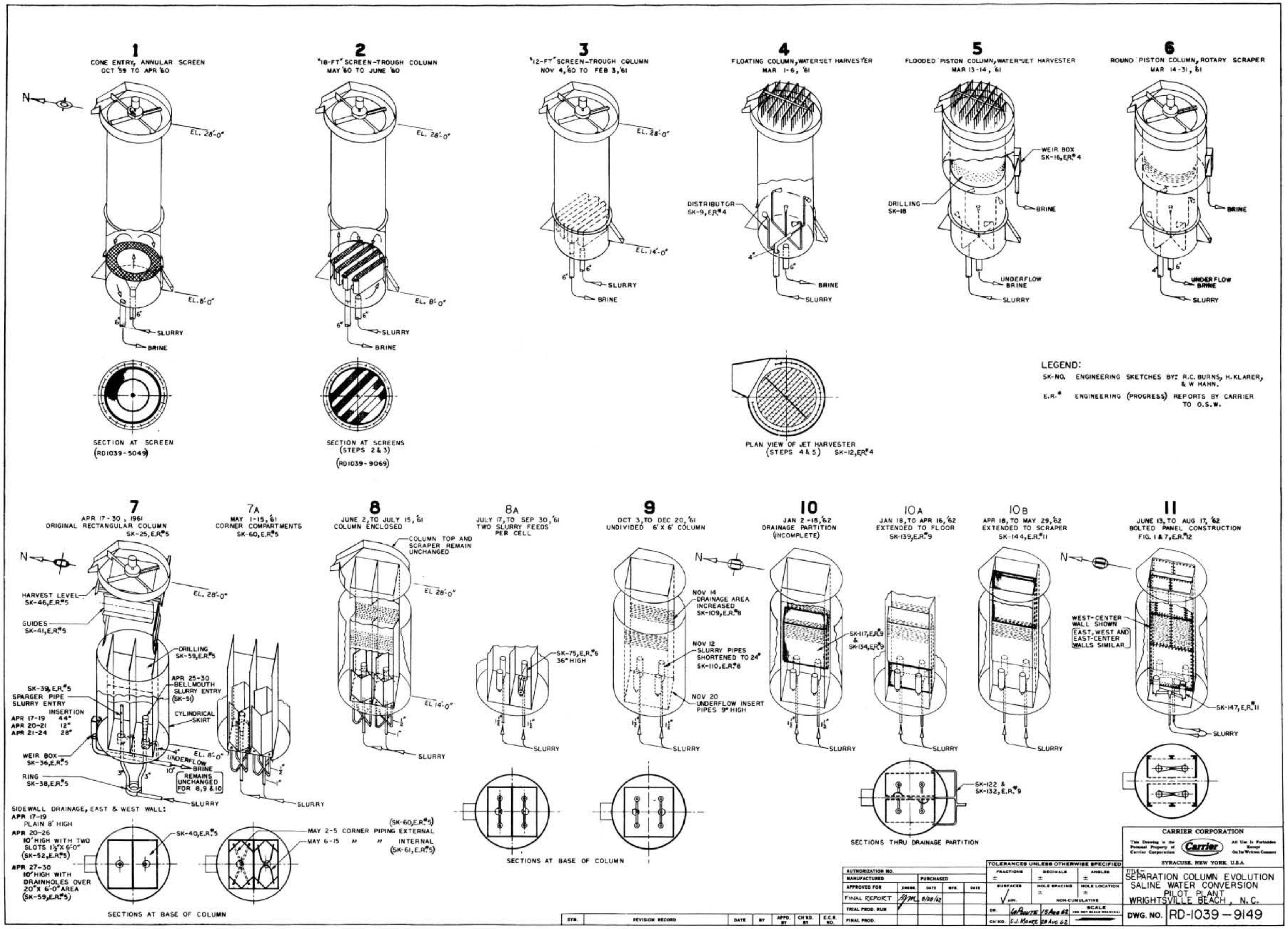


STEAM TRACED
E-1 DEAERATOR EJECTOR
TO ATMOSPHERE
TO HOTWELL
BYPASS
FEED TANK V-1
FEED TANK V-2
TRAP ASSY
DRAIN
P-1 DEAERATOR CHARGE PUMP
V-3 DEAERATOR
P-2 FREEZER CHARGE PUMP
BASKET STRAINERS
FI-7 (BRINE)
FI-6 (WATER)
H-3 TUBE SHELL
H-4 TUBE SHELL
P-13 BRINE DISCHARGE PUMP
FX-12
WET BULB CHAMBER
V-4 FREEZER
TO SIGHT GLASSES
BRINE TEE-HEATER
P-3 SLURRY PUMP
P-4 BRINE RECIRC. PUMP
SCRAPER
V-5 SEPARATION COLUMN
V-6 MELT TANK
(DRAIN)
SEWER
P-5 FRESH WATER PUMP
KNOCKOUT DRUM
H-6 ABSORBER
P-9 ABS. RECIRC. PUMP
P-10 GEN. CHG. PUMP
H-7 SHELL TUBE
H-9 CONDENSER
H-8 GENERATOR
STM SEP
E-3 HOGGING EJECTOR
E-24 PURGE EJECTOR
P-6 CONDENSATE PUMP
H-11 AUX. EVAPORATOR
REGULATING VALVE
THERMOSTAT
PRESSURETROL
AT PANEL
ALCO VALVE
P-20 COMPRESSOR
H-10 AUX. CONDENSER
PENN VALVE
H-12 COOLING TOWER
P-7 COOLING WATER PUMP
P-11 EJECTOR SUMP PUMP
FROM E-2
FROM E-1
V-8 HOTWELL
LOW PRESSURE STEAM
HIGH PRESSURE STEAM
STEAM HOSE
FROM POWERHOUSE
CDSR. POT
FRT-19
Carrier
CARRIER CORPORATION
VALVING DIAGRAM
SALINE WATER PROJECT
CAUTION
NONE
RD1039-9113
DO NOT REVISE THIS DRAWING WITHOUT REFERRING TO DRAWING NO.


1
CONE ENTRY, ANNULAR SCREEN
OCT '59 TO APR '60
EL. 28'-0"
EL. 8'-0"
SLURRY
BRINE
SECTION AT SCREEN
(RD1039-5049)
2
"18-FT" SCREEN-TROUGH COLUMN
MAY '60 TO JUNE '60
EL. 28'-0"
EL. 8'-0"
SLURRY
BRINE
SECTION AT SCREENS
(STEPS 2 & 3)
(RD1039-9069)
3
"12-FT" SCREEN-TROUGH COLUMN
NOV 4, '60 TO FEB 3, '61
EL. 28'-0"
EL. 14'-0"
SLURRY
BRINE
4
FLOATING COLUMN, WATER-JET HARVESTER
MAR 1-6, '61
DISTRIBUTOR
SK-9, E.R.# 4
BRINE
SLURRY
PLAN VIEW OF JET HARVESTER
(STEPS 4 & 5) SK-12, E.R.# 4
5
FLOODED PISTON COLUMN, WATER-JET HARVESTER
MAR 13-14, '61
WEIR BOX
SK-16, E.R.# 4
DRILLING
SK-18
BRINE
UNDERFLOW
BRINE
SLURRY
6
ROUND PISTON COLUMN, ROTARY SCRAPER
MAR 14-31, '61
BRINE
UNDERFLOW
BRINE
SLURRY
LEGEND:
SK-NO. ENGINEERING SKETCHES BY: R.C. BURNS, H. KLARER, & W HAHN.
E.R.# ENGINEERING (PROGRESS) REPORTS BY CARRIER TO O.S.W.
7
APR 17-30, 1961
ORIGINAL RECTANGULAR COLUMN
SK-25, E.R.# 5
HARVEST LEVEL
SK-46, E.R.# 5
GUIDES
SK-41, E.R.# 5
EL. 28'-0"
DRILLING
SK-59, E.R.# 5
APR 25-30
BELLMOUTH
SLURRY ENTRY
(SK-51)
CYLINDRICAL
SKIRT
SK-39, E.R.# 5
SPARGER PIPE
SLURRY ENTRY
INSERTION
APR 17-19 44"
APR 20-21 12"
APR 21-24 28"
WEIR BOX
SK-36, E.R.# 5
EL. 8'-0"
UNDERFLOW
BRINE
REMAINS UNCHANGED FOR 8, 9 & 10
RING
SK-38, E.R.# 5
SLURRY
SIDEWALL DRAINAGE, EAST & WEST WALL:
APR 17-19
PLAIN 8' HIGH
APR 20-26
10' HIGH WITH TWO SLOTS 1½" X 6'-0"
(SK-52, E.R.# 5)
APR 27-30
10' HIGH WITH DRAINHOLES OVER 20" X 6'-0" AREA
(SK-59, E.R.# 5)
SK-40, E.R.# 5
SECTIONS AT BASE OF COLUMN
7A
MAY 1-15, '61
CORNER COMPARTMENTS
SK-60, E.R.# 5
SLURRY
MAY 2-5 CORNER PIPING EXTERNAL (SK-60, E.R.# 5)
MAY 6-15 " " INTERNAL (SK-61, E.R.# 5)
8
JUNE 2, TO JULY 15, '61
COLUMN ENCLOSED
COLUMN TOP AND SCRAPER REMAIN UNCHANGED
EL. 28'-0"
EL. 14'-0"
SLURRY
8A
JULY 17, TO SEP 30, '61
TWO SLURRY FEEDS PER CELL
SK-75, E.R.# 6
36" HIGH
SLURRY
9
OCT 3, TO DEC 20, '61
UNDIVIDED 6' X 6' COLUMN
NOV 14
DRAINAGE AREA INCREASED
SK-109, E.R.# 8
NOV 12
SLURRY PIPES SHORTENED TO 24"
SK-110, E.R.# 8
NOV 20
UNDERFLOW INSERT PIPES 9" HIGH
SLURRY
SECTIONS AT BASE OF COLUMN
10
JAN 2-18, '62
DRAINAGE PARTITION
(INCOMPLETE)
SK-117, E.R.# 9
&
SK-134, E.R.# 9
SLURRY
10A
JAN 18, TO APR 16, '62
EXTENDED TO FLOOR
SK-139, E.R.# 9
SLURRY
SK-122 &
SK-132, E.R.# 9
SECTIONS THRU DRAINAGE PARTITION
10B
APR 18, TO MAY 29, '62
EXTENDED TO SCRAPER
SK-144, E.R.# 11
SLURRY
11
JUNE 13, TO AUG 17, '62
BOLTED PANEL CONSTRUCTION
FIG. 1 & 7, E.R.# 12
WEST-CENTER WALL SHOWN
EAST, WEST AND EAST-CENTER WALLS SIMILAR
SK-147, E.R.# 11
SLURRY
CARRIER CORPORATION
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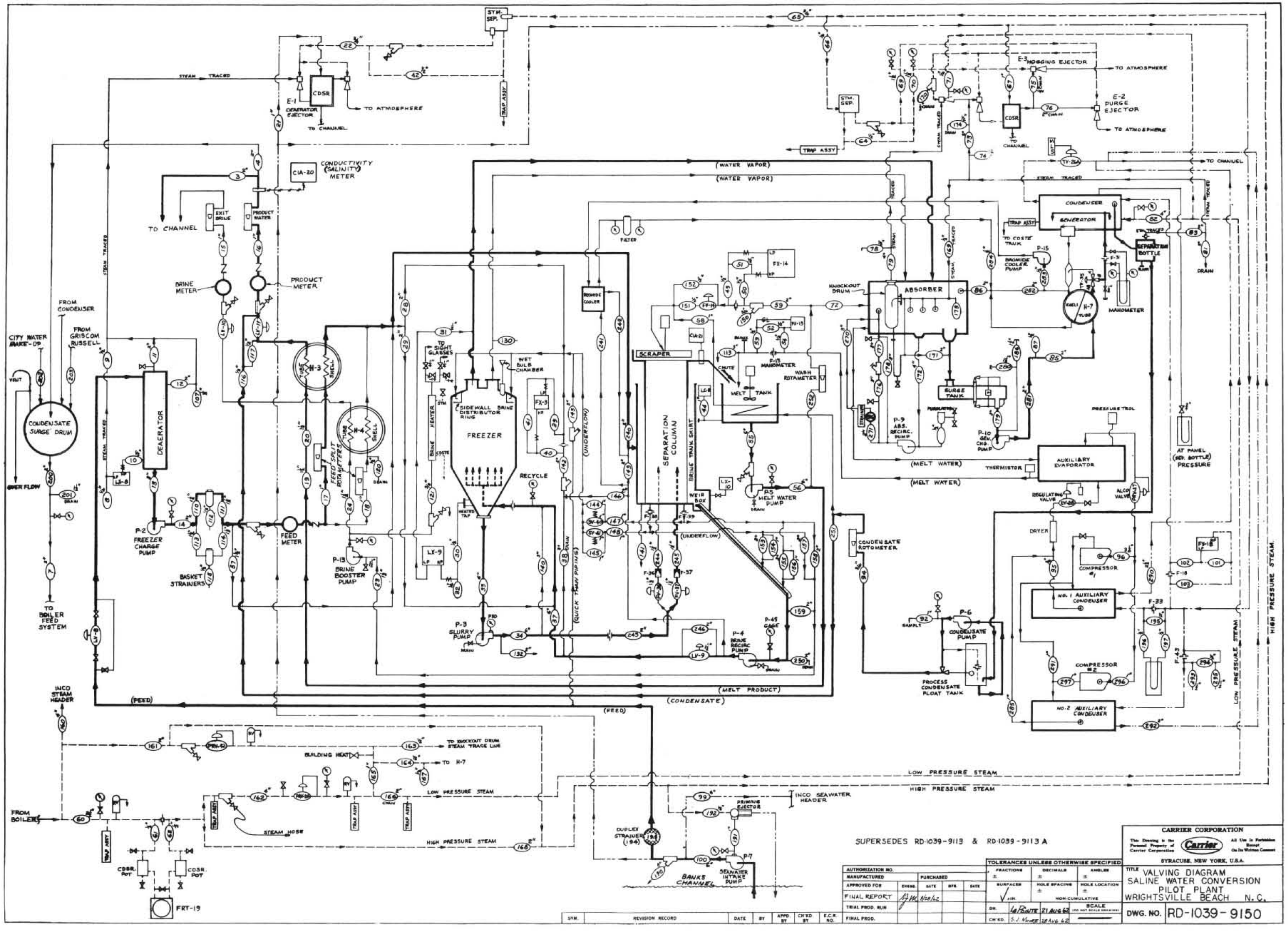
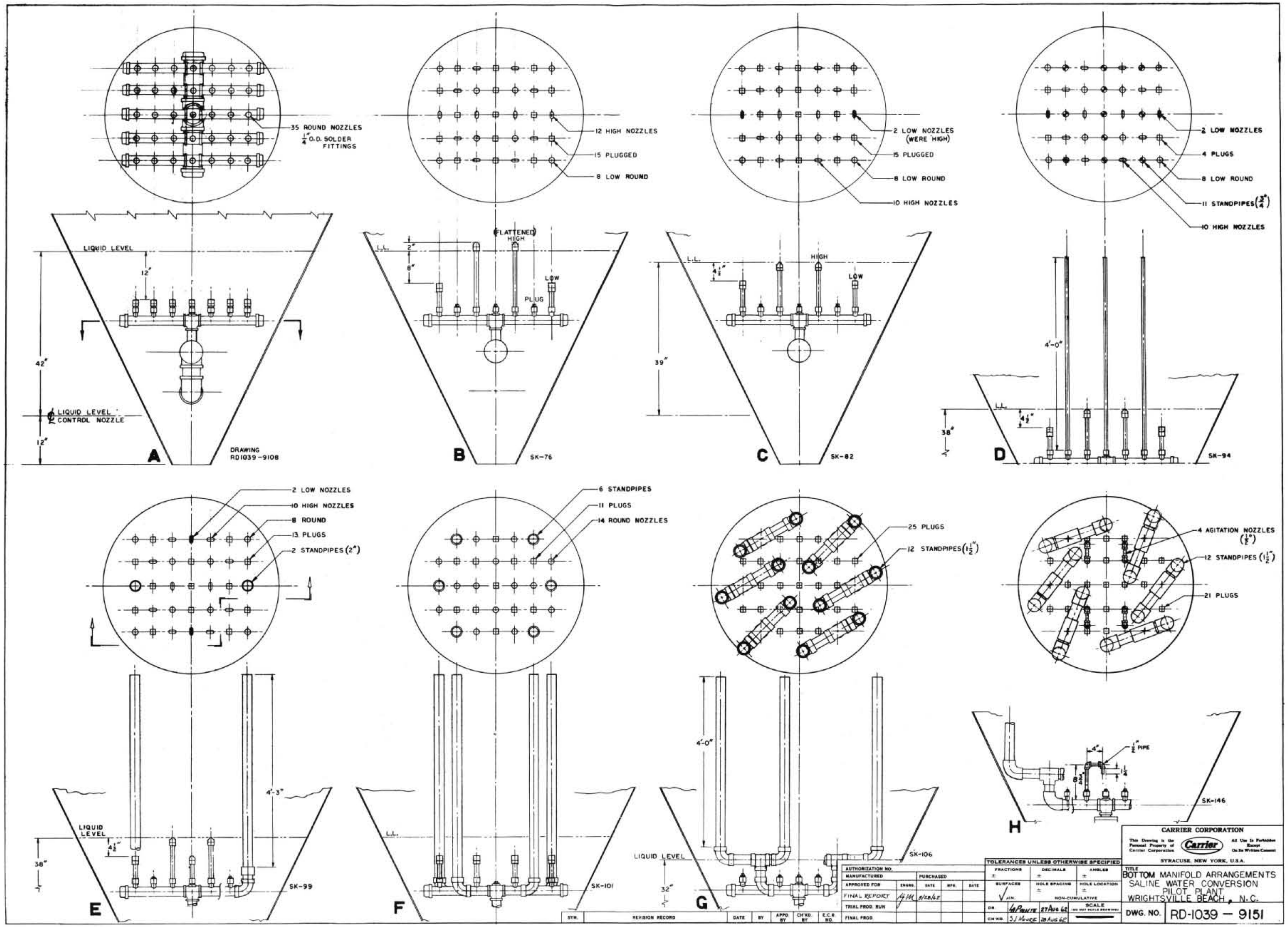
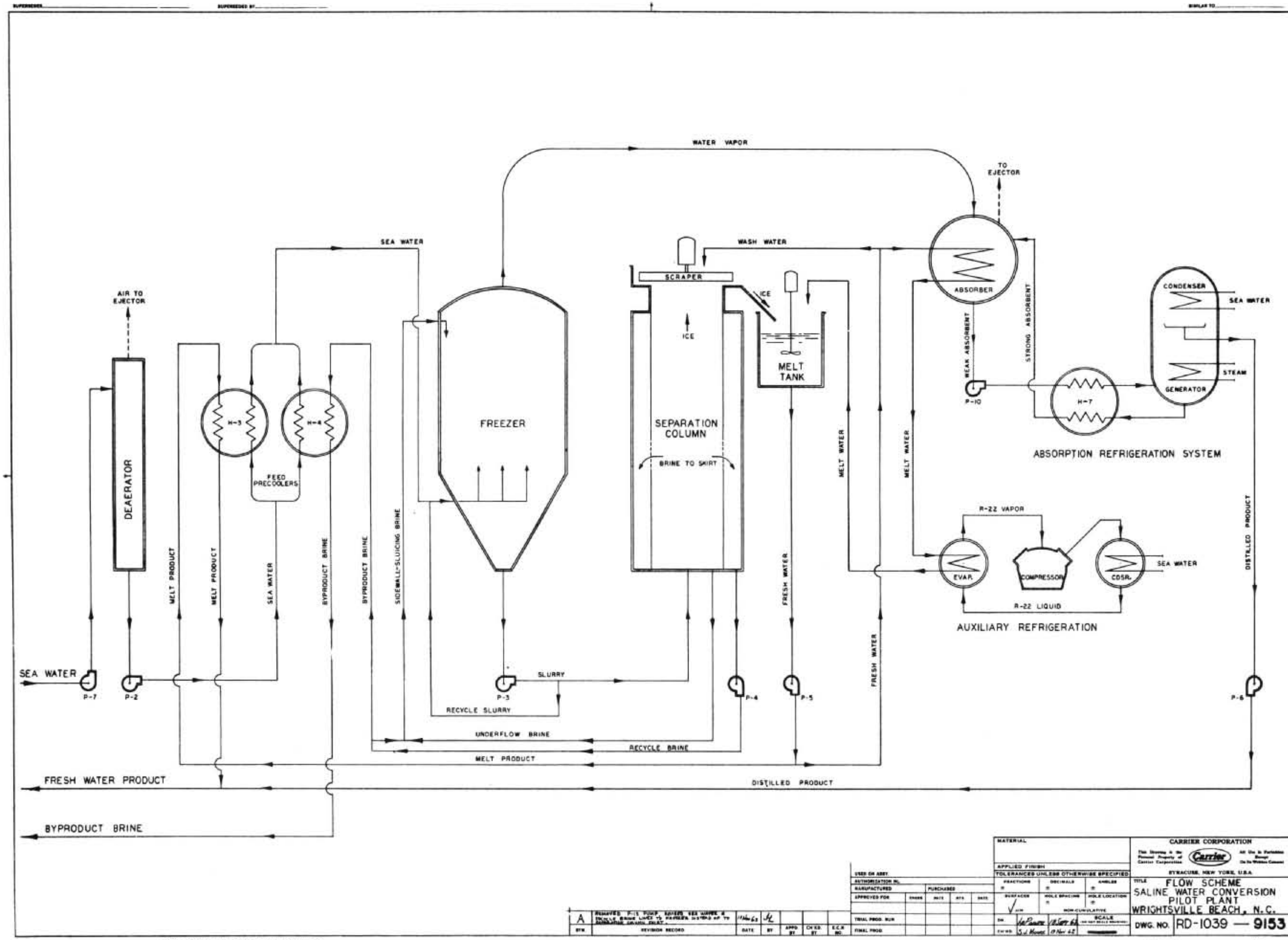
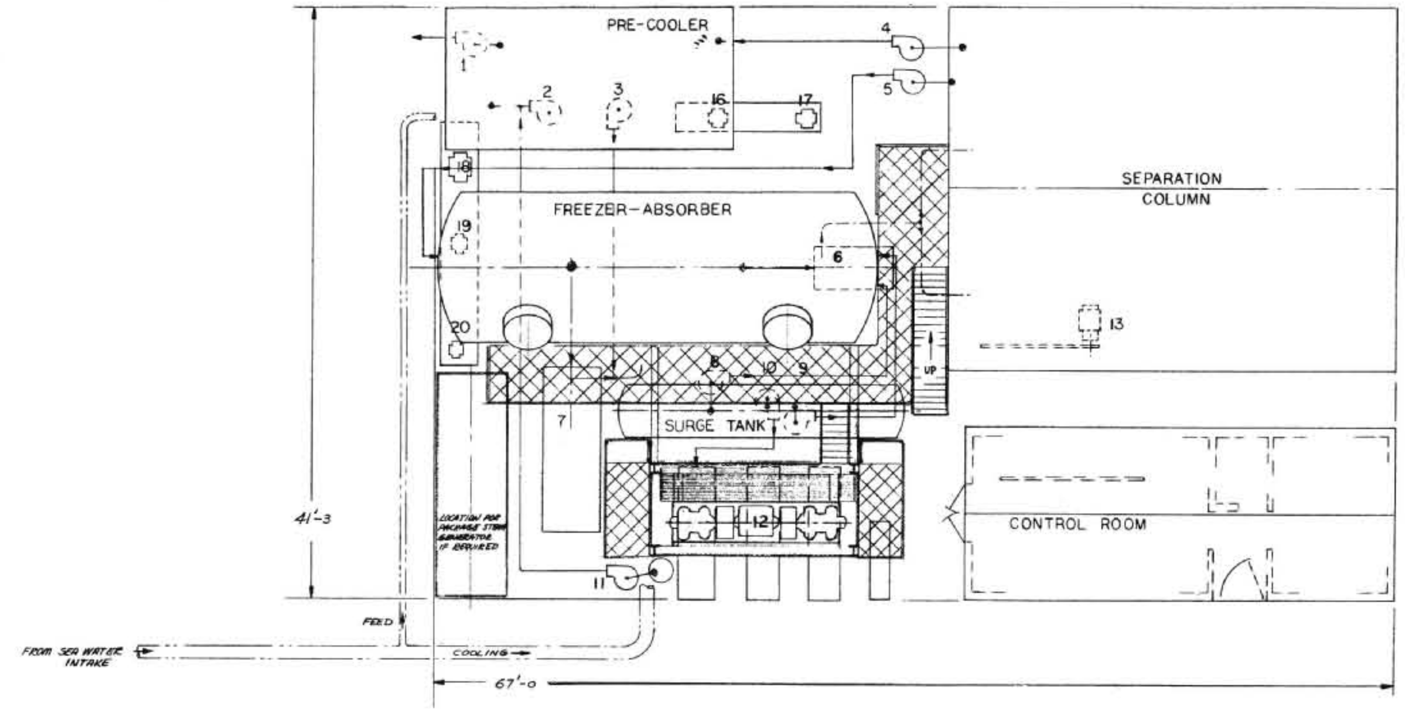
SYRACUSE, NEW YORK, U.S.A.
TITLE — SEPARATION COLUMN EVOLUTION
SALINE WATER CONVERSION
PILOT PLANT
WRIGHTSVILLE BEACH, N.C.
DWG. NO. RD-1039-9149
TOLERANCES UNLESS OTHERWISE SPECIFIED
FRACTIONS ±
DECIMALS ±
ANGLES ±
SURFACES
HOLE SPACING ±
HOLE LOCATION ±
NON-CUMULATIVE
SCALE
DR. LeBOUTE 15 Aug 62
CH'KD. S.J. Moore 29 Aug 62
AUTHORIZATION NO.
MANUFACTURED
PURCHASED
APPROVED FOR
FINAL REPORT 8/20/62
TRIAL PROD. RUN
FINAL PROD.
SYM.
REVISION RECORD
DATE
BY
APPD. BY
CH'KD. BY
E.C.R. NO.

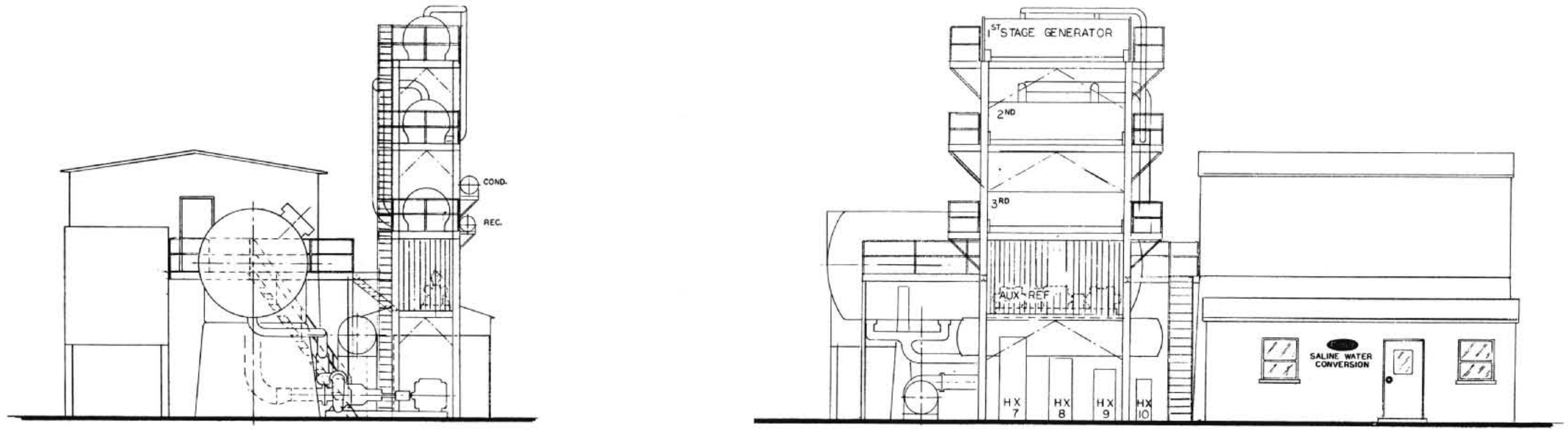

SUPERSEDES RD-1039-9113 & RD-1039-9113 A
CARRIER CORPORATION
SYRACUSE, NEW YORK, U.S.A.
VALVING DIAGRAM
SALINE WATER CONVERSION
PILOT PLANT
WRIGHTSVILLE BEACH N. C.
DWG. NO. RD-1039-9150
FINAL REPORT
CONDENSATE SURGE DRUM
DEAERATOR
FREEZER
SEPARATION COLUMN
ABSORBER
CONDENSER
GENERATOR
AUXILIARY EVAPORATOR
NO. 1 AUXILIARY CONDENSER
NO. 2 AUXILIARY CONDENSER
COMPRESSOR #1
COMPRESSOR #2
LOW PRESSURE STEAM
HIGH PRESSURE STEAM
(WATER VAPOR)
(MELT WATER)
(MELT PRODUCT)
(CONDENSATE)
(FEED)
TO CHANNEL
TO ATMOSPHERE
FROM BOILER
CITY WATER MAKE-UP
BANKS CHANNEL
INCO STEAM HEADER
INCO SEAWATER HEADER


35 ROUND NOZZLES
1/4" O.D. SOLDER FITTINGS
LIQUID LEVEL
12"
42"
℄ LIQUID LEVEL CONTROL NOZZLE
12"
A
DRAWING RD1039-9108
12 HIGH NOZZLES
15 PLUGGED
8 LOW ROUND
(FLATTENED) HIGH
L.L.
2"
8"
PLUG
LOW
B
SK-76
2 LOW NOZZLES (WERE HIGH)
15 PLUGGED
8 LOW ROUND
10 HIGH NOZZLES
HIGH
LOW
39"
C
SK-82
2 LOW NOZZLES
4 PLUGS
8 LOW ROUND
11 STANDPIPES (3/4")
10 HIGH NOZZLES
4'-0"
38"
D
SK-94
2 LOW NOZZLES
10 HIGH NOZZLES
8 ROUND
13 PLUGS
2 STANDPIPES (2")
4'-3"
LIQUID LEVEL
38"
E
SK-99
6 STANDPIPES
11 PLUGS
14 ROUND NOZZLES
L.L.
F
SK-101
25 PLUGS
12 STANDPIPES (1 1/2")
4'-0"
LIQUID LEVEL
32"
G
SK-106
4 AGITATION NOZZLES (1/2")
12 STANDPIPES (1 1/2")
21 PLUGS
1/2" PIPE
H
SK-146
CARRIER CORPORATION
SYRACUSE, NEW YORK, U.S.A.
TITLE BOTTOM MANIFOLD ARRANGEMENTS
SALINE WATER CONVERSION
PILOT PLANT
WRIGHTSVILLE BEACH, N.C.
DWG. NO. RD-1039 — 9151
TOLERANCES UNLESS OTHERWISE SPECIFIED
FINAL REPORT
REVISION RECORD


WATER VAPOR
TO EJECTOR
SEA WATER
WASH WATER
SCRAPER
ICE
AIR TO EJECTOR
ABSORBER
CONDENSER
SEA WATER
WEAK ABSORBENT
STRONG ABSORBENT
MELT TANK
STEAM
GENERATOR
P-10
H-7
H-3
H-4
FREEZER
SEPARATION COLUMN
DEAERATOR
BRINE TO SKIRT
FEED PRECOOLERS
ABSORPTION REFRIGERATION SYSTEM
MELT WATER
R-22 VAPOR
MELT PRODUCT
SEA WATER
BYPRODUCT BRINE
SIDEWALL-SLUICING BRINE
FRESH WATER
EVAP.
COMPRESSOR
CDSR.
DISTILLED PRODUCT
R-22 LIQUID
AUXILIARY REFRIGERATION
SEA WATER
P-7
P-2
P-3
SLURRY
P-4
P-5
P-6
RECYCLE SLURRY
UNDERFLOW BRINE
RECYCLE BRINE
MELT PRODUCT
FRESH WATER PRODUCT
DISTILLED PRODUCT
BYPRODUCT BRINE
CARRIER CORPORATION
SYRACUSE, NEW YORK, U.S.A.
FLOW SCHEME
SALINE WATER CONVERSION
PILOT PLANT
WRIGHTSVILLE BEACH, N.C.
DWG. NO. RD-1039 — 9153

NO	NAME
1	PRODUCT BOOSTER PUMP
2	CONDENSATE PUMP
3	FEED PUMP
4	BRINE DISCHARGE PUMP
5	MELT WATER PUMP
6	SLURRY TRANSFER PUMP
7	SLURRY RECYCLE PUMP
8	ABSORBER PUMP
9	EVAPORATOR PUMP
10	GENERATOR PUMP
11	PROCESS CONDENSATE PUMP
12	AUX. REF. COMPRESSORS MOTOR
13	SCRAPER DRIVE MOTOR
	SEAL WATER PUMP
	SEA WATER INTAKE PUMP
16	DEAERATION VACUUM PUMP A
17	" " " B
18	PURGE VACUUM PUMP A
19	" " " B
20	" " " C





SYM.	REVISION RECORD	DATE	BY	APPD. BY	CHKD. BY	E.C.R. NO.	

USED ON ASSY.
AUTHORIZATION NO. 1033-514
MANUFACTURED / PURCHASED
APPROVED FOR
TRIAL PROD. RUN
FINAL PROD.

MATERIAL
APPLIED FINISH
TOLERANCES UNLESS OTHERWISE SPECIFIED
FRACTIONS ± DECIMALS ± ANGLES ±
SURFACES ✓ HOLE SPACING ± HOLE LOCATION ±
NON-CUMULATIVE
SCALE 1/8" = 1'-0
DR'N. KCB 2-22-63

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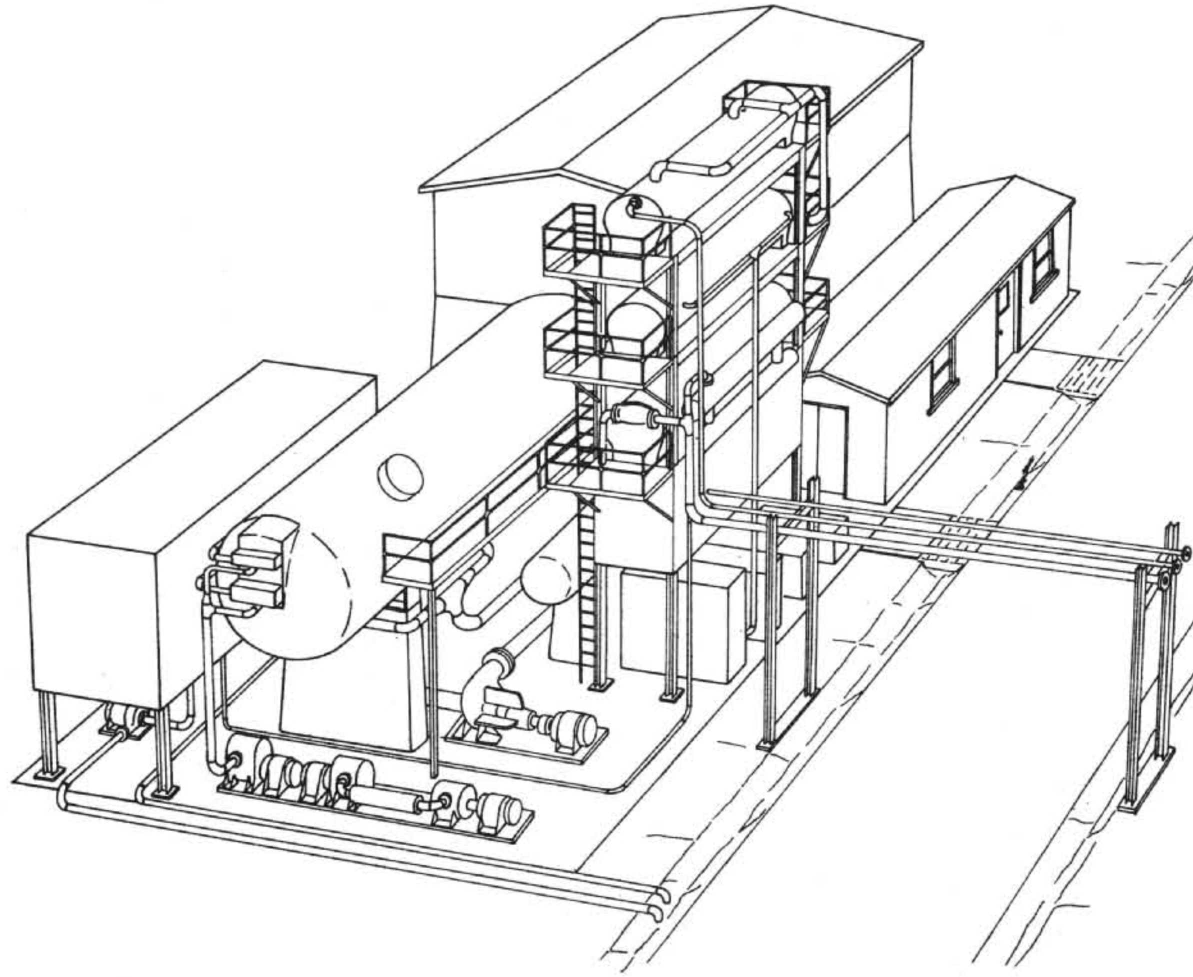
SYRACUSE, NEW YORK, U.S.A.

TITLE
GENERAL ARRANGEMENT OF PLANT
165,000 GPD
ABSORPTION-FREEZING PROCESS

DWG. NO. R 1039-9171



SYM.	REVISION RECORD	DATE	BY	APPD. BY	CHKD. BY	E.C.R. NO.

USED ON ASSY.
AUTHORIZATION NO. 1039-S14
MANUFACTURED / PURCHASED
APPROVED FOR | ENGRG. | DATE | MFG. | DATE
TRIAL PROD. RUN
FINAL PROD.

MATERIAL
APPLIED FINISH
TOLERANCES UNLESS OTHERWISE SPECIFIED
FRACTIONS ± | DECIMALS ± | ANGLES ±
SURFACES | HOLE SPACING ± | HOLE LOCATION ±
NON-CUMULATIVE
DR. LaPointe | 7 FEB 63 | SCALE (DO NOT SCALE DRAWING)
CH'KD. RCB | 2-22-63 | ———

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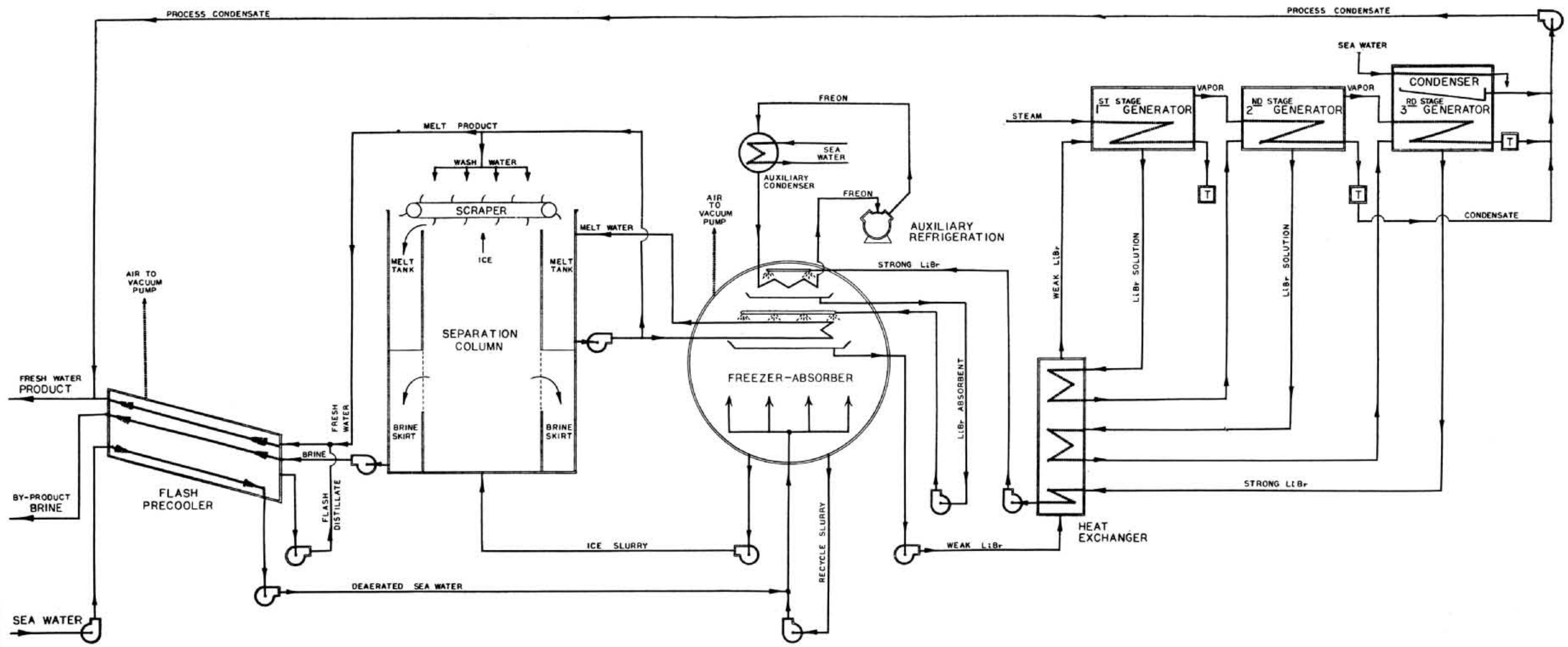
TITLE
PERSPECTIVE OF PROPOSED PLANT
165,000 GPD
ABSORPTION-FREEZE PROCESS

DWG. NO. R1039-9174



SYM.	REVISION RECORD	DATE	BY	APPD. BY	CH'KD. BY	E.C.R. NO.

USED ON ASSY.					
AUTHORIZATION NO.					
MANUFACTURED		PURCHASED			
APPROVED FOR	ENGRG.	DATE	MFG.	DATE	
Final Report	S.J.M.	2/5/63			
TRIAL PROD. RUN					
FINAL PROD.					

MATERIAL

APPLIED FINISH

TOLERANCES UNLESS OTHERWISE SPECIFIED

FRACTIONS	DECIMALS	ANGLES
±	±	±
SURFACES	HOLE SPACING	HOLE LOCATION
√ MIN.	±	±
	NON-CUMULATIVE	

DR.	[illegible]	5 FEB 63	SCALE
CH'KD.	S.J. Moore	5 FEB 63	NONE

CARRIER CORPORATION

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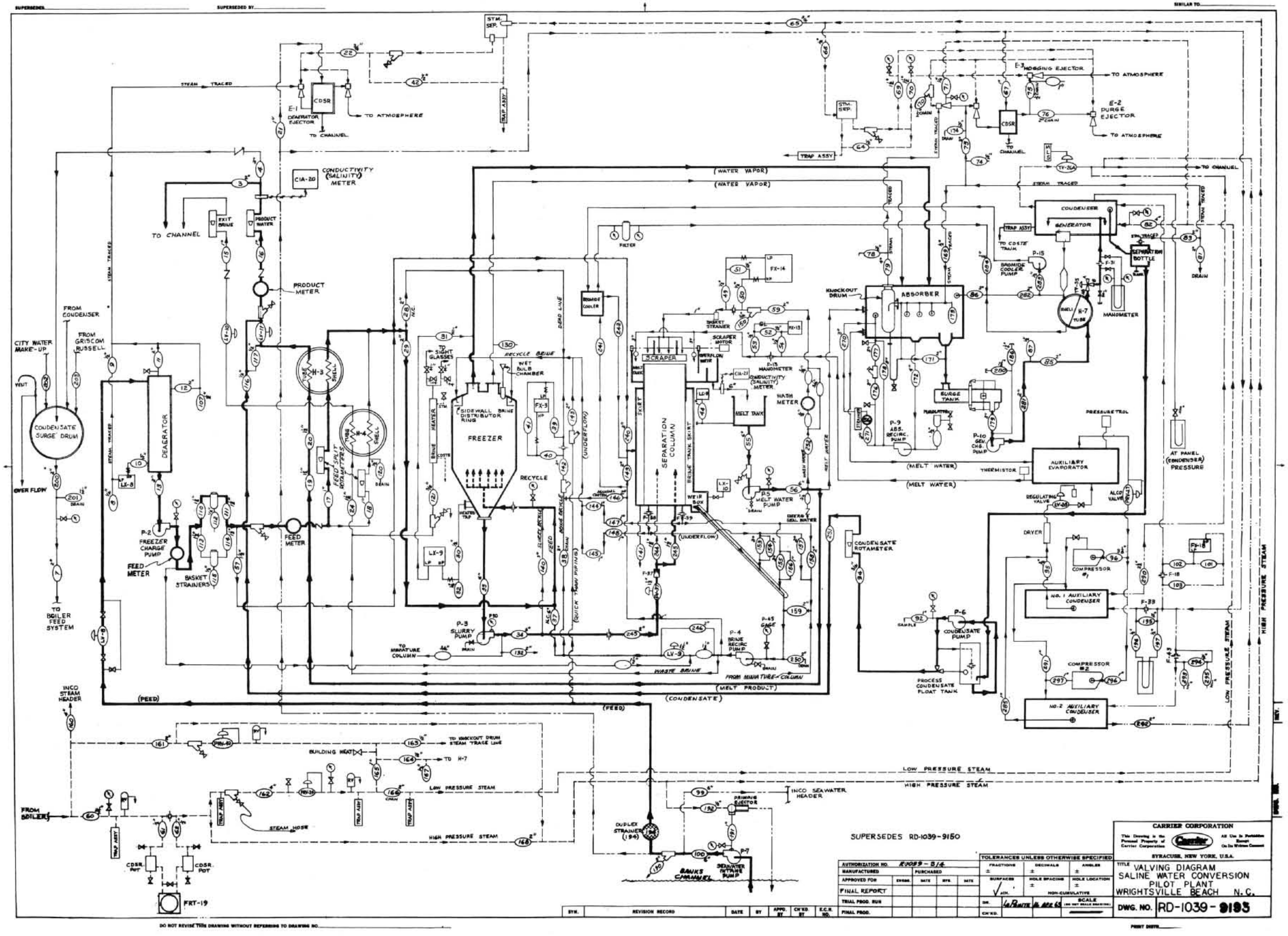
SYRACUSE, NEW YORK, U.S.A.

TITLE FLOW SCHEME
SALINE WATER CONVERSION PLANT
ABSORPTION-FREEZING PROCESS

DWG. NO. R1039-9177


CITY WATER MAKE-UP
FROM CONDENSER
FROM GRISCOM RUSSELL
CONDENSATE SURGE DRUM
OVER FLOW
TO BOILER FEED SYSTEM
DEAERATOR
FREEZER CHARGE PUMP
FEED METER
BASKET STRAINERS
DEARATOR EJECTOR
TO ATMOSPHERE
TO CHANNEL
CONDUCTIVITY (SALINITY) METER
EXIT BRINE
PRODUCT WATER
PRODUCT METER
FEED SPLIT ROTAMETERS
SIDEWALL BRINE DISTRIBUTOR RING
FREEZER
BRINE HEATER
SIGHT GLASSES
WET BULB CHAMBER
RECYCLE BRINE
RECYCLE
SLURRY PUMP
TO MINIATURE COLUMN
DEAD LINE
(WATER VAPOR)
(UNDERFLOW)
(QUICK THAW PIPING)
SEPARATION COLUMN
SCRAPER
SCRAPER MOTOR
BRINE TANK SKIRT
WEIR BOX
MELT TANK
WASH METER
MELT WATER PUMP
BRINE RECIRC PUMP
FROM MINIATURE COLUMN
WASTE BRINE
(MELT PRODUCT)
(CONDENSATE)
(FEED)
CONDENSATE ROTAMETER
CONDENSATE PUMP
PROCESS CONDENSATE FLOAT TANK
KNOCKOUT DRUM
ABSORBER
SURGE TANK
ABS. RECIRC. PUMP
GEN. CHG. PUMP
BROMIDE COOLER PUMP
CONDENSER
GENERATOR
SEPARATION BOTTLE
MANOMETER
AT PANEL (CONDENSER) PRESSURE
(MELT WATER)
THERMISTOR
AUXILIARY EVAPORATOR
PRESSURETROL
REGULATING VALVE
ALCO VALVE
DRYER
COMPRESSOR #1
NO. 1 AUXILIARY CONDENSER
COMPRESSOR #2
NO. 2 AUXILIARY CONDENSER
MOGGING EJECTOR
PURGE EJECTOR
TRAP ASSY
INCO STEAM HEADER
TO KNOCKOUT DRUM STEAM TRACE LINE
BUILDING HEAT
TO H-7
LOW PRESSURE STEAM
HIGH PRESSURE STEAM
FROM BOILERS
STEAM HOSE
CDSR. POT
FRT-19
INCO SEAWATER HEADER
PRIMING EJECTOR
DUPLEX STRAINER (194)
BANKS CHANNEL
SEAWATER INTAKE PUMP
SUPERSEDES RD-1039-9150
AUTHORIZATION NO. R-1089-B14
TOLERANCES UNLESS OTHERWISE SPECIFIED
CARRIER CORPORATION
SYRACUSE, NEW YORK, U.S.A.
TITLE VALVING DIAGRAM
SALINE WATER CONVERSION
PILOT PLANT
WRIGHTSVILLE BEACH N.C.
DWG. NO. RD-1039-9193
DO NOT REVISE THIS DRAWING WITHOUT REFERRING TO DRAWING NO.