Development of Propane Hydrate Desalting Process

United States Department of the Interior
Development of Propane Hydrate Desalting Process

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Contract No. 14-01-0001-341
14-01-0001-1238
As the Nation’s principal conservation agency, the Department of the Interior has basic responsibilities for water, fish, wildlife, mineral, land, park, and recreational resources. Indian Territorial affairs are other major concerns of America’s “Department of Natural Resources”.

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

**FOREWORD**

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

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

This report summarizes work under USDI-OWS Contracts 14-01-0001-341 and 14-01-0001-1238 carried out by Sweet Water Development Company on the propane hydrate saline water conversion process. Initial bench scale pilot plant work by Sweet Water Development Company on the process resulted in OWS participating in a second bench scale pilot plant. The data from this work served as a basis for the design of a 20,000 gpd pilot plant at Wrightsville Beach, North Carolina. This plant began operation in March of 1965.

The startup of the plant revealed problem areas in the crystallizer and separation-wash system. The hydrate being formed stuck to and built up on the walls of the crystallizer until flow through the vessel was blocked and plant operations became impossible. This problem was solved by eliminating the original pump-around system, installing internal agitation, and coating the inside wall surfaces. This proved to be the solution to the problem and resulted in controlled operation of the crystallizer.

The next problem area was found to be the cyclones which did not achieve hydrate-brine separation in the separation wash system. Plant testing failed to improve this separation and attention was turned to small scale testing of cyclones to analyze the problem. The small scale testing established that several cyclone configurations were capable of separating most of the brine from the hydrate, but none could remove enough brine to be suitable for Sweet Water's process. The addition of propane diluent in the feed did not prove beneficial to cyclone separation because of the poor wetting of hydrate by liquid propane.

It appears that much better separation of hydrate from brine could be achieved in a cyclone if a method of increasing the wetting angle between hydrate and liquid propane or a method for improving crystal habit were found. In lieu of changing the wetting angle or improving crystal habit, it seemed advisable to consider another wash separation system for the pilot plant. Several types of wash separation devices were considered, however, a pressurized wash column was chosen for installation.

During the wash column design and fabrication period, test work was continued on the crystallizer. Early tests showed the crystallizer could achieve 65% of design capacity. By increasing the power input to the mixers from an original value of 0.8 x 10^-3 hp/gal to 7 x 10^-3 hp/gal design production capacity of propane hydrate was achieved. During this period the melter volumetric heat transfer coefficient was determined for two internal configurations.
The work on the pressurized wash column indicates that it can be an effective wash separation device for the hydrate process. A product water concentration of 761 ppm of total dissolved solids was produced. This testing indicates additional screen area and equal distribution of this screen area would result in the wash column achieving its design function.

Economic analysis of the process appears in this report and projects a capital cost of $921,700 (turn key basis) for a one million gallons per day plant. The cost of water from this plant using electrical power at 7 mills per kwh and fifteen year straight line depreciation is 69 cents per thousand gallons of fresh water.

The plant estimate for the one million gallons per day plant considers the use of multiple units in parallel to attain capacity. Estimates of larger capacity plant appear to be a repetition of this parallelism and thus no significant reduction in either unit capital charge or product water cost is foreseen for plants of 5 and 10 millions of gallons per day.
I. Introduction

This report summarizes work carried out under USDI-OSW Contract 14-01-0001-341 and Contract 14-01-0001-1238 between the Office of Saline Water and the Sweet Water Development Company of Dallas, Texas, on a 20,000 gpd propane hydrate forming saline water conversion process pilot plant at Wrightsville Beach, North Carolina.

The Office of Saline Water, United States Department of the Interior, has as its objective the development of a method or methods to produce fresh water from sea water or saline waters at an economical cost.

A novel approach to solution of this problem involves the use of various organic and inorganic gaseous agents, which, when dissolved in saline waters at appropriate conditions of pressure and temperature, form solid hydrates - that is, solid crystalline substances containing the gaseous agent and pure water, coupled by weak molecular bonds, but containing none of the ions of the aqueous solution. These crystals can be washed free of adhering brine and melted to recover water and the gaseous agent. One attractive material for hydrate formation is propane. Propane is available commercially at an economical price, forms a hydrate that freezes with water over a range of temperatures from 32°F to 42.1°F, and contains roughly 17 molecules of combined water per molecule of propane. Propane is easily handled as a process material. The propane also acts as a medium to transfer energy in the crystal formation and melting steps of the process.

II. Process Background

Original investigations on this hydrate process were carried out by the inventor, Mr. V. C. Williams, in 1958 and 1959 in Puerto Rico. Additional studies led to the granting of a United States Patent (2,974,102) early in 1961. Shortly thereafter, also in early 1961, the Sweet Water Development Company of Dallas, Texas was formed to advance the development of the process by the construction and operation of a first small scale pilot plant in St. Louis, Missouri. Additional patents, assigned to Sweet Water Development Company, were applied for in February 1962. On May 1, 1962, a "first" contract was entered into between Sweet Water and OSW for the construction and operation of a second small scale plant.

The basic design data for Sweet Water's 20,000 gpd Wrightsville Beach pilot plant were obtained from the operation of these two small scale units in St. Louis, Missouri. The first unit was operated solely by Sweet Water. The second small scale unit represented the first OSW participation and was sponsored under contract 14-01-0001-259. This contract was completed in December 1962 by the submission of a report on the second generation bench scale operations and a preliminary design of a 20,000 gpd pilot plant.
III. Plant Design and Construction

Engineer-architect for the Wrightsville Beach 20,000 gpd pilot plant was The Lummus Company of New York, N. Y., using Sweet Water data from the small scale units. In addition to the process design of the plant itself, a number of support facilities had to be designed and constructed. The latter, for convenience designated as the "civil" phase, consisted of the craneway, extension of the existing OSW test station pipe rack, instrument and plant air lines, steam lines, and seawater supply lines. This phase of design, and its construction, was actually an addition to the OSW test station facility at Wrightsville Beach.

The design of the civil phase was completed on November 4, 1963. The design of the pilot plant was completed March 20, 1964. Specifications were issued for construction bids for each of the above phases on their respective completion dates.

The construction of the civil phase began March 23, 1964, with the successful bidder, Reagan Construction Company of Wilmington, N. C. undertaking the work. The civil construction phase was completed July 1, 1964.

The construction of the plant began July 16, 1964 and was performed by Sam P. Wallace & Company, Dallas, Texas. The plant construction was essentially completed by January 15, 1965, although the last piece of equipment, a temperature recorder, did not arrive for installation until well into April, after preliminary plant operations had begun in mid-March.

A photograph of the completed plant follows as Photo A.

IV. Plant Equipment Description

All the equipment used in the Wrightsville Beach pilot plant is standard off-the-shelf chemical process equipment. All gas and liquid streams in the plant which do not contain solids are measured by orifice meters. Streams which contain solid material are metered by magnetic flow meters. The output signal from all these instruments is read directly or transmitted pneumatically. Standard air motor control valves are used. All motors and electrical switches, junctions, etc. are explosion proof since the plant and area are classified Class I, Group D, Division II because of the propane. A description of other plant equipment as originally installed is shown in Table IA. The construction drawings for this equipment were transmitted to OSW upon plant completion. A list of equipment for the plant at the completion of contract -1238 appears as Table IB.

V. Description of Process Flow

A description of the process flow system appears in the Appendix. The process flow using the originally installed cyclone wash system and the process flow using the pressurized wash column are discussed. A process flow diagram is also shown for each of the separation wash systems.
PHOTOGRAPH A

SWEET WATER DEVELOPMENT COMPANY
SALINE WATER CONVERSION PILOT PLANT
WRIGHTSVILLE BEACH, N. C.

- 3 -
<table>
<thead>
<tr>
<th>Item No.</th>
<th>Name</th>
<th>Description</th>
</tr>
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<tbody>
<tr>
<td>FA 101</td>
<td>Primary knock-out pot</td>
<td>2½ ft Diameter x 5 ft tank</td>
</tr>
<tr>
<td>FA 102</td>
<td>Secondary knock-out pot</td>
<td>1½ ft Diameter x 5 ft tank</td>
</tr>
<tr>
<td>FA 103</td>
<td>Propane hold tank</td>
<td>3 ft Diameter x 9 ft tank</td>
</tr>
<tr>
<td>FA 106</td>
<td>Reject brine hold tank</td>
<td>3½ ft Diameter x 8 ft tank</td>
</tr>
<tr>
<td>FA 107</td>
<td>Propane flash drum</td>
<td>4½ ft Diameter x 11 ft tank</td>
</tr>
<tr>
<td>FA 108</td>
<td>Melter recycle hold tank</td>
<td>4 ft Diameter x 7 ft tank</td>
</tr>
<tr>
<td>FA 109</td>
<td>Propane Water separator</td>
<td>6½ ft Diameter x 8½ ft tank</td>
</tr>
<tr>
<td>FA 112</td>
<td>Slurry surge tank</td>
<td>2½ ft Diameter x 10 ft tank</td>
</tr>
<tr>
<td>FC 101 - 108</td>
<td>Cyclones</td>
<td>18 in Diameter</td>
</tr>
<tr>
<td>FC 110 - 111</td>
<td>Cyclones</td>
<td>21 in Diameter</td>
</tr>
<tr>
<td>FC 112</td>
<td>Cyclone</td>
<td>24 in Diameter</td>
</tr>
<tr>
<td>DC 101</td>
<td>Maker</td>
<td>5 ft Diameter x 30 ft tank, agitated</td>
</tr>
<tr>
<td>DC 102</td>
<td>Melter</td>
<td>2½ ft Diameter x 34 ft tank, disc and donut</td>
</tr>
<tr>
<td>GB 101</td>
<td>Primary Compressor</td>
<td>Varidrive reciprocating, non-lube</td>
</tr>
<tr>
<td>GB 102</td>
<td>Secondary Compressor</td>
<td>Varidrive reciprocating, non-lube</td>
</tr>
<tr>
<td>GA 101-123</td>
<td>Pumps</td>
<td>Standard centrifugal</td>
</tr>
<tr>
<td>EG 101</td>
<td>Deaerator</td>
<td>14 in Diameter x 20 ft Plastisol lined, packed</td>
</tr>
<tr>
<td>DA 101</td>
<td>Depropanizer</td>
<td>14 in Diameter x 20 ft Plastisol lined, packed</td>
</tr>
<tr>
<td>DA 102</td>
<td>Depropanizer</td>
<td>14 in Diameter x 20 ft Plastisol lined, packed</td>
</tr>
<tr>
<td>EA 101-102</td>
<td>Product water and reject brine-sea water heat exchangers</td>
<td>Shell and tube, fixed tube sheet</td>
</tr>
<tr>
<td>EA 104</td>
<td>Seawater chiller</td>
<td>Shell and tube, fixed tube sheet</td>
</tr>
<tr>
<td>EA 103</td>
<td>Propane Condenser</td>
<td>Shell and tube, fixed tube sheet</td>
</tr>
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</table>
**TABLE IB**

**FINAL PLANT EQUIPMENT DESCRIPTION**

**WASH COLUMN SYSTEM**

<table>
<thead>
<tr>
<th>Item No.</th>
<th>Name</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>FA 101</td>
<td>Primary knock-out pot</td>
<td>2½ ft Diameter x 5 ft vertical tank</td>
</tr>
<tr>
<td>FA 102</td>
<td>Secondary knock-out pot</td>
<td>1½ ft Diameter x 5 ft vertical tank</td>
</tr>
<tr>
<td>FA 103</td>
<td>Propane hold tank</td>
<td>3 ft Diameter x 9 ft horizontal tank</td>
</tr>
<tr>
<td>FA 108</td>
<td>Melter recycle hold tank</td>
<td>4 ft Diameter x 7 ft vertical tank</td>
</tr>
<tr>
<td>FA 109</td>
<td>Propane Water Separator</td>
<td>6½ ft Diameter x 8½ ft horizontal tank</td>
</tr>
<tr>
<td>FA 110</td>
<td>Propane Recovery Flash Tank</td>
<td>2½ ft Diameter x 3½ ft horizontal tank</td>
</tr>
<tr>
<td>FA 111</td>
<td>Propane Recovery Flash Tank</td>
<td>2½ ft Diameter x 3½ ft horizontal tank</td>
</tr>
<tr>
<td>FC 114</td>
<td>Reject Brine Separation Cyclone Maker</td>
<td>12 in Diameter</td>
</tr>
<tr>
<td>DC 101</td>
<td></td>
<td>5 ft Diameter x 30 ft agitated horizontal tank</td>
</tr>
<tr>
<td>DC 102</td>
<td>Melter</td>
<td>2½ ft Diameter x 34 ft packed tower</td>
</tr>
<tr>
<td>GB 101</td>
<td>Primary Compressor</td>
<td>Varidrive, reciprocating, non-lube</td>
</tr>
<tr>
<td>GB 102</td>
<td>Secondary Compressor</td>
<td>Varidrive, reciprocating, non-lube</td>
</tr>
<tr>
<td>GA 101, 103, 105, 107, 108, 122, 123</td>
<td>Pumps</td>
<td>Standard Centrifugal</td>
</tr>
<tr>
<td>EG 101</td>
<td>Deaerator</td>
<td>14 in Diameter x 20 ft Plastisol lined, packed</td>
</tr>
<tr>
<td>DA 101</td>
<td>Depropanizer</td>
<td>14 in Diameter x 20 ft Plastisol lined, packed</td>
</tr>
<tr>
<td>DA 102</td>
<td>Depropanizer</td>
<td>14 in Diameter x 20 ft Plastisol lined, packed</td>
</tr>
<tr>
<td>EA 101 - 102</td>
<td>Product water and reject brine-seawater heat exchangers</td>
<td>Shell and tube, fixed tube sheet</td>
</tr>
<tr>
<td>EA 103</td>
<td>Propane condenser</td>
<td>Shell and tube, fixed tube sheet</td>
</tr>
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VI. Process Description - General

The process consists of three basic unit operations as follows:

1) **Make** crystalline propane hydrate \( (3C_3H_8 \cdot 53H_2O) \) by contacting salt water and liquid propane.

2) **Separate and wash** the propane hydrate crystals.

3) **Melt** the washed propane hydrate by contact with gaseous propane to obtain fresh water and liquid propane.

Two of the operations above, making and melting hydrate, are reactions requiring a transfer of heat. An understanding of the factors governing the operation of the process, and especially the make and melt steps, results from a study of Figure I, the equilibrium phase diagram for a propane-propane hydrate-seawater salts brine system. The propane hydrate equilibrium (in pure water) data as determined by a residual treatment are represented by the relation:

\[
\log P = 27.50600 - \frac{12848.8}{T} + \Delta
\]

- \( P \) = psia
- \( T \) = °R = °F + 459.69
- \( \Delta \) = residual

from Miller and Strong's data (AGA Proc. 27, 80-94 (1945)) and Deaton and Frost's data (U. S. Bur. Mines Monograph #8, 1946), with the residuals correlated by group "lever arm" combination. Correlated data are

**TABLE II**

<table>
<thead>
<tr>
<th>( t ) °F</th>
<th>( P ) psia</th>
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<tr>
<td>32</td>
<td>23.75</td>
</tr>
<tr>
<td>33</td>
<td>26.64</td>
</tr>
<tr>
<td>34</td>
<td>29.99</td>
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<td>35</td>
<td>33.81</td>
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<td>37</td>
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<td>40</td>
<td>62.53</td>
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<td>41</td>
<td>70.38</td>
</tr>
<tr>
<td>42</td>
<td>79.14</td>
</tr>
<tr>
<td>42.1 (critical)</td>
<td>80.20</td>
</tr>
</tbody>
</table>

In the crystallizer with sea salts brine present, the freezing points (tabulated above) for the hydrate would be depressed because of the salts
Propane - Propane Hydrate - Seawater
Equilibrium Diagram
effect. A. J. Barduhn - Syracuse University Progress Report October 1 to December 31, 1960, "Demineralization of Sea Water by Hydrate Formation" - states that a useful generalization is that all hydrates appear to have about the same depression of the critical decomposition temperature per unit amount of salt. This depression lies between 1.0 and 1.1°F per weight percent NaCl up to 10% NaCl. The depression of the formation temperature of ice in salt water also lies in this range and varies from 1.00 to 1.08°F/‰ in the range 0 to 6% NaCl.

However, as shown by H. M. Hendrickson (Ref. Eng. 49, 3 - 37 (1958)) among others, the freezing point depression for sea salt brine is about 93.5% of that for NaCl brines at the same weight percent up to about 8 wt %. It is inferred from these references that the critical decomposition temperature of propane hydrate would be depressed about 0.935 x 1.1 or 1.03°F per weight percent sea salt brine linearly as the concentration goes from 0 to 8 wt % sea salt. The salt content is determined from the chlorinity by an empirical equation derived from data in "The Oceans": Sverdrup, Johnson, Fleming (Prentice-Hall, Publisher, 1942) instead of the old 1902 Copenhagen equation.

\[ S \frac{^\circ}{oo} = 1.8147 \ Cl \frac{^\circ}{oo} \]

where

\[ S \quad = \quad \text{Salinity ("sea salts")} \]
\[ Cl \quad = \quad \text{Chlorinity} \]
\[ \frac{^\circ}{oo} \quad = \quad \text{Parts per thousand - milles} \]

The important operating control in the crystallizer is the pressure in the vapor space, which sets the vaporization temperature of the propane liquid and thus establishes the driving force for hydrate formation. The pressure is carefully measured with test gauges and controlled by a pressure recorder controller (PRC) in the control room. The vapor pressure of propane is determined from Stearns and George data (IEC 35, 602 (1943)), using the residual analysis method also:

\[ \log P = 5.49009 - \frac{1798.529}{T} + \Delta \]

where

\[ P \quad = \quad \text{psia} \]
\[ T \quad = \quad \frac{^\circ}{R} = t ^\circ F + 459.69 \]
\[ \Delta \quad = \quad \text{residual} \]

By smoothing:

<table>
<thead>
<tr>
<th>t ℉</th>
<th>P</th>
<th>Δ</th>
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<tbody>
<tr>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>0</td>
<td>37.81</td>
<td>0.00000</td>
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<tr>
<td>10</td>
<td>45.85</td>
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<tr>
<td>40</td>
<td>77.80</td>
<td>0.00018</td>
</tr>
<tr>
<td>50</td>
<td>91.50</td>
<td>0.00000</td>
</tr>
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</table>

- 8 -
For the maker to form hydrate, a departure from equilibrium must occur. This departure, as previously mentioned, is controlled by careful regulation of the operating pressure of the maker. Temperature has proved to be a poor operating control because the liquid propane droplets, bulk brine, hydrate crystals, and brine at the crystal brine interface all must have different temperatures for crystallization to proceed. The mass in the maker is a uniform mixture, thus the temperature obtained by thermometry is some unknown combination of the actual temperatures of the various components existing in the maker. Barduhn, et al - Syracuse University Progress Report July 1, 1965, June 30, 1966 - "Studies on Ice Crystal Growth Rates and Hydrate Reaction Kinetics" - tabulate the many component temperatures which can exist in an operating hydrate crystallizer.

The degree of departure from equilibrium will obviously control the rate of hydrate formation in the maker and hydrate decomposition in the melter. A number of methods for describing this departure from equilibrium exists. de Graauw and Rutten, Technological University Delft, in their Interim Report of September 1965, prefer to choose a concentration difference converted from pressure differences by Henry's Law. A pressure difference driving force could easily be defined to describe the deviation from equilibrium for hydrate formation or equivalently a difference in fugacity, but since hydrate formation and hydrate decomposition are dependent on heat transfer steps, Sweet Water prefers to define an overall temperature difference obtained by means other than thermometry as the overall driving force. This driving force is defined from the equilibrium curve as the difference between the equilibrium temperature of propane corresponding to the operating pressure of the maker subtracted from the equilibrium temperature of hydrate at the operating pressure in a brine solution equal to that of the bulk brine solution. By way of illustration, if the operating pressure of the maker were 70 psia and the bulk brine concentration were 4.00%, the driving force from Figure I would be 37.25°F minus 33.75°F or 3.50°F.

VII. Plant Start-up and Modification

Plant operations began March 18, 1965. Operating technicians for plant operations were hired in the Wilmington, North Carolina, area and schooled in the operation of the process prior to plant start-up. Several mechanical problems were encountered on plant start-up. These were

1) The control valve and spray nozzle on the deaerator were undersized. The valve size was increased and the spray nozzle was removed.

2) The designed flow rate of liquid propane into the maker could not be achieved. A pump was installed in this line to provide available pressure to obtain the flow.
3) The vapor head in the maker did not have sufficient flow area and caused a pressure drop between the maker and primary compressor suction. The vapor head was revised to provide sufficient area.

4) The reject brine hold tank did not allow propane to settle out and separate from the discharge brine. The inlet nozzle was revised to move the turbulence caused by the feed stream away from the outlet reject brine nozzle.

5) Water tended to collect in low places in propane lines. When conditions were favorable, hydrate formed and plugged these lines. Drains were installed in these low places.

6) An undersized control valve on the vapor line from the chiller resulted in excessive pressure drop. This line was revised to include a by-pass for the valve.

7) Ten of the pumps in the plant had lantern rings of cast iron installed by the manufacturer instead of the specified teflon lantern rings. The result was corrosion of the cast iron and pump shaft seizure. The iron rings were chiseled out and replaced with the proper rings.

The most serious obstacle to plant operations was in the maker. The initial design of the maker involved the use of an elaborate pump around system for agitation. The fluid in the maker made three longitudinal passes inside the crystallizer. On plant start-up, as the maker began to form hydrate, the hydrate would build on the walls until flow was completely obstructed. The problem was diagnosed as being too little agitation. Attempts at plant operation ceased on May 29, 1965 for revisions to the maker. The troublesome pump around system was eliminated and six top entering 1 horsepower mixers were added to provide internal agitation. To prevent hydrate sticking to the walls, a chlorinated rubber base paint was used on the internal surfaces. This paint blistered and peeled after six months service and was later replaced with a phenolic epoxy coating which has since given good service. The plant was restarted in August, 1965, and the modification to the crystallizer proved successful in that it allowed the maker to produce hydrate for further plant testing. The coating greatly lowered the tendency of the hydrate to stick to the walls of the vessel.

During the period in which the maker problems were occurring, a check out of the other sections of the plant revealed that most were capable of performing their design function. The seawater feed section filtered and deaerated in excess of 35 gpm of sea water. Oxygen content of the water from the deaerator was 0.7 ppm. Heat exchangers EA 101 and EA 102 were able to handle in excess of this flow rate with a 2°F approach.
The product and reject brine stripping section proved capable of depropanizing the discharge streams from the plant to within the limits of specifications for product water when proper operation of the propane-water gravity separators was achieved. Inefficient separation of propane and water in the gravity separators results in entrained propane in the water to these depopropanizers, overloads the vacuum system, and decreases the depopropanization as well as resulting in high propane losses from the plant.

The compressor section and melter section have performed over the ranges of operation that have been imposed upon them. Some difficulty has been encountered with the vari-drive unit on the primary compressor. These problems have been mechanical and minor in nature. These problems include excessive belt wear, bearing problems, and adjustment problems on the pneumatically controlled unit.

VIII. **Cyclones - General Discussion**

Cyclones as separation-washing devices appear to have basic advantages over other type of solids separator-washer devices in other freezing processes. They have a high capacity, no moving parts, are inexpensive, adaptable to a wide range of separation problems and offer flexibility of operation.

Cyclones, and especially hydraulic or liquid cyclones, are widely used in the mining industries and from this industry most of the existing technology has evolved. It is interesting to note that the first hydraulic cyclone patent was issued in the United States in 1891 to E. Bretney, Pat. No. 453,105 but widespread usage of liquid cyclones did not occur until the late 1940s. Their design and usage remains primarily an art, or is based on operating experience for a given cyclone separation. However, recent books on cyclones, "Cyclones in Industry", edited by K. Rietema and C. G. Verves (Elsevier Publ Co, 1961) and "The Hydrocyclone" by D. Bradley (Pergamon Press, 1965), contribute valuable information to the field.

The design of a cyclone configuration involves optimum geometric symmetry and Rietema reports that Kelsall states the optimum geometric ratios are

\[
\frac{L_c}{D_c} = 5 \\
\frac{L_v}{D_c} = 0.4 \\
\frac{F_c}{D_c} = 0.28 \\
\frac{O_c}{D_c} = 0.34
\]
where

\[ D_C = \text{cyclone diameter} \]
\[ L_C = \text{overall cyclone length} \]
\[ I_V = \text{vortex finder length} \]
\[ F_C = \text{diameter of feed nozzle} \]
\[ O_C = \text{overflow diameter} \]

Rietema also states that the division of cyclone length into the cylinder and cone section is not critical although generally the cone section should be as long as possible and the cone angle always kept below 30°.

Other investigators have derived design criteria similar to those of Kelsall. D. A. Dahlstrom (personal communication, December 4, 1961, to V. C. Williams) sets forth design equations giving the design of the cylindrical section as having the relation

\[ \frac{O_C + 2F_C}{D_C} = 0.5 \text{ to } 0.6 \]

when

\[ O_C = 1.0 \text{ to } 1.5F_C \]

Dahlstrom states that the vortex finder length should be such that it extends below the bottom of the tangential inlet nozzle \( \frac{1}{2} \) to 1 in. The cylindrical section length extends 6 in. or \( D_C \), whichever is smaller, below the bottom of the vortex finder. Dahlstrom recommends a 10° to 15° included angle for the cone section. Cyclones designed according to Rietema and Kelsall are similar to those designed according to Dahlstrom, differing only in minor particulars. Thus, it is seen that cyclone design as well as their use is a general rather than a specific science.

Hydrate as pumped from the maker in the Sweet Water process is surrounded by and slurried by brine. The purpose of the cyclone system is to displace the brine from the hydrate slurry using liquid propane. Complete separation would be achieved in this manner in one cyclone if all the brine were prevented from getting into the propane hydrate-liquid propane slurry. Since some of the brine adheres to the hydrate and must be reduced in salt concentration before the hydrate is melted, a number of stages are required. In each stage, water of salinity lower than the adhering brine contacts the adhering brine on the hydrate, thus lowering its salinity by dilution. This process continues until the salinity of the melted hydrate and adhering brine is less than 500 ppm. A cyclone in the Sweet Water process receives as feed a mixture of brine, propane hydrate, and liquid propane. Complete separation is achieved when the lighter components, propane (sp gr: 0.53) and propane hydrate (sp gr: 0.88) are completely removed in a brine-free stream from the cyclone overflow. The heaviest component, brine (sp gr: 1.02), leaves the cyclone underflow, propane and propane hydrate free.
To define the separation resulting from the operation of a cyclone in the Sweet Water Hydrate Process the following factors are defined:

Solids Factor \( (F_S) \) = \( \frac{\text{hydrate reporting to the overflow}}{\text{hydrate in the feed to the cyclone}} \)

Propane Factor \( (F_P) \) = \( \frac{\text{liquid propane reporting to overflow}}{\text{liquid propane in the feed to the cyclone}} \)

Brine Factor \( (F_B) \) = \( \frac{\text{brine reporting to the underflow}}{\text{brine in the feed to the cyclone}} \)

Other definitions useful in defining cyclone operation are:

Overflow ratio \( (\frac{OF}{F}) \) = \( \frac{\text{overflow rate (vol.)}}{\text{total feed rate (vol.)}} \)

Diluent Ratio \( (D) \) = \( \frac{\text{diluent feed rate (vol.)}}{\text{total feed rate (vol.)}} \)

Slurry Concentration (Volumetric) \( (SC_V) \) = \( \frac{\text{hydrate volume}}{\text{hydrate volume + brine volume}} \)

Examination of the above definitions and factors as applied to the Sweet Water process reveals that perfect cyclone separation is achieved when \( F_S, F_B, \) and \( F_P \) are 1.00. Imperfect separation is reflected by a decrease in separation factors \( F_S, F_D, \) and \( F_P \).

For a cyclone operating with diluent ratio \( D \) and slurry concentration (volumetric) \( SC_V \), the limits of \( F_S, F_B, \) and \( F_P \) are shown in Figure II as a function of the overflow ratio, \( \frac{OF}{F} \). The solid lines show perfect cyclone separation. The dashed lines show the case where no separation occurs. For an operating cyclone, the separation factors must fall on or between the solid and dashed lines for a given separation factor. In the case of perfect separation, note that as the overflow rate is increased, the lightest component, propane, is first separated. As the overflow ratio continues to increase, the intermediate density component, propane hydrate, is separated. This continues until all the hydrate has been separated at point A, the only point at which perfect separation is achieved. Further increases in \( \frac{OF}{F} \) results in decreased \( F_B \).
IX. **Cyclone Testing - Plant**

The battery of cyclones originally installed in the plant consisted of one 24 in. diameter cyclone, two 21 in. cyclones, and eight 18 in. cyclones. After the initial mechanical and crystallization problems of the plant were solved, attention was focused on the testing of these units. Slurry flow rates from 50 to 180 gpm and propane diluent rates from 50 to 150 gpm were fed to these cyclones. The maximum feed rate tested was 250 gpm. Underflow rates from 20 gpm to 120 gpm were used in the testing. Marked separation was noted between the liquid propane and brine components of the feed. The separation of propane hydrate from the brine was so slight, however, that the difference in salt concentration of the overflow and underflow streams was, under the best condition of high feed rate and the optimum overflow ratio, only about 0.2% salts of the 3.5% salt concentration fed to the cyclone. During the course of this testing the physical variables of cyclones were changed as much as the design would allow. The underflow orifices were changed, the vortex finder diameter was increased, and the vortex finder length was increased and decreased, all with no change on the separating ability of the cyclones.

At this point in the program it was decided to try cyclones capable of developing greater centrifugal forces than the cyclones in the plant. A 14 in. cyclone was purchased from Heyl and Patterson, a commercial manufacturer of cyclones, and installed. Feed rates of brine and propane hydrate up to 180 gpm and propane diluent rates up to 150 gpm were tested on this cyclone. The maximum feed rate tested was 200 gpm. The results were approximately the same as those obtained from earlier plant tests. The cyclone was capable of achieving separation of liquid propane and brine, $F_B$ and $F_p$ approximately equal to 1.00, simultaneously; but the hydrate reported to the underflow instead of the overflow under these conditions, $F_S$ near 0.0. In tests without liquid propane in the feed, (i.e. only a hydrate-brine slurry as feed), the hydrate concentration in the overflow remained approximately equal to its concentration in the feed and the underflow.

The highest pressure drop, an indication of vortex strength, achieved in the plant throughout the testing of the 24 in., 21 in., 18 in., and 14 in. was 10 psi at 200 gpm feed using the 14 in. diameter cyclone. It was decided to try higher pressure drops by constructing and installing an 8 in. diameter cyclone. This cyclone was tested at maximum feed rate of 200 gpm and achieved pressure drops of 40 psi at this maximum rate. The results were the same as for the larger cyclones. The cyclone was capable of achieving separation of brine and liquid propane, but the propane hydrate still reported to the underflow when separation occurred between propane and brine. In tests in which no propane diluent was used, the hydrate concentration in the overflow and underflow stream remained approximately equal to that of the feed.
X. Cyclone Operational Problems

At this point in testing the performance of the pilot plant, the poor separation of the cyclones prevented the continuation of the planned plant testing program. The reason for the poor hydrate separation was thought to stem from either a change in the crystal habit or faulty cyclone design.

The hydrate crystal configuration in the St. Louis bench scale plants had been cubic. The hydrate crystals in the Wrightsville Beach plant appeared to be flat thin platelets.

The other possibility, faulty cyclone design, led to the intensive program of cyclone testing described in the following sections. Small scale testing appeared to be more expedient and to have more inherent flexibility than plant testing. To accomplish small scale cyclone testing a test panel containing a cyclone, metering and sampling apparatus was constructed and piped to operate on a small side stream from the plant. This testing focused attention on the small scale cyclone testing described below.

XI. Small Scale Cyclone Testing

Five different basic cyclone configurations and modification of the basic configuration were tested. The cyclones are listed in Table III. The drawings of these cyclones appear in the appendix, their drawing number being indicated in the table and show the dimension of the cyclone tested. A number of runs with a $2\frac{1}{2}$ in. diameter cyclone of standard configuration was made. When the results of these runs were similar to those obtained in the plant cyclones, a number of modifications to cyclone design was made.

The first modification of the $2\frac{1}{2}$ in. diameter cyclone was to install in the cone of a cyclone a porous inverted cone or porous inverted truncated cone. The idea of the insert was to force hydrate into the overflow without disrupting cyclone spiral flow patterns by preventing the hydrate from moving through the porous insert into the underflow. Ten such inserts were tested, a typical example of which is shown in Dwg. 1040. The materials used to form the cones was 30 mesh nylon cloth, 16 mesh stainless steel wire cloth, 40 mesh nylon screen, and etched stainless steel having 150 micron round holes on 0.0155 in. centers. The base diameter of these cone inserts varied from 2 in. to $2\frac{1}{2}$ in. The $2\frac{1}{2}$ in. diameter cone inserts forced all the cyclone underflow to pass through them. The cyclone was in effect working like a combination cyclone and filter. The smaller diameter cones were supported by 1/8 in. metal spiders to prevent disrupting cyclone flow as much as possible. Feed rates ranging from 2 to 6 gpm were tested. The cyclones failed in every case to discharge on overflow stream.
<table>
<thead>
<tr>
<th>Cyclone Name</th>
<th>Cyclone Body Diameter</th>
<th>Modifications Tested</th>
<th>Dwg. No.</th>
<th>Page No.</th>
</tr>
</thead>
<tbody>
<tr>
<td>2½ in.</td>
<td>2½ in.</td>
<td>1) Standard configuration</td>
<td>1000</td>
<td>72</td>
</tr>
<tr>
<td></td>
<td></td>
<td>2) Cyclone inserts</td>
<td>1040</td>
<td>73</td>
</tr>
<tr>
<td></td>
<td></td>
<td>3) Split nozzle inlet</td>
<td>1050</td>
<td>74</td>
</tr>
<tr>
<td>Doxie (R)</td>
<td>10 mm</td>
<td>1) Standard configuration</td>
<td>3000</td>
<td>75</td>
</tr>
<tr>
<td>Unidirectional Flow</td>
<td>½ in.</td>
<td>1) Body length of 7 in.</td>
<td>4000</td>
<td>76</td>
</tr>
<tr>
<td></td>
<td></td>
<td>2) Body length of 5 in.</td>
<td>4010</td>
<td>77</td>
</tr>
<tr>
<td>van Tongeren</td>
<td>3/4 in.</td>
<td>1) Standard configuration</td>
<td>5000</td>
<td>78</td>
</tr>
<tr>
<td></td>
<td></td>
<td>2) Various spiral inlets</td>
<td>5010</td>
<td>79</td>
</tr>
<tr>
<td>Plexiglas</td>
<td>1 in.</td>
<td>1) With and without vortex finder</td>
<td>6000</td>
<td>80</td>
</tr>
</tbody>
</table>
having a hydrate concentration sufficient to reduce its melted salinity more than 0.3% when fed with a slurry having a melted salinity of 3.5%.

The next test using the 2½ in. cyclone involved a change in the method of feeding the cyclone. On the chance that the poor separation was resulting from emulsification of propane and water droplets being mixed in feed tees before the cyclone, it was reasoned that by feeding the brine-hydrate slurry and the liquid propane streams into the cyclone unmixed, separation would be improved. Drawing 1050-A shows the conventional method used in feeding diluent and slurry. Drawing 1050-B and Drawing 1050-C show the two variations tested. The feed nozzle consisted of a pipe that had a divider plate soldered longitudinally to keep the diluent and slurry segregated before they entered the cyclone. The configuration shown in C was expected to give the best results since the brine on entering the cyclone and being subjected to the centrifugal forces would move outward, forcing its way through the propane diluent, leaving the hydrate near the center of the cyclone where it could be taken out with the propane in the overflow. This obviously was not the problem since separation no better than in previous tests was achieved.

A cyclone was constructed of plexiglas, see Drawing 6000, and tested in order to view the flow patterns existing inside a cyclone being fed liquid propane, propane hydrate, and brine. The cyclone was operated at feed rates up to 10 gpm. It was impossible to observe a sharp interface of any of the three feed components inside the cyclone. The test did reveal that the middle of the cyclone contained a propane gas core. The feed pressure to the cyclone was always greater than 120 psig during the testing. The propane being fed to the cyclone had a temperature range from 38-40°F. The equilibrium pressures for propane at 38-41°F from Figure 1 is 62 to 64 psig. In all cases the propane gas core was present, therefore a pressure gradient as great as 20 to 60 psi existed from the feed point to the center of the cyclone near the vortex, although the pressure drop across the entire cyclone varied from 8 to 60 psi. This pressure gradient from the slower moving feed to the fast moving vortex, followed by a partial recovery of pressure in the exit streams, can be attributed to the pressure-velocity relationship for a fluid stream. Photograph B, following, shows the gas core existing in the 1 in. cyclone operating at a feed rate of 6 gpm with and without a vortex finder. The 1 in. cyclone did little or no separation without a vortex finder. When a vortex finder was installed the cyclone showed an ability to achieve separation factors for brine and propane approaching 1.00. The solids separation factor was poor for the cyclone and represented little improvement over completely inefficient separation in the range of feed rate tested (2 to 9 gpm).
PHOTOGRAPH B
INTERNAL FLOW PATTERNS
OF ONE INCH PLEXIGLAS CYCLONE

Without Vortex Finder

With Vortex Finder
XII. Small Scale Cyclone Testing - Successful Apparatus

A number of cyclones which were successful in achieving separation of liquid propane, brine, and propane hydrate were tested. These cyclones include a Doxie (R) cyclone, a unidirectional flow cyclone, and a van Tongeren type cyclone.

The Doxie (R) cyclone, shown in Drawing 3000, is manufactured by the Dorr-Oliver Company, and consists of six parallel 10 mm cyclones. The unit is an off-the-shelf item, and as such was designed to operate up to 3,000 psi, although this high pressure was not a requirement for the Sweet Water application. From one to all six of the parallel 10 mm cyclones in the unit can be operated at the same time by plugging off the cyclones to be taken out of service.

The unidirectional flow or UF cyclone design was based on the premise that better separation could be achieved in a cyclone with less energy requirement if the cyclone overflow were not forced to undergo a reversal of flow before leaving the cyclone. This cyclone is shown in Drawing 4000. In operation the cyclone received the slurry feed at its upper end through the tangential inlet. This inlet imparts spin necessary for cyclone separation, moving the heavier components to the outside of the spiral flow pattern while the lighter components are concentrated in or near the center of the cyclone. The spiral flow pattern moves down the cylindrical tube, the heavy phase, brine in this case, is withdrawn from the bottom tangential outlet. The lighter phase, after concentrating in the center of the cyclone is withdrawn without undergoing any change in flow direction from the center by the tube inserted axially into the bottom of the cyclone. This tube corresponds to the vortex finder in a standard cyclone. The tube shown in the drawing inserted axially from the top of the cyclone is used to feed liquid propane to the cyclone at the center of the spiral flow pattern in the cyclone, the point at which it was desired to concentrate the liquid propane.

A second UF cyclone, UF II, was tested to find out the effect of shorter body length. This cyclone is shown in Drawing 4010. Time did not permit further testing of variations in cyclone configuration for optimization of the UF cyclone design.

The van Tongeren type cyclone is similar to a UF cyclone in that the overflow does not undergo a reversal in direction of flow. One difference between the cyclones is in the method of imparting spin to the motive fluid. The van Tongeren type cyclone uses a spiral insert to start the fluid spinning. Another difference is that in the van Tongeren type cyclone the spinning fluid passes through a convergent-divergent section in which the spin is accelerated before reaching the discharge nozzles. A drawing of the van Tongeren cyclone tested is shown in Drawing 5000. This drawing shows the original spin producing device installed, a 3 in. piece of a 3/4 in. boring bit. Drawing 5010 shows two other spin producing devices
which were tested, produced no apparent increase in separating ability, and caused higher cyclone pressure drop than the original boring bit.

The original testing of the Doxie (R) cyclone was on a test panel to which were added sample devices, meters, pressure gauges, and other apparatus as needed. This panel was designed to accommodate two cyclones which could be operated either in parallel or series. Metering of dilute slurries and propane was accomplished with rotameters. Thick slurries gave metering problems. Attempts were made to measure flows with a nutating disc meter but were unsuccessful. The thick slurry tended to plug, causing variations in flows and pressures. It was necessary to obtain the flow rates for streams of thick slurries (over 15% solids) by difference using a material balance on the system.

The hydrate concentration in a given stream was calculated from the salinity differences of the bulk brine and melted slurry in the stream. The method of calculation follows.

**Solids Concentration Calculation - Salinity Method**

\[
S_F = \text{filtrate salinity, or the salinity of the aqueous liquid phase in the slurry}
\]

\[
S_H = \text{salinity of hydrate} = 0
\]

\[
S_o = \text{salinity of aqueous phase in overflow settling bomb (melted)}
\]

\[
S_u = \text{salinity of aqueous phase in underflow settling bomb (melted)}
\]

\[
S_f = \text{salinity of aqueous phase in feed settling bomb (melted)}
\]

\[
X = \text{wt fraction hydrate in slurry}
\]

\[
SC = \text{solids concentration} = 100X
\]

In the feed:

\[
S_F (1 - X) + S_H X = (1) S_f
\]

\[
S_F - S_F X = S_f
\]

\[
- S_F + S_F X = -S_f
\]

\[
S_F X = S_F - S_f
\]

\[
X = \frac{S_F - S_f}{S_F}
\]

\[
SC \text{ of Feed} = 100 \left( \frac{S_F - S_f}{S_F} \right)
\]

and similarly
Solids Concentration Calculation - Salinity Method
Cont'd

\[ \text{SC of overflow} = 100 \frac{(S_F - S_0)}{S_F} \]

and

\[ \text{SC of underflow} = 100 \frac{(S_F - S_u)}{S_F} \]

The accuracy of the method depends heavily on obtaining a hydrate free brine sample. Two methods were used to get brine samples. One was to filter the hydrate from the slurry and the other was to use a cyclone capable of separating a hydrate free brine sample. The latter method was found to be easier to run and to be more accurate.

XIII. Small Scale Cyclone Testing - Results

The results of the testing of the 10 mm conventional design cyclone, UF I, UF II, and the van Tongeren cyclones are shown in Figures III, IV, and V. As shown in Figure III the flow capacity of the tested cyclones varied widely at a given pressure drop. For a given pressure drop the cyclone having the largest flow rate was the van Tongeren followed by the UF II, UF I, and a single Doxie (R) 10 mm cyclone. The cyclones did not vary a great deal in their ability to separate brine into the underflow as shown by their brine separation factors, \( F_B \), in Figure IV. The cyclones were very similar in their ability to separate liquid propane and propane separation factors, \( F_P \), of 1.00 were easily attained in all of the cyclones. The cyclones did show a marked difference with respect to the solids separation factors, \( F_S \), as shown in Figure V. The best solids separation for a given overflow ratio was achieved by the Doxie (R) 10 MM cyclone followed by the UF I, UF II, and the van Tongeren.

The effect of increased pressure drop on \( F_S \) and \( F_B \) (obtained by increased feed rate) at a given overflow ratio is shown in Figure VI. This particular curve is for the UF II and differs from the other cyclones only in its degree of displacement from 1.00 values of \( F_S \) and \( F_B \). This curve illustrates that increasing pressure drop for a given cyclone is beneficial for \( F_S \) and has a small favorable effect on \( F_B \).

Figures VII, VIII, and IX illustrate the effect on cyclone separation when liquid propane is fed to the cyclone. Figure VII shows the dependence of separation factors \( F_S \) and \( F_B \) for the UF II cyclone on overflow ratio without the presence of propane diluent. Figure VIII shows the displacement of these curves resulting from liquid propane, diluent ratio of 0.15, being fed to the cyclone. It will be noted that \( F_B \) is improved but \( F_S \) is lowered.
Figure III. Pressure Drop vs. Feed Rate for Test Cyclones - With No Diluent

Note: Pressure drop is numerical average of feed-to-overflow and feed-to-underflow pressure drops.
Figure IV.
Effect of Overflow Ratio on Hydrate Recovery
(PH) for Test Cyclones at 22 psi Pressure
Graph No. Digital

- 25 -
Effect of Pressure Drop on Separation Factors

HE-51 Cyclone Without Draining
(Overflow Ratio 0.4)

Figure VI
Effect of Overflow Ratio on Separation Factor

U.E.N. Cycles Without Diluent
(Slurry Feed Rate 2.43 gph)

Figure VII
Effect of Overflow Ratio on Separation Factors

UF: II Cyclone With Diluent

Feed Rate: Slurry 24 kgpm
Diluent: Air 8 kgpm
P = 1 at All Rates

Figure VII
Effect of Pressure Drop on Separation Efficiency

UF-II Cyclone: Water Dispersed

Diluent Ratio: 0.15
Overflow Ratio: 0.36
FP = 0.30 All Runs

Figure II
for a given overflow rate. This is expected, since with propane present a
greater overflow ratio would be required to get the same \( F_g \), but the same
maximum \( F_g \) should be eventually attained. This was not the case. A
maximum \( F_g \) of 0.97 is possible without diluent, while a maximum \( F_g \) of 0.82
is obtained with diluent present. This is an indication that the hydrate
is behaving as an entirely separate fluid in the cyclone.

Figure IX, as compared to Figure VI, shows the effect of propane diluent on
\( F_g \) and \( F_B \) with increasing pressure drop. The brine separation factor \( F_B \) has
been improved, but the solids separation, \( F_S \), has suffered.

In all cases where solids separation was being achieved in the cyclones with
good brine rejection and propane separation, the cyclone overflow would
consist of a thick viscous aqueous phase in liquid propane. The two phases
showed no tendency to intermix and increasing the amount of propane acted
only to make the bulk fluid less viscous. The thick aqueous phase con-
sisted of a maximum of 35\% solids in brine. This indicates that the
brine was not displaced from the compacted hydrate slurry by the propane.
The fact that the hydrate could not be dispersed in liquid propane indicates
that propane does not wet the surface of propane hydrate.

The performance of cyclones in Sweet Water's process would be greatly
improved if propane, instead of brine, could be made to wet preferentially
the hydrate surface. Twenty-six surfactants ranging from hydrocarbon-sea-
water emulsion breakers to hydrocarbon-seawater emulsion stabilizers were
tested in an attempt to accomplish this purpose. Twenty-four of the
surfactants were lab samples from the Tretolite Company, St. Louis, Mo.,
one was Sulfanole No. 9, made by Sun Chemical Corp., Wood River Junction,
R. I., and one was Calgon, made by the Calgon Corp., Pittsburg, Pa. In
the tests, the cyclone slurry feed rate, diluent feed, overflow and under-
flow rates were held constant. Any benefit resulting from surfactant
addition would be reflected by an increase solids concentration in the
cyclone overflow.

Each surfactant tested was injected in the cyclone feed stream at a rate
giving a surfactant concentration of 10 ppm in the feed stream. The
results indicated that none of the surfactants tested caused the propane
to wet preferentially the hydrate surface. Separation of hydrate from
the surrounding brine was no better than when no surfactant was used.
In fact, some of the surfactants lowered the solids concentration in
the cyclone overflow.

Doxie (R) 10 mm cyclones were tested in series, the overflow from the first
cyclone being fed to the second cyclone. The testing was not successful in
getting overall cyclone separation factors greater in two stages of separa-
tion than in one cyclone. The thick viscous nature of the overflow from
the first cyclone was a problem to operation and metering on the second
cyclone. Dilution of the feed to the second cyclone with liquid propane
resulted in easing some of the operational problems but resulted in an
overall separation of hydrate in two stages which did not exceed that from
one.
XIV. Alternative Wash Separation Device Studies

The difficulty in the use of cyclones in the hydrate process remains achieving displacement of interstitial brine with propane or improving crystal habit or size to achieve better separation of solid and aqueous phases. In view of the failure to improve displacement of brine or to improve crystal habit, it was necessary to consider alternative methods of hydrate separation and washing. Four methods of wash separation were considered. These were 1) a conventional gravity drain wash column, 2) a pressurized wash column, 3) a wash column fed with a heavy hydrate-brine slurry from a cyclone overflow stream and 4) a screening jig of the "Sweco Center Discharge Separator" type. None of the proposed wash separators were judged as technically impractical. The gravity drain wash column design required such a large wash column that it was dropped from consideration (see wash column theory discussion). The cyclone and wash column combination as well as the "Sweco Center Discharge Separator" appeared to require longer periods for development. The choice was thereby narrowed to a pressurized wash column. Design of such a device began immediately after receiving OSW approval to proceed.

XV. Pilot Plant Maker Testing

During the design period for the pressurized wash column the crystallization rates of propane hydrate were studied in the maker. A number of plant runs were made in the course of this study. Table IV summarizes these runs. It will be noted that the independent variables studied were primarily driving force and pump out rate.

The initial testing as shown in Run A-7 indicated the maximum production capacity of the maker was 76 lbs of hydrate as water per minute at 130 gpm slurry pump out rate. For 20,000 gpd of hydrate as water from the maker a production rate of approximately 116 lbs of hydrate as water per minute is required.

Pangborn and Barduhn of Syracuse University Research Institute reported that the production rate of methyl bromide hydrate was dependent upon hydraulic power input to the stirrer up to a value of 1x10^-2 hp/gal stirred. Pangborn and Barduhn's data indicated that Sweet Water's maker agitation system was underpowered. It therefore appeared logical to increase the power input to the maker. To test this hypothesis, the blade size on the six turbine mixers on the maker were increased from their original height of 2-19/31 in. to 3-3/16 in. The turbine blade assembly diameter was also increased from 13 to 14 1/2 in. The power input with 1750 gallons operating volume was thereby increased from its original 0.8x10^-3 hp/gal to 1.6x10^-3 hp/gal. The effect of this increase in power input was tested in Run C-1, and resulted in raising the maximum attainable production rate of the maker from 76 to 86 lb/min of hydrate as water.
<table>
<thead>
<tr>
<th>Run #</th>
<th>Motor Size (hp)</th>
<th>Turbine Diameter (in.)</th>
<th>Maker Volume (gal)</th>
<th>Pump Out Rate (gpm)</th>
<th>Run Variable</th>
<th>Run Complete</th>
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<tbody>
<tr>
<td>A-1</td>
<td>1</td>
<td>13</td>
<td>1750</td>
<td>50</td>
<td>ΔT, Pump Out</td>
<td>Yes</td>
</tr>
<tr>
<td>A-2</td>
<td>1</td>
<td>13</td>
<td>1750</td>
<td>Varied</td>
<td>ΔT, Feed Temp</td>
<td>Yes</td>
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</tr>
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<td>Yes</td>
</tr>
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<td>A-6</td>
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<td>A-7</td>
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<td>130</td>
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<td>Yes</td>
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<tr>
<td>B-1</td>
<td>1</td>
<td>13</td>
<td>1750</td>
<td>Varied</td>
<td>ΔT, Pump Out</td>
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<tr>
<td>C-1</td>
<td>1</td>
<td>14(\frac{1}{2})</td>
<td>1750</td>
<td>130</td>
<td>ΔT</td>
<td>Yes</td>
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<tr>
<td>C-2</td>
<td>1</td>
<td>14(\frac{1}{2})</td>
<td>1750</td>
<td>Varied</td>
<td>Pump Out</td>
<td>No</td>
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<td>1</td>
<td>14(\frac{1}{2})</td>
<td>1700</td>
<td>Varied</td>
<td>Pump Out</td>
<td>Yes</td>
</tr>
<tr>
<td>C-4</td>
<td>1</td>
<td>14(\frac{1}{2})</td>
<td>1750</td>
<td>130</td>
<td>ΔT</td>
<td>Yes</td>
</tr>
<tr>
<td>D-1</td>
<td>3</td>
<td>13</td>
<td>1750</td>
<td>130</td>
<td>ΔT, Filtered Seawater</td>
<td>Yes</td>
</tr>
<tr>
<td>D-2</td>
<td>3</td>
<td>13</td>
<td>1750</td>
<td>130</td>
<td>ΔT</td>
<td>Yes</td>
</tr>
<tr>
<td>D-3</td>
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<td>13</td>
<td>1750</td>
<td>130</td>
<td>ΔT</td>
<td>Yes</td>
</tr>
<tr>
<td>E-1</td>
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<td>1400</td>
<td>130</td>
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</tr>
<tr>
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<td>2200</td>
<td>130</td>
<td>ΔT</td>
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</tr>
<tr>
<td>F-2</td>
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<td>13(\frac{1}{2})</td>
<td>2100</td>
<td>Varied</td>
<td>Pump Out</td>
<td>Yes</td>
</tr>
</tbody>
</table>
The success experienced in increasing production rate by increasing the stirring power led to tests in which the power input to the maker was further increased. The six 1 hp mixer motors originally installed with the 13 in. diameter turbines had been operated overloaded with the 14½ in. diameter turbines in order to achieve the increase from $0.8 \times 10^{-3}$ to $1.6 \times 10^{-3}$ hp/gal. It was necessary to purchase new motors for further increases in power input. The largest motors made to fit the existing mixer assemblies was 3 hp. By the installation of 3 hp motors and using the old 13 in. diameter turbine assemblies (blade height 2-19/31 in.) a power input of $7 \times 10^{-3}$ hp/gal stirred was obtained. With this power input a production rate of 120 lbs/min of hydrate as water was obtained at a driving force of 3.2°F (Run D-1).

The favorable results of increasing the power input to $7 \times 10^{-3}$ hp/gal stirred made further increases in power input a logical extension of the resting. Since 3 hp motors were as large as would fit the mixers, the only way this could be achieved was by overloading these motors. This was accomplished by increasing the turbine assembly diameter from 13 to 13½ in. Attempts to operate with this power input and the same maker volume as in previous tests (1750 gal of stirred liquid) were not successful. The additional mixing caused splashing and excess mixer cavitation. The operating level and as a result the volume in the maker had to be raised. The net result of increasing both power input and volume turned out to be about a stand off as far as mixing power input was concerned. The power input with the larger blades and higher maker operating volume was $6.0 \times 10^{-3}$ hp/gal stirred for Run F-1 as compared with the $7 \times 10^{-3}$ hp/gal stirred in Run D-1.

The experimental data for Run D-1 are shown as Figure X and represent the data taken for a typical run. Each of the data circles represents the average of 8 hours operation under smooth plant operating conditions. The straight line portion of the curves are straight line least squares fits. Figure XI shows the effect of increasing the power input to the maker. The curves were obtained from data taken in Runs A-3, C-1, D-1, and F-1. It is believed the drop off in production rate in runs A-3 and C-1 as the driving force increased in the runs results from the relatively low powered mixers and their inability to agitate the resulting thicker slurry in the maker.

There were two sources of error during all of the testing. The biggest problem was in obtaining representative samples. Since there was no installed hydrate separation system in the plant, the amount of hydrate present in the slurry from the maker was calculated from test samples. This hydrate production computation was based on a salinity balance of a sample of the melted slurry and a sample of the bulk brine from the slurry (see pages 21 and 22). The precision of the method is hindered by the difficulty in obtaining a representative bulk brine sample. This sample must be hydrate free for a precise result from the calculations.
Figure X  Pulp Dia., Mass Production Rate vs AFA  
Function of Driving Force

Average Mass Volume = 750 g/l
Pulp Dia, Area = 2.5 cm
Through Put  = 12 inches
Turbine Speed  = 300 rpm
Heater Driven Input  = 7 x 10^-5 hp/gal/min

Temperature vs. Driving Force (kcal)

- 34 -
The separation of this bulk brine sample was obtained by feeding slurry through two Doxie type (1.0 cm) cyclones. The underflow from the first cyclone was the feed to the second cyclone. The second cyclone was operated with a high overflow to underflow ratio and the underflow from the cyclone was used as the bulk brine sample.

The other source of error lay in determining the temperature driving force for crystallization (ΔT). The driving force was derived from the equilibrium diagram for propane-propane hydrate-seawater mixtures (see page 7). This curve is for pure propane. The pilot plant used the purest commercially available propane, HD-5, as specified by Natural Gas Processors Association publication 2140-62. These specifications call for a minimum of 90 liquid volume percent propane. Although the delivered product was usually 90 to 95% propane it was never more than 98% propane. Other components with different vapor pressures tend to shift the equilibrium boiling point, and as the period of operation with a given charge of propane lengthened, the equilibrium shifted more. This inherent error is believed a larger factor in the tests in which the driving force is small, and probably explains some of the data scatter and particularly the differences between the production rates of various runs at the smaller driving forces of less than 1.0°F.

The size of the hydrate crystals was determined from permeability data using the Carmen-Kozeny method of particle size determination by pressure drop across a bed of known porosity. The effective particle diameter determined by this method was found to vary from 20 to 40 microns. Particle sizes were generally found to be inversely proportioned to the driving force for crystallization.

The effective particle size determined from permeability data on ice slurries gives values which are usually 2 to 3 times smaller than the actual particle size as determined by photographic means. Deviation from spherical shape accounts for this difference. Using this rule of thumb, the permeability tests are in agreement with particle size determined from photographs of propane hydrate crystals. These pictures, taken with a microscope and Polaroid camera, show particles ranging from 50 to 200 microns. In general, the shape of the crystals is flat platelets although some cubic crystals were observed. Photographs of propane hydrate crystals appear in the Appendix.

Several maker runs were made in which the pump out rate from the maker was the controlled variable. Because of control valve size, the lowest pump out rate which could be controlled from the maker was 50 gpm. The usual procedure was to begin at 50 gpm and increase the pump out rate slowly over a period of time. An example of data from one of these runs, Run D-3, appears in Figure XII. As in the driving force runs, each data point represents the average data for 8 hours of smooth plant operations. The straight line segment of the data is a least squares straight line fit. The curves for all of the variable pump out runs appear in Figure XIII. The driving force used in all of the runs except one was approximately 1.8°F (actually 1.7°F - 2.0°F), the exception being Run D-3.
Figure XII: Run D-3: Maker Production as a Function of Pump Out Rate.

Maker Volume = 1600 gal.
Driving Force = 2.4 - 2.5°F.
Two runs were made in which the amount of excess propane (unreacted or unvaporized propane) in the discharge from the maker was varied. The range of excess propane tested was from 2% to 20% by volume. The data showed that excess propane had little or no effect on maker production rate.

XVI. A. Pilot Plant Melter Testing

The operating efficiency of the melter in the propane hydrate process is measured by its heat transfer characteristics and the pressure drop across the unit. The melter is a direct contact heat exchanger and for effective heat transfer must intimately contact a hydrate slurry and propane gas. Two types of melter internals designed to promote intimate slurry-gas contact were tested during the program. The first of these was the disc and donut system originally installed. The disc and donut system consisted of 15 discs 24 in. diameter and 14 donuts with the "hole" 18 in. in diameter, the outside diameter of the donuts being the same as the inside diameter of the melter, 29-3/8 in. These discs and donuts were spaced 1 ft apart and the arrangement was 28 ft in length.

The other type of melter internals tested was Flocor packing manufactured by the Ethyl Corporation. Flocor consists of a combination of PVC flat sheets and thermoformed PVC sheets specially designed and sinusoidally corrugated in two directions. This packing has a 98% void factor and was selected to achieve intimate gas-slurry contact with a low pressure drop across the melter. The Flocor was installed in one ten ft and two eight ft sections with individual support plates for each section. A slurry distribution plate was designed and installed over the top of the packing.

The tests on the melter for both types of internals were made by feeding a given liquid slurry rate of known hydrate composition to the melter and controlling the driving force for heat transfer to achieve something less than complete melting of the hydrate. By varying the slurry feed rate to the melter different heat loads were attained.

The plant as constructed had two pumps that recycled water through the melter for dilution of incoming slurry. It was believed that this recirculation system would improve heat transfer in the melter. The effect of this recirculation system was investigated during the course of the work on the melter.

In scaling up melter design it is convenient to use a volumetric heat transfer coefficient, and this was used for evaluation of the performance of the two packings. It should be emphasized that while this coefficient can be used to scale up the Flocor it cannot be used to scale up the disc and donut internal arrangement. The volumetric heat transfer coefficient for the disc and donut is useful, however, for comparison. Figure XIV shows the volumetric heat transfer coefficients of the disc and donuts as well as the Flocor and the effect on this coefficient when the recirculation water is added to the feed so that the liquid feed rate is increased.
Figure XIX: Effect of total liquid feed to melter on melter volumetric heat transfer coefficient.

Comparison of floor packing with disc & donut arrangement.

Total liquid feed to melter (gpm)

(Hydrate feed rate constant)
to the melter. The hydrate feed rate remained constant during these runs. Figure XV shows the gas pressure drop across the melter with each of the types of packing tested. Examination of the data in Figures XIV and XV indicate that while Flocor had a lower pressure drop the disc and donut arrangement had the higher volumetric heat transfer coefficient.

It is obvious that further testing is needed to improve heat transfer in the plant melter. Intuitively the melter volumetric heat transfer coefficient could be greatly improved by finding an internal arrangement which would eliminate at least some of the large void space in both the disc and donut and Flocor packings while maintaining a low pressure drop across the melter, minimizing the energy required in the primary compressor. Packing other than Flocor needs to be evaluated with this objective in mind. Such testing was planned later in the program but more important work on the pressurized wash column preempted these plans.

XVI. B. Small Scale Venturi Eductor Testing

Primarily because of its low capital cost, a venturi eductor was considered as a melter-contactor for the pilot plant. The venturi had a further advantage in the hydrate process since it could be driven by a relatively high pressure stream of hydrate slurry which would be available when a pressurized wash column was installed in the plant. For the venturi melter to work properly, it must accomplish three functions. First, it must entrain into the throat of the venturi sufficient propane to melt hydrate; secondly, it must compress the slurry-gas mixture to sufficiently high pressure in the divergent section of the venturi to reach melting conditions; and third, it must achieve enough heat transfer to melt the hydrate.

The development work on the venturi initially concentrated on the first two of the above problems, entrainment of propane gas and the compression of this gas. To accomplish this work a small pilot unit was constructed. A flow diagram of the installation appears as Figure XVI of this report.

The venturi eductor, as applied in the Sweet Water Pilot Plant, used the high pressure of a motive slurry fluid converted to kinetic energy at the throat of the eductor to entrain propane gas and reconverted the kinetic energy of the resultant gas-slurry stream to pressure at the outlet of the diffuser. The above energy conversions of the fluids involved, neglecting inefficiencies, are in accordance to the equation

\[ \Delta H = \frac{u_2^2 - u_1^2}{2 \ gc} \]

where

- \( \Delta H \) = pressure change in feet of fluid
- \( u \) = velocity in feet per second
- \( gc \) = gravitational constant
FIGURE XI: COMPARISON OF MELTER PRESSURE DROP
VS. TOTAL LIQUID FEED TO MELTER.
DESCRIPTION OF EQUIPMENT

<table>
<thead>
<tr>
<th>ITEM NO</th>
<th>NAME</th>
<th>DESCRIPTION</th>
</tr>
</thead>
<tbody>
<tr>
<td>-</td>
<td>RECIPIENT TANK</td>
<td>VERTICAL TANK (APPROXIMATELY 1500 G)</td>
</tr>
<tr>
<td></td>
<td>LIQUID RETURN PUMP</td>
<td>20 gpm @ 15 psi head</td>
</tr>
<tr>
<td>R-1</td>
<td>ROTAMETER</td>
<td>100% = 12.98 ppm @ 65 psig and 40°F</td>
</tr>
<tr>
<td>R-2</td>
<td>ROTAMETER</td>
<td>100% = 10.01 gpm @ 95°F, 1.00</td>
</tr>
<tr>
<td>R-3</td>
<td>HAND CONTROL</td>
<td>100% = 5.45 ppm @ 70 psig and 60°F.</td>
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<td>HC</td>
<td>VALVES</td>
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<td>PG</td>
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<td>LG</td>
<td>LEVEL GAUGES</td>
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</tr>
</tbody>
</table>

Figure XVI

Sweetwater Development Co.
1300 Expressway Tower
Dallas, Texas

Process Flow Diagram for Venturi Eductor Testing
The theoretical energy required to compress the propane gas is given by the equation

\[ W = \frac{kRT}{k-1} \left[ 1 - \left( \frac{P_2}{P_1} \right)^{\frac{k-1}{k}} \right] \]

where

- \( W \) = work
- \( R \) = gas constant
- \( T \) = suction temperature
- \( \frac{P_2}{P_1} \) = compression ratio
- \( k \) = ratio of specific heats for the gas

Using these equations it is possible to compute the theoretical performance of the eductors for comparison with test results.

A number of eductor designs were tested using a plexiglas eductor constructed so that various liquid nozzles and diffuser sections could be tested. Table V lists the eductors tested and gives the critical design dimensions of each of the eductors.

The efficiency of energy recovery for several of the eductors tested appears as Figure XVII. The best of the plexiglas eductors achieved an energy recovery efficiency of 18% at a compression ratio of 1.03. Test eductor CR-1, manufactured by Croll Reynolds, was the best eductor tested, achieving a 19.5% efficiency at a 1.055 compression ratio.

After the most efficient eductor from an energy recovery consideration had been determined, testing began to determine the heat transfer coefficient existing in an extension of the diffuser of a venturi eductor. The diffuser extension tested consisted of 100 ft of 1\( \frac{1}{2} \) in. schedule 40 pipe. The diffuser was constructed of 10 lengths of 10 ft each of this pipe connected by 180° standard radius pipe returns. The diffuser shown on the flow diagram in Figure XVI was not included in the original energy recovery work but was added for the heat transfer testing.

Selected heat transfer data appear as Figure XVIII. The tests show quite a bit of scatter but it is possible to see that the heat transfer coefficient is approximately 1,400 Btu/hr cu ft - OF.
## SWEET WATER DEVELOPMENT COMPANY

### TABLE V

**TEST EDUCTOR CRITICAL DESIGN DIMENSIONS**

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<tr>
<th></th>
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<td>0.058</td>
<td>0.082</td>
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<td>0.110</td>
<td>0.129</td>
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<td>4.95</td>
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<td>0.335</td>
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<td>0.378</td>
<td>0.209</td>
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<td>1.380</td>
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<td>P-6</td>
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<td>0.50</td>
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<td>1.437</td>
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<td>5.991</td>
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</table>

*Short diverging section at discharge of liquid nozzle gave pattern spread.
Figure XVII: Test Diffuser Volumetric Heat Transfer - VS Driving Force

\[ \Delta T = T_e - T_h \]

- \( T_e \) = Propane Equilibrium Temperature
- \( T_h \) = Hydrate Melting Temperature
XVII. **Pressurized Wash Column Theory**

Sketches showing the basic operating principles of a flooded and pressurized wash column for propane hydrate are found in Figure XIX. These operating principles are essentially the same as those outlined by A. J. Barduhn (Ref. 1, page 62) for drained wash columns. A cylindrical wash column with screens mounted on the shell of the vessel is used for the illustration. Slurry containing 10 to 20 percent hydrate, aqueous basis, is fed into the bottom of the column and moves upward. The hydrate consolidates at some level and becomes a moving packed bed having approximately 40 percent solids concentration, aqueous basis. The brine component of the slurry flows up through the bed and leaves the column in the horizontal direction through screens mounted on the wall of the column. The hydrate bed continues up past the screens and encounters wash water which has a small downward velocity relative to the wall of the column, which occupies the interstices of the bed in the upper part of the column. This wash water displaces and washes brine from the hydrate and is discharged through the screens along with the reject brine. The washed hydrate is removed from the bed by a scraper at the top of the column, and is then re-slurried and removed from the column. Product water is used both for washing the hydrate and for reslurrying the washed hydrate. Only that part of the product water discharged through the screens, usually equivalent to 5 to 10 percent of the washed hydrate production, is consumed in washing. The height of the wash column must be at least the sum of the heights required for all of these operations.

From the inlet at the bottom of the column to the screens, the distance must be adequate to provide the force necessary to push the hydrate bed upward. The required force is the sum of the liquid head in the column, buoyant force of the hydrate in brine and wash water, the downward force of the scraper on the bed, the frictional forces between the wall of the column and the moving bed, and the downward force applied at the top of the wash column to drive the wash water down through the rising bed. The force applied at the top of the bed is usually large compared to scraper, buoyant, and frictional forces. Excluding liquid head, the required force at the bottom of the bed is therefore slightly larger than the force applied at the top of the bed. The hydrate bed below the screens required for generating the upward force is called the driving section. The height of the driving section adjusts itself to give the proper pressure drop to the brine flowing up through the hydrate bed to provide this force. The permeability and the porosity of the hydrate bed determine the height of the driving section. A space below the bottom of the driving section is usually provided for slurry distribution to the bed, as well as for operating variations such as particle size, slurry concentration and operational upsets.

Screen section height is based on two important considerations. First, the screens should have adequate clear opening area to insure that the reject brine and wash water encounter a small pressure drop in passing
**Figure XIV** Flooded and Pressurized Wash Column Operation

- **A. Hydrate Movement**
- **B. Liquid Flow Pattern in Bed**
- **C. Salinity of Interstitial Liquid Along Wash Column**
For the design of Sweet Water's 20,000 gpd wash column, the following values were used:

\[ B_0 = 5 \times 10^{-8} \text{ cm}^2 \text{ which corresponds to an effective particle diameter of about 30 microns} \]

\[ \rho = 1.03 \frac{\text{gm}}{\text{cm}^3} \]

\[ \kappa = 1.8 \times 10^{-2} \frac{\text{gm}}{\text{cm sec}} \]

The permeability used above was computed using pilot plant data on pressure drop through a packed hydrate bed. The Carmen-Kozeny equation then reduces to:

\[ \left( \frac{Q}{A} \right) = 2.80 \times 10^{-3} \frac{\text{cm}}{\text{sec}} \left( \frac{h}{ft} \right) \]

Using English units, the equation can be written in the following form:

\[ \left( \frac{gpm}{ft^2} \right) = \left( 0.102 \frac{gpm}{ft \ psi} \right) \left( \frac{psi}{ft} \right) \]

where

\[ \left( \frac{gpm}{ft^2} \right) = \text{water flow through the bed} \]

\[ \text{cross sectional area of bed} \]

\[ \left( \frac{psi}{ft} \right) = \text{pressure drop along bed length} \]

\[ \text{length of bed} \]

Note that \( \left( \frac{gpm}{ft^2} \right) \) can be reduced to units of length/time, and represents the superficial velocity of water with respect to the bed. In the washing section of a wash column, the minimum value of \( \left( \frac{gpm}{ft^2} \right) \), \( V_{min} \), is defined as the downward velocity of wash water relative to the bed which is numerically equal to the upward velocity of the bed relative to the wall of the vessel. In other words, \( V_{min} \) is the downward velocity of wash water relative to the bed when the downward velocity of wash water relative to the wall of the vessel is 0.

\( V_{min} \) has been related to pressure drop along bed length, and these quantities may now be used to compute the minimum wash column cross sectional area required for producing the plant design capacity of fresh water.

\[ W = \nu A \epsilon \gamma \chi \]

where

\[ W = \text{product water flow - gpm} \]

\[ \nu = \text{velocity of bed relative to wall of wash column} = V_{min} \left( \frac{gpm}{ft^2} \right) \]

\[ A = \text{cross sectional area of cake - ft}^2 \]
\[ \varepsilon = \text{solids fraction porosity} = 0.42 \text{ from average of permeability test data} \]
\[ \rho_H = \text{specific gravity of propane hydrate} = 0.88 \]
\[ x = \text{weight fraction product water in propane hydrate} - 0.88 \]

Using this relationship, the following table was constructed for Sweet Water's 20,000 gal/day propane hydrate plant.

### TABLE VI

**Areas and Diameters for 20,000 gpd Propane Hydrate Wash Column at Various Imposed Pressure Drops**

<table>
<thead>
<tr>
<th>Imposed Pressure Drop - top of bed to screens - psi</th>
<th>Water Production gpd</th>
<th>20,000 gal/day Wash Column ft² Area Required ft²</th>
<th>Diameter ft</th>
</tr>
</thead>
<tbody>
<tr>
<td>0 (gravity only)</td>
<td>21</td>
<td>952</td>
<td>34.8</td>
</tr>
<tr>
<td>2</td>
<td>116</td>
<td>172</td>
<td>14.8</td>
</tr>
<tr>
<td>4</td>
<td>212</td>
<td>94</td>
<td>10.9</td>
</tr>
<tr>
<td>6</td>
<td>308</td>
<td>65</td>
<td>9.1</td>
</tr>
<tr>
<td>8</td>
<td>403</td>
<td>50</td>
<td>8.0</td>
</tr>
<tr>
<td>10</td>
<td>499</td>
<td>40</td>
<td>7.1</td>
</tr>
<tr>
<td>12</td>
<td>594</td>
<td>34</td>
<td>6.6</td>
</tr>
</tbody>
</table>

The washing section height may be estimated using a method discussed by Allen J. Barduhn (Ref. 1). For a propane hydrate wash column, the theoretical washing section height is estimated to be on the order of one inch.

Methods for calculating brine crown height have been set forth by H.F. Wiegandt, R. L. Von Berg and J. P. Leinroth (Ref. 2); W. J. Hahn (Ref. 3); J. C. Orcutt, F. O. Mixon, and F. J. Hale (Ref. 4); and others. The method used in estimating the brine crown height for Sweet Water's wash column was based essentially on methods appearing in Ref. 4; however, the method as used appeared in an earlier preliminary report on the same work by F. O. Mixon (Ref. 5). The analog method given by H. F. Wiegandt et al (Ref. 2) is particularly interesting because the results are obtained in graphical form, and the method can be readily adapted to any wash column configuration. The method makes use of a physical model which eliminates much of the complexity involved in the mathematical model.

Standard hydraulic calculations are used to assure that the reject brine and product water do not encounter an excessive pressure drop in the screens themselves. To assure that the pressure drop in the bed near the screens is not excessive compared to that in other parts of the column, some method of estimating flow potential gradient must be used.
through the screens. Second, the overall area of the screens must be large enough to insure that the reject brine and wash water do not undergo too large a percentage of the available pressure drop between the top (or bottom) of the bed and the screens while passing through the bed in the immediate vicinity of the screens. This might lead to a well compacted cake near the screens and an inadequate compaction of the cake in other areas of the column, which in turn would result in poor washing.

The height of the column above the screens must include space for the brine crown, for adequate washing and hold time for operational adjustments. The brine crown results from the fact that some horizontal flow potential must be applied to the brine to move it horizontally through the bed to the screens and through the screens. Roughly speaking, the brine in the bed "piles up" high enough to provide the necessary horizontal force. Generally, the highest part of the brine crown is found in that part of the column farthest from the screens. For the example given in Figure XIX-B, this is at the center of the column.

The washing section lies just above the brine crown, and is the section in which most of the washing takes place. The washing is accomplished by a combination of the wash water displacing brine from the interstices of the bed, and of the wash water diluting and carrying off the brine adhering to the surface of the crystals. The interstitial wash water flow is laminar, and most of the brine is removed from the bed by displacement. Below the washing section, the interstitial liquid salinity is essentially the same as that of reject brine. Above the washing section, the interstitial liquid salinity is essentially the same as product water. This is graphically shown in Figure XIX-C.

The basic wash column design parameters - such as height of the packed bed, cross sectional area, and operating pressures - may be related using the Carmen-Kozeny equation for fluid flow through porous media:

\[
Q = \frac{B_0 \rho gh}{\mu L}
\]

where

\begin{align*}
Q & = \text{volume of fluid per unit time} = \text{cm}^3 \\
A & = \text{cross sectional area of porous medium perpendicular to flow of fluid} = \text{cm}^2 \\
B_0 & = \text{permeability of porous medium} = \text{cm}^2 \\
\rho & = \text{density of fluid} = \text{g/cm}^3 \\
g & = \text{gravitational constant} = 980 \text{ cm/sec}^2 \\
h & = \text{hydraulic head loss through bed along direction of flow} = \text{cm} \\
\mu & = \text{viscosity of fluid} = \text{g cm/sec} \\
L & = \text{length of porous medium along direction of fluid flow} = \text{cm}
\end{align*}
The analog wash column design method set forth by H.F. Wiegandt et al (Ref. 2) involves graphical construction of lines of constant flow potential. In a pressurized wash column, these lines radiate out from the screens (0% flow potential) to the top and bottom of the bed (100% flow potential). Using this method the screen area and height may be designed so that the distance from the 0% to the 10% lines (pressure drop near the screens) does not differ excessively from the distance between the 90% and 100% lines (pressure drop near the top or bottom of the bed).

The driving section height must be great enough to subject the reject brine passing up through the bottom of the bed to a pressure drop great enough to push the bed upward through the column. The basic equations for pressure drop in porous media apply in the design of driving section height.

Some leeway should be left near the top of the cake for changes in brine crown height resulting from process fluctuations. Also, some space should be left below the driving section for slurry distribution, and for fluctuations in driving section length. In a wash column used for development work, some extra height should be provided to facilitate future modifications.

XVIII. Description of Pressurized Wash Column

An annular wash column with 96 inches I.D. outer shell and 18 inches O.D. inner shell was chosen for Sweet Water's 20,000 gpd design. The cross sectional washing area with this configuration is 47.5 sq ft including provision for eight drain tubes. The imposed back pressure required for producing 20,000 gpd fresh water would therefore be between 8 and 10 psi/ft. A schematic cutaway drawing of this wash column is shown in Figure XX.

The harvested hydrate is scraped in toward the center of the column and is reslurried. The product slurry then passes down through the inner annular shell of the column, and is removed through a nozzle in the bottom of the column. Water for reslurrying harvested hydrate and for washing could be provided by either a sluice ring mounted just above the scraper blades around the outside diameter of the column or by showers mounted on the scraper arms.

Drain tube mounted screens were chosen over shell mounted screens for the wash column because they do not require disruption of the shell of the vessel, and because the screen height or position can be more readily changed later as needed. Eight drain tubes were originally mounted on a header in the bottom of the wash column.

The slurry feed to the wash column enters through two nozzles in the bottom of the vessel, and is distributed by a tee shaped header mounted
Figure XV

Sweet Water Development Co.
1500 Expressway Tower,
Dallas, Texas

-Cutaway-
Wash Column Internals
on each nozzle, giving a total of four connected slurry entrances for the column. Elbows are mounted at both ends of the cross bar of each tee so the slurry is discharged downward toward the bottom head of the column. This was done to prevent any bed erosion that might result from discharging the slurry either upward or horizontally.

Figure XXI shows a graphical representation of the design heights of the various sections of Sweet Water's wash column. The bases for the design were discussed in the previous section.

Beginning at the bottom for descriptive purposes, two feet was provided for slurry distribution below the driving section. The driving section was calculated to be 1 1/2 ft long and an additional 3/4 ft was added to allow for process fluctuation. The resulting driving section height was 3-1/4 ft. The screens originally installed in the wash column consisted of 1/16 in. holes drilled in 4 in. Sch. 80 PVC drain pipes, with the holes on 1/4 in. centers horizontally and vertically. The screen section formed in this manner was made 9 in. high, giving a total clear screen opening area of 51.5 sq in. and an overall screen area of 7.07 sq ft. This should result in minimal pressure drop across the screens with the reject brine flow used in this design.

The brine crown height was calculated to be a maximum of 24 1/2 in. above the screens. The conservative assumptions that the screens were on two shells of an annulus and that the screen height was less than 0.1 in. were used in this calculation. Another 5 1/2 in. was then added as a safety factor to bring the brine crown height to an estimated 30 in.

Although the height of the washing section was calculated to be on the order of one in., six in. was provided for this purpose. One foot was allowed between the top of the cake and the top of the washing section to allow for fluctuations in brine crown height.

The total design height required for producing 20,000 gpd fresh water in the propane hydrate process is 10 ft. An additional 2.75 ft was added to this height to provide for any future changes which might become necessary. Initially, this height was added to the slurry distribution section. During the period January 1, 1968 to August 31, 1968, six screen arrangements were tested. Schematic drawings of these screen arrangements are found in the appendix.

The 20,000 gpd wash column was installed in Sweet Water's pilot plant as shown in the appended Process Flow Diagram FS-68-1. A great deal of flexibility was incorporated into the installation. The pressure for operating the column is supplied by the slurry transfer pump GA-103 and by pump GA-107 which furnishes both wash water and washed hydrate re-slurring water to the top of the column. PRC-107 controls pressure in
Figure XXI
Sweet Water Development Co.
1300 Expressway Tower
Dallas, Texas.
Wash Column
Original Design Heights
As Installed Jan. 1, 1968

Dwg. No. 6808
the top (and therefore in the bottom) of the column by operating a control valve in the washed slurry stream leaving the column. The reject brine leaving the column is flow controlled by FRC-125, and the screen pressure depends on the setting of this control valve. The pressure used to force wash water down through the rising hydrate bed is therefore supplied by pump GA-107 and is controlled by the combination of the top pressure controller and the reject brine flow controller. The slurry transfer rate and the wash-reslurry water rate are both flow controlled. This instrumentation has proved quite satisfactory for controlling wash column operations.

XIX. Wash Column Pilot Plant Testing

Two critical factors in wash column operation, both involving the scraper, became apparent soon after the wash column was put into operation on December 27, 1967. First, the proper amount of reslurrying water must be supplied to the top of the column. If not enough reslurrying water is supplied, washed hydrate builds up in the top of the column. This buildup overloads and stalls the scraper, and forces a plant shutdown. Too much reslurrying water, which is supplied at melter liquid temperature, can erode or partially melt the hydrate bed. An ammeter connected to the scraper motor gives an indication of scraper loading. Wash and reslurrying water flow was adjusted to keep the scraper partially loaded (indicating little or no bed erosion), but not overloaded. In addition, the melter liquid temperature was kept as low as operations would permit.

The second critical factor in wash column operation is the pressure balance between the top and bottom of the column. Under normal operations, the pressure at the bottom of the column is slightly larger than that at the top, but only enough to push the packed hydrate bed upward at the required rate. If, during a plant upset, the pressure at the bottom of the column suddenly becomes too large compared to the pressure at the top of the column, the hydrate bed exerts an excessive force on the scraper and bends the arms upward. In normal operations, the opposing forces at the top and bottom of the column are both on the order of 100 tons.

On several occasions, the PVC drain tubes originally installed cracked or broke at points of stress concentration. Attaching wire cloth screens to the PVC tubes proved to be a problem. For these reasons, carbon steel drain tubes were used in the last two screen arrangements, and have proved structurally sound. The monel wire cloth screens were soldered directly to the steel tubes, and this method of attachment has been very satisfactory.
During most of the pilot plant testing, the wash column was operated with 100 percent net wash, i.e., all of the product water produced was returned and used as wash water. The plant was normally operated on closed loop with no reject brine or product water leaving the plant. Only enough seawater was added to maintain an adequate aqueous inventory in the plant. With the plant on closed loop, product water circulates through the wash column center shell, the melter, the melter recycle hold tank, the separator, and back to the top of the wash column. The product water inventory in this cycle is normally about 2,000 gallons. Product water salinity is therefore a measure of the ability of the wash column to dilute this cycle with washed or partially washed hydrate.

During all the wash column runs, the brine salinity in the crystallizer was kept below 6.0 percent (42 percent seawater conversion), by draining liquid from the reject brine stream from the wash column and adding either seawater or city water to the crystallizer. This was done to limit the primary compressor load, to limit the amount of washing required, and to prevent crystal habit deterioration that might occur with higher brine salinity.

Most of the wash column runs were conducted with a constant slurry transfer flow rate and a constant pressure at the top of the wash column. Product water inventory was controlled by making small adjustments in the reject brine flow which also controls the amount of wash water leaving the column through the screens.

An outline of all the wash column runs is found in Appendix G.

The pilot plant testing has shown most of the design calculations to be accurate. The assumptions that the pressures at the top and bottom of the bed would be nearly equal and that both would be considerably larger than the screen pressure (all corrected for hydrostatic head) were found to be valid. The pressure difference between the bottom and top of the bed required for driving the bed upward was usually about 2 psi, corrected for hydrostatic head.

Samples taken from taps mounted on the side of the wash column showed the top of the brine crown to be 2 to 3 ft above the top of the screens under normal operating conditions. It was impossible to determine the actual washing section height from these samples, but there was no indication that this height exceeded the design washing section height. The samples indicated driving section height to be on the order of one foot.

Permeability was the only quantity for which actual test data differed significantly from design data. The permeability obtained from actual wash column testing was much higher than the value used in the design.
Screen arrangements 1 and 2 (see Appendix F) consisted of 1/16 in. holes drilled in the PVC drain tubes. These screens allowed a considerable amount of hydrate to pass through into the reject brine stream, even when the hydrate bed was apparently formed. Subsequent screen arrangements using 30 x 150 mesh or 24 x 150 mesh monel wire cloth did not allow hydrate to enter the reject brine stream. Screen arrangement 4, with wire cloth, and screen arrangement 2, with 1/16 in. holes, are practically the same in all other respects. The minimum product water salinity using screen arrangement 4 was about half the minimum product water salinity attained with screen arrangement 2.

Screen arrangement 4 gave lower product water salinity than screen arrangement 3, probably because of crowding of pressure gradient in the bed near the small screens used in arrangement 3.

Screen arrangements 5 and 6 consisting of 16 tubes with 2 ft high screen gave better washing than any of the previous arrangements. The minimum product water salinity attained in all the wash column testing was 761 ppm using screen arrangement 5. It is not known how much the improved washing was due to the fact that arrangements 5 and 6 have four times the overall screen area of any of the previous arrangements, and how much it was due to the improved screen distribution obtained by using sixteen tubes instead of eight.

In many of the wash column runs the product water salinity reached a minimum and leveled out near this minimum. Thus the overall salinity of hydrate and interstitial liquid leaving the wash column was the same as that in the product water inventory. Since the hydrate contains pure water, this means some brine was reaching the top of the wash column. In stable operations the data indicated the top of the brine crown was always below the top of the hydrate bed. These data were obtained from samples taken from taps on the outside wall of the wash column, where maximum brine crown height should occur. When the product water salinity reached its minimum and stabilized, the brine reaching the top of the column was either occluded on the surface of the crystals or a non-uniformity in the hydrate bed was allowing brine to reach the top of the bed somewhere other than at the sampling taps. For reasons given below the latter condition seems more likely.

During the most successful runs using screen arrangements 5 and 6, full plant design wash water flow was attained using only 1 to 2 psi/ft imposed pressure gradient from the top of the bed to the screens. This is compared to the approximate design value of 10 psi/ft which was based on permeability test data. With a 10 percent slurry concentration and 100 percent net wash, the upward reject brine flow generates a pressure gradient in the driving section of the column which is about five times
the pressure gradient imposed in the top of the column by the wash water. The bed in the wash column was therefore continuously formed with a pressure gradient 5 to 10 psi/ft. The bed in the permeability tester is formed with a pressure gradient ranging from 50 to 200 psi/ft. Within the range of bed formation pressure gradients required to operate the wash column at design flows and assuming that there was no channeling in the wash column, it appears that the permeability of the bed formed strongly depends on the pressure gradient used in the formation of the bed.

It follows that additional screen area and more uniform screen distribution should be incorporated into the wash column. This can be done by modifying the existing drain tubes, and by adding more drain tubes. In addition, the effect of formation pressure gradient on hydrate bed permeability should be determined. This approach to developing the wash column was planned when work on the project was cancelled.

XX. Process Economics

Principal items of equipment (PIE) have been designed for a one million gallons per day propane hydrate water desalting plant. These designs were based on pilot plant knowledge and a considered extrapolation of this knowledge. The installed cost of these items were then obtained from manufacturers of the equipment where required, and using standard cost estimating procedures corrected for 1967 prices from Nelson's Cost-imating, Chilton's Cost Engineering in the Process Industries, and Vilbrandt and Dryden's Chemical Engineering Plant Design (see Appendix I) for this estimating.

The economics of the process is summarized below as:

<table>
<thead>
<tr>
<th><strong>Plant Capital Cost</strong></th>
<th>$ 921,700</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Turn Key Basis</strong></td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th><strong>Capital Cost Per Daily Gallon</strong></th>
<th>$ 0.922</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th><strong>Total Operating Cost Per Stream Day</strong></th>
<th>$ 690.54</th>
</tr>
</thead>
</table>

<table>
<thead>
<tr>
<th><strong>Cost per Thousand Gallons of Fresh Water</strong></th>
<th>$ 0.6905</th>
</tr>
</thead>
</table>

- 60 -
XXI. **Recommendations**

The highly favorable economics appearing in the preceding section demand that the development of the propane hydrate process be continued. Based on the test results given in this report, the following recommendations are presented.

Wash column development should be continued by incorporating additional screen area and more uniform screen distribution into the existing wash column. This can be done by modifying the existing drain tubes and by adding more drain tubes. The effect of formation pressure gradient on hydrate bed permeability should be determined.

The actual liquid volume utilized for crystallization in the maker is not known. The effect of very short hold times on crystallizer production should be studied. This can be done by using partitions to reduce the limits of active volume in the existing maker.

The effect of hydrostatic head on propane hydrate crystallization should be studied. This information will be vital in scaling up the maker for large capacity plant designs.

The present melter is a packed column used for transferring heat from a rising propane vapor stream to a descending liquid propane and water-hydrate slurry stream. With this counter current arrangement, an increase in column capacity or volumetric heat transfer coefficient is usually accompanied by an increased pressure drop for the vapor stream, and the column capacity is limited by the flooding velocity. If the streams are contacted co-currently, the vapor stream pressure drop is very low compared to the counter current arrangement, and flooding velocity is not as critical. The application of the co-current melter to the propane hydrate process should be investigated in the pilot plant. The existing melter, with minor piping changes, can be used for this testing.

The use of the venturi-educator as a melter has been shown to be feasible if certain process conditions exist. When more development has been carried out and propane hydrate process conditions are more firmly established, the feasibility of using a venturi-educator melter should be re-evaluated.
XXII. References


XXIII. APPENDIX
APPENDIX A

Cyclone Wash System Process Flow Description

A flow sheet for the pilot plant as constructed and with minor revisions appears as Drawing FS-64-I (Appendix A, page 66). Using this flow sheet as a guide, and the inlet seawater as a beginning, the flow pattern can be traced. The incoming seawater, before entering the plant, passes through a filter and then to a vacuum deaerator where dissolved gases such as CO₂, O₂, and N₂ are removed. The deaerated seawater is then split into two streams and passes through heat exchangers EA 101 and EA 102. In these heat exchangers the incoming seawater is counter-currently cooled by the discharge product water and reject brine streams from the plant.

After being cooled in these heat exchangers, the deaerated seawater passes through a chiller, EA 104, and is admitted into the washing system where it performs part of the washing operation. The flow pattern in the washing section is as follows. It first enters FA 112 where it settles and is removed at the bottom of the tank. It then passes through pump GA 119 into cyclone FC 110. The brine reports to the underflow (as it does in all of the cyclones) of FC 110 through pump GA 120 and is then admitted into cyclone FC 111. From the underflow of FC 111, the incoming seawater joins the recycle brine stream and is fed to the maker or crystallizer, DC 101.

The seawater and recycled brine are intimately mixed with liquid propane in the maker. Part of the propane vaporizes, cooling the mass and removing the latent heat of formation of propane hydrate. The remaining propane reacts with the water to form propane hydrate crystals. The slurry thus formed is removed from the maker through pump GA 103 and is fed into the cyclone system. Before entering the cyclones, liquid propane is added to the slurry. The feed to the first cyclone is now a mixture of brine, hydrate, and liquid propane. In cyclone FC 112, a major portion of the brine is removed as an underflow and goes to the reject brine gravity separator FA 106. From separator FA 106 a reject brine stream is taken from the plant. The remaining brine overflows from FA 106 and is recycled to the maker. The propane and hydrate mix in the slurry stream that was fed to FC 112 reports to the overflow of the cyclone due to their lighter densities and is then fed consecutively from cyclone FC 111 through the nine remaining cyclones. In each of these cyclones the lighter hydrate and liquid propane mix reports to the cyclone overflow stream. Before entering each of these cyclones, a stream of lower salinity water is contacted with the feed stream to the cyclone. This stream comprises the wash water for each stage and is the underflow from that particular cyclone.

After having passed through the cyclones, the liquid propane-propane hydrate slurry leaves cyclone FC 101 in the overflow and is fed into the melter DC 102. The slurry falls through the disc and donut melter from top to bottom where it is directly contacted with rising propane gas. The conditions in the melter are controlled such that the propane gas condenses, giving up its latent heat to decompose hydrate. The net result of this exchange is fresh water and liquid propane.
The fresh water and liquid propane flow from the melter to tank FA 109, a gravity separator, in which the propane rises and fresh water is removed from the bottom. The product water leaves the process through heat exchanger EA 101. Before being rejected from the system, however, the product water is cleared of any residual propane in depropanizer DA 101.

The propane, since it is not consumed process-wise, is continually being recycled. The flow pattern of the propane can be picked up in either the melter, the compressor, or maker section. For descriptive purposes, choose the liquid propane supply existing in the propane-water separator FA 109. Liquid propane from this tank serves two purposes. One purpose is to act as a source of diluent to the cyclone system. Thus it is seen that a flow of liquid propane moves from the propane separator to the flash tank FA 107. In this tank the propane which was at the equilibrium conditions of the melter, flashes to the equilibrium conditions of the maker, thus cooling itself so that it does not melt hydrate when it is contacted with the slurry from the maker.

The remaining liquid propane from tank FA 109 acts as a cooling medium and hydrating agent in the crystallizer. It is sparged into the maker where it is intimately contacted with the brine phase. Some of the propane enters the clathrate as the hydrating agent. The remaining propane vaporizes and is drawn through the primary knock-out pot FA 101 into the primary compressor GB 101. In GB 101 the gas is compressed from maker operating conditions, (approximately 60 psig) to melting conditions, (approximately 70 psig). In the melter, DC 102, the gas intimately contacts the slurred hydrate and most of it condenses and flows as liquid propane to the propane water separator, FA 109.

The propane which does not condense in the melter must be compressed in the secondary compressor, GB 102, to a pressure such that it will condense against cooling water in heat exchanger EA 103. The heat removed by cooling water in EA 103 is the heat that must be rejected to the surroundings to make up for heat leak, work inputs, and inefficient heat exchange of entering and departing streams from the process. The condensed propane in EA 103 is held in a tank FA 103 for a short time for surge purposes prior to being fed back to the propane separator FA 102 to begin its cycle once again.
Wash Column Process Flow Description

A process flow diagram follows for the propane hydrate pilot plant using the pressurized wash column. The process operates much the same with a pressurized wash column as with a cyclone wash system. Since the process operation with the cyclones was discussed in the previous section, this discussion will be limited to the differences in the two systems.

The incoming seawater is not used to wash hydrate in the pressurized wash column system as it was in the cyclone system and this stream can be fed directly into the maker. This eliminates the need for the chiller shown on the process flow diagram with the cyclone wash system.

The remaining difference in the two systems is of course the flow system associated with the wash column. Slurry from the maker is pumped through slurry transfer pump GA 103 and fed into the bottom of the wash column. Brine from the slurry is withdrawn from the wash column, leaving the hydrate as a consolidated porous plug which rises due to hydraulic forces against a descending wash water stream. The rising cake is cut up by a harvester and mixed with enough water from pump GA 107 to reslurry the hydrate and transport it to the melter where the hydrate is melted by direct contact with propane gas.

The reject brine removed from the wash column flows to a reject brine separation cyclone, FC 114. The overflow from the cyclone is recycled to the maker. The underflow from this cyclone is the reject brine from the system, and passes through heat exchanger EA 102 to the propane recovery flash tank FA 110. This tank was installed in the pilot plant primarily to recover entrained propane which sometimes found its way into the reject brine stream during plant upsets. The propane recovery flash tank on the product water line out of the plant serves the same purpose. After passing through the propane recovery flash tank the dissolved propane is removed from the reject brine in depropanizer DA 102.

The piping with the wash column installed was arranged so that the maker and melter were capable of operation without the wash column. This type of operation was accomplished by by-passing the wash column (line no. 20 on the diagram) and feeding the melter directly from the maker. The aqueous phase from the separator FA 109 can be returned to the maker through pump GA 122. This allows the pilot plant to operate on a closed system without a seawater feed, reject brine or product water flow.

It will be noted that several other items of equipment associated with the cyclone and propane diluent systems have been eliminated with the installation of the wash column. Most important of these from a capital cost standpoint are tanks FA 107, FA 112, and FA 106 as well as their associated piping and instrumentation.
APPENDIX C

Crystals - Photographic Method

In order to observe the crystals being produced in the maker, a viewer was constructed using two \( \frac{\frac{1}{2}}{\frac{1}{2}} \) in. thick \( \times \) 3 in. diameter pyrex discs clamped inside a brass support housing such that a thin layer of slurry could pass between the discs. This 3 in. diameter window was used to view the crystals, and with the aid of a microscope and a Polaroid camera, the pictures shown on the following page were made. These pictures were taken at a 50X magnification. Pictures were also taken at 100X. The pictures taken at 50X were found to be superior because of the limited depth of field at 100X. The pictures that were taken of hydrate crystals showed particles ranging from 50 to 200 microns. In general it appears that the crystals are flat plates although some cubic crystals were observed.
Crystals - Permeability Tester

A permeability tester was constructed to gain added information about the hydrate crystals being produced. The permeability tester was constructed of plexiglas. The tester has a smaller tube 1½ in. I.D. mounted inside a larger tube, the annular area being filled with slurry to prevent heat leak during operation. Slurry is admitted inside the smaller tube which has a filter medium arranged so that a cake of hydrate builds inside the smaller tube. A brine flow through this cake is metered and the resulting pressure drop over a given length of the bed is measured to allow the permeability to be calculated. If the void fraction or porosity of the bed is known, the particle size may be calculated. The porosity of the bed was measured to average 0.575. Using this porosity the particle size was calculated to range from 20 to 40 microns. Experiments with ice have shown that the actual average particle is usually 2 to 3 times the value obtained from a permeability test due to departure of the particles from a spherical shape. Using this rule of thumb the particle sizes obtained by photography and permeability tests are in general agreement.
APPENDIX E

OVERFLOW
5/8" O.D. X 0.065" WALL

METAL SCREW

CYCLONE INLET

3/4" O.D. X 0.065" WALL

2 1/2" O.D. X 0.065" WALL

METAL SCREW

UNDERFLOW
5/8" O.D. X 0.065" WALL

METAL SCREW

1 3/8"

2 3/8"

7"

12"

Not to Scale

SWEET WATER DEVELOPMENT CO.
1500 EXPRESSWAY TOWER,
DALLAS, TEXAS

STANDARD CYCLONE
2 1/2 INCH

DWG NO C1000

- 72 -
APPENDIX E

Not to Scale

SWEET WATER DEVELOPMENT CO.
1500 EXPRESSWAY TOWER
DALLAS, TEXAS

CYCLONE WITH FLOATING INSERT - 30 MESH WIRE CLOTH INSERT

DWG. No. C 1040

- 73 -
(A) Conventional

Slurry → Mixing Tee → Mixed Feed → 1/2" O.D. Feed Nozzle → Diluent → Cyclone Cross-Section at Inlet.

(B) Split Feed

Slurry Toward Cyclone Wall → Flat Plate Separator → Streams Contact

Diluent Toward Cyclone Center → 3/4" O.D. x 18" Long

(C) Split Feed

Diluent Toward Cyclone Wall → Diluent → Slurry Toward Cyclone Center

Conventional & Split Feed Cyclone Configuration

Sweet Water Development Co.,
1500 E. Pershing Tower,
Dallas, Texas
Dwg. No. 1050
DILUENT INLET

$\frac{1}{4}$" COPPER COMP. FITTING; $\frac{1}{4}$" THRU DRILLED.

$\frac{1}{4}$" TUBING TANGENTIAL INLET

$\frac{5}{8}$" O.D. x $\frac{1}{8}$" I.D. TUBE

SCALE: FULL SIZE
MATERIAL: COPPER

$\frac{1}{4}$" COMPRESSION FITTING SAME AS ABOVE. SOLDER IN PLACE.

UNDERFLOW

$\frac{1}{4}$" TUBING

OVERFLOW

$\frac{3}{4}$" TUBING TANGENTIAL TO INSIDE WALL

SWEETWATER DEVELOPMENT CO.
1300 EXPRESSWAY TOWER
DALLAS, TEXAS

UNIDIRECTIONAL FLOW TYPE CYCLONE
UF-I

Dwg. No. 4000
Diluent Inlet

\( \frac{1}{4} \)" Tubing

\( \frac{1}{4} \)" Copper Comp. Fitting; \( \frac{1}{4} \)" Thru Drilled.

\( \frac{5}{8} \)" O.D. x \( \frac{1}{2} \)" I.D. Tube

\( \frac{1}{4} \)" Tubing Tangential to Inside Wall.

Underflow

\( \frac{1}{4} \)" Compression Fitting Same as Above. Solder in Place.

Overflow

Fed Inlet

Scale: Full Size
Material: Copper

Sweet Water Development Co.
1500 Expressway Tower
Dallas, Texas

Unidirectional Flow Type Cyclone
UF-11

Dwg. No. 4010
APPENDIX E

Cyclone Diagram

scale 1\(\frac{1}{2}\)" = 1"

Vortex Finder Inserted Here

Overflow

Drill 5/8"

3/4" NPT

1/4" NPT

3/8" O.D. x 20Stubs Hard Brass Tube

All Brass

1/2"of 3/8" O.D. Comp. Tube Fitting, Sweat Soldered to Tube. Thru-Drilled 3/8"


Scale = Full Size

Detail of Vortex Finder

Sweet Water Development Co.
1500 Expedition Tower
Dallas, Texas

Van Tongeren Cyclone

DWG No. 5000

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APPENDIX E

Spiral Device "A"

3/4" Boring Bit

Spiral Device "B"

3/4" Ships Auger

Spiral Device "C"

1/4" Tubing (Flattened) Wound on 5/8" Rod.

Comparative Drawing of Van Tongeren Cyclone Impellers

Note: Drawings not to scale. All impellers are 5" long & 3/4" o.d.

Sweet Water Development Co.
1500 Expressway Tower
Dallas, Texas

Spiral Devices For
Van Tongeren Cyclone

DWG.N° 5010
NOTE: ORIGINAL CYCLONE DID NOT HAVE VORTEX FINDER, THUS NECESSITATING INSTALLATION OF PART A-2 & INSERT "1".

PART A-2

"O" RING

INSERT "1"

PART B

3/8" NPT OVERFLOW

3/8" NPT

PART C

3/8" NPT

PART D

UNDERFLOW

2.005 ± 0.001" O.D.

0.263 ± 0.001" I.D.

1.000 ± 0.001" I.D.

EXPLODED VIEW

NO SCALE

SWEET WATER DEVELOPMENT CO.
1300 EYEXPRESSWAY TOWER
DALLAS, TEXAS

1" PLEXIGLAS CYCLONE
ASSEMBLY

DWG. NO. 6000

- 80 -
PLAN

SCALE 3/4"=1'-0"

VEssel Shell
96" I.D.

58" D

Center Pipe
18" O.D.

(8) Drain Pipes
4.500" O.D.
(Screens Mounted on O.D.)

Sweet Water Development Co.
1500 Expressway Tower,
Dallas, Texas

Wash Column
First Through Fourth Screen Arrangements
Jan. 1, 1968 to May 26, 1968

DWG. No. G801
APPENDIX F

ELEVATION

Sweet Water Development Co.
1500 Expressway Tower, Dallas, Texas
WASH COLUMN
SIXTH SCREEN ARRANGEMENT
JULY 6, 1968 - AUGUST 31, 1968

Dwg. No. G807
### SUMMARY OF WASH COLUMN OPERATIONS - 1968

<table>
<thead>
<tr>
<th>Run No.</th>
<th>Run Time</th>
<th>Screen Arr. (See Appendix F)</th>
<th>Slurry Feed Rate gpm</th>
<th>Hydrate Prod. Rate as Water-kgpd</th>
<th>Pressures - psig Corrected for Hydrostatic Head</th>
<th>Minimum Prod. Water Salinity ppm</th>
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<tr>
<td>1</td>
<td>4</td>
<td>1</td>
<td>67-112</td>
<td>14-15</td>
<td>104-106</td>
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<td>2</td>
<td>5</td>
<td>1</td>
<td>70</td>
<td>12</td>
<td>104-106</td>
<td>103</td>
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<tr>
<td>3</td>
<td>6</td>
<td>1</td>
<td>65</td>
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<td>96-120</td>
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<td>7</td>
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<td>11-25</td>
<td>96-106</td>
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<td>82-108</td>
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<tr>
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<td>6</td>
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<td>8</td>
<td>104</td>
<td>93-100</td>
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<td>17</td>
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<td>6</td>
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<td>98-97</td>
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<tr>
<td>19-24</td>
<td>17</td>
<td>6</td>
<td>Var.</td>
<td>Var.</td>
<td>Var.</td>
<td>Var.</td>
</tr>
</tbody>
</table>

*Verified by Wrightsville Beach Test Facility Laboratory
Subject Invention Statement

There were no subject inventions conceived or reduced to practice by Sweet Water Development Company during the course of USD1-OSW contracts 14-01-0001-341 and 14-01-0001-1238, nor are any such applications contemplated as a result of the work under these contracts.
COST ESTIMATE

1,000,000 GPD PLANT

A cost estimate for a 1,000,000 gpd plant follows. This estimate was derived from pilot plant knowledge with a considered extrapolation to an actual large scale plant. This extrapolation was obtained from information and knowledge from manufacturers, engineering consultants, OSW personnel and OSW reports. The derivation of the cost estimate is on the following pages and is preceded by a process flow description for the proposed plant along with the process flow diagram and material balance.

The estimated costs on this 1,000,000 gpd plant are as follows:

<table>
<thead>
<tr>
<th>Description</th>
<th>Amount</th>
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<tr>
<td>Capital Cost - Total</td>
<td>$ 921,700</td>
</tr>
<tr>
<td>This is for a complete plant, including equipment, all service facilities, road, supply and disposal lines, etc., working capital and shakedown tests.</td>
<td></td>
</tr>
<tr>
<td>Capital Cost per daily gallon</td>
<td>$ 0.9217</td>
</tr>
<tr>
<td>Production Cost per thousand gallons Fresh Water</td>
<td>$ 0.6905</td>
</tr>
<tr>
<td>This production cost includes power at $0.007/kwh, labor, maintenance, supplies, administration, and 15 yr straight line depreciation.</td>
<td></td>
</tr>
</tbody>
</table>
A process diagram for a 1,000,000 gpd plant follows. Using this flow sheet as a guide and inlet seawater as a beginning, the flow pattern can be traced. The incoming seawater after being pumped from the supply line through pump P-1 is split into two streams and passes through heat exchangers HE-1 and HE-2. In these heat exchangers, the incoming seawater is counter-currently cooled by the discharge product water and reject brine streams from the plant. After being cooled in these heat exchangers, the seawater streams again join before being fed into the maker.

The hydrate free aqueous feed is intimately mixed with liquid propane in the maker. Part of the propane vaporizes, cooling the mass and removing the latent heat of formation of propane hydrate. The remaining propane reacts with the water to form propane hydrate crystals. The slurry thus formed is removed from the maker through pump P-2 and is fed into the wash column. Brine from the slurry is withdrawn from the wash column, leaving the hydrate as a consolidated porous plug which rises due to hydraulic force against a descending wash water stream. The rising cake is cut up by a harvester and mixed with enough sluice water from pump P-3 to reslurry the hydrate and transport it to the melter.

The reject brine removed from the wash column flows to a reject brine cyclone. The underflow from this cyclone is the reject brine from the system. The overflow from the cyclone is recycled to the maker. The reject brine from the cyclone underflow passes through heat exchanger HE-2 and then to the depropanizer where soluble propane is removed. The dissolved propane is recovered from the discharge of the vacuum pump on the depropanizer and it is compressed to make pressure and introduced into the flow to the suction of the primary compressor.

The hydrate and water slurry from the wash column is fed to the top of the melter where it is directly contacted with propane gas. The conditions in the melter are controlled so that as the slurry and propane gas descend through the packed section, the propane gas condenses, giving up its latent heat to decompose hydrate. The net result of this exchange is fresh water and liquid propane. The bottom of the melter acts as a receiver tank and allows any propane gas not condensed to be withdrawn to the secondary compressor. The fresh water and liquid propane from the melter flow by gravity to the propane water separator. The separation occurs in this tank by gravity. The layered propane floats over a baffle into the propane receiver section and is returned to the maker to begin the crystallization step already explained. The product water is withdrawn from the separator and flows through heat exchanger HE-1 to the depropanizer where soluble propane is removed under vacuum. The recovered propane is compressed to make pressure and introduced into the flow to the suction of the primary compressor.

The propane, since it is not consumed process-wise, is continually being recycled. The pattern of the propane can be either picked up at the melter or maker section. For description purposes, choose the liquid propane
APPENDIX I

supply existing in the separator. The propane from this tank is sparged into the maker where it is intimately contacted from the brine phase. Some of the propane enters the clathrate as the hydrating agent. The remaining propane vaporizes and is drawn into the primary compressor C-1. In C-1 the gas is compressed for maker operating conditions (approximately 60 psig) to melting conditions (approximately 70 psig). In the melter the gas entering contacts the slurried hydrate and most of it condenses and flows as liquid propane to the separator. The propane which does not condense in the melter must be compressed in the secondary compressor C-2 to a pressure such that it will condense against cooling water in the condenser, HE-Condenser. The heat removed by the cooled water in the condenser is the heat that must be rejected to the surroundings to make up for heat leak, work input, and inefficient heat exchange of entering and departing streams from the process. The liquid propane from the condenser is fed back to the propane side of the separator to begin its cycle once again.
APPENDIX I

PROCESS EQUIPMENT DESCRIPTION

Maker

From pilot plant work, acceptable data reported in the literature, and consultant's information, the design bases for the maker are:

1) The plant achieves an overall conversion of 42% by volume of the incoming seawater.

2) The hold time based on total feed (hydrate free) is 10 to 12 minutes.

3) Wash water requirement is to be 5 - 10% of fresh water produced.

4) The maker produces a slurry of 20 wt % bound-water hydrate.

5) The agitation power input to the maker is $1 \times 10^{-2}$ hp/gal.

From these bases the maker was designed as two vertical tanks 18 ft in diameter and 12 ft high. Agitation is by a top entering axial flow turbine which imparts both shear type agitation and liquid turnover. The tank has four vertical baffles spaced 90° apart and extending two feet from the wall toward the tank centerline.

The height chosen allows 10 ft of active liquid volume and 2 ft of straight shell as well as the upper vessel elliptical head as a gas disengaging space. The inside walls of the maker are coated with an epoxy paint.

The installed cost of each of two vessels complete with agitator, epoxy coating, footings, and foundation is $30,800, and the total installed cost is $61,600.

Melter

The melter was designed on the basis of an overall heat transfer coefficient of 10,000 Btu/hr, cu ft, OF and a 2F driving force in a tower 14 ft in diameter and 22 ft seam to seam. The top 18 ft of the column is filled with packing. The bottom 4 ft and the elliptical head act as tank capacity for the sluice water pump, P-3 on the flow sheet, and gas disengaging space.

The flow through the melter is cocurrent with both the gas and liquid flowing down the column. This flow pattern gives the gas-liquid contact advantages
of a dumped packing while the gas pressure drop usually associated with a dumped packing is minimized.

The installed cost of the melter complete with footings, packing, distributor and support plates is $27,600.

Wash Column

The wash columns are designed on the basis of permeability data from the pilot plant and the equations for the operation of a pressurized wash column as set forth in our final report on contract 14-01-0001-341. Using a back pressure of 25 psi/foot of cake, a production rate of 1.090 gpd/sq ft is achieved. The necessary area is obtained in three 20 ft diameter wash columns, each with a 10 ft seam to seam straight section.

The geometry of each column is the same as that of the pilot plant wash column. The feed enters at the bottom and rises until it reaches the screen section in the drain tubes which are mounted on supports at the tank bottom. The cake is scraped inward and removed through a pipe which extends from the top of the cake to the bottom of the vessel. The inside of the column is coated with an epoxy paint.

The installed cost for each wash column including internal piping, epoxy coating, a harvester and footings is $52,500. The total installed cost for the three wash columns is $157,500.

Propane Water Separator

The propane water separator design is based on pilot plant operating data. A hold time of 5 minutes is provided for the water side of the tank where the gravity separation occurs and 2 minutes hold liquid time for the propane side. Both hold times are computed for the tank operating at 75% of its total volume. The tank designed is 10 ft in diameter and 16 ft seam to seam.

The installed cost of the tank including the required internal baffle and footings is $9,200.

Reject Brine Cyclone

The reject brine cyclone is designed for a flow rate of 3,000 gpm and a 4 psi pressure drop. The resulting cyclone has a diameter of 70 in. The 1967 installed cost of a 70 in. diameter cyclone as obtained from Nelson's Cost-imating is $5,300.
APPENDIX I

Feed Heat Exchangers

A total of two heat exchangers is required. These are a product water-seawater feed heat exchanger and a reject brine-seawater feed heat exchanger.

For the feed streams the heat exchanger developed by Colt Industries, Inc. of the stacked aluminum plate type are used. These are designed to handle 78°F inlet seawater and achieve a 28°F approach with the cold process streams. The seawater-product water exchanger will consist of four of these exchangers each selling for $8,532. The total purchase price for the four would, therefore, be $34,128. Vilbrandt and Dryden (Chemical Engineering Plant Design) state that the installation of a heat exchanger ranges from 10 to 20% of the purchase price. Using an average of 15%, the installed cost is

\[(\$34,128)(1.15) = \$39,200.\]

The seawater-reject brine exchangers are of the same type as those above but because of the added heat load of the larger stream the purchase price for each of the 4 units required would be $10,740 or a total of $42,960. Installed, the exchangers cost

\[(\$42,960)(1.15) = \$49,300.\]

Propane Condenser

The propane condenser has a heat load of 9.3 x 10⁶ Btu/hr. Using aluminum tube steel shell design, an overall fouled heat transfer coefficient of 185 Btu/hr, sq ft, F for this service is specified by manufacturers. The resulting required heat transfer area using a 10°F mean temperature difference is 5,050 sq ft. From Nelson's Cost-mating and correcting for the aluminum tube, the 1967 installed cost of the exchanger is $12,000.

Depropanizers

The depropanizer is designed based on a superficial gas velocity of 1 ft/sec. The required area for each depropanizer is 3.4 sq ft. To conserve space and save money, both depropanizers were designed into one shell but completely separated by a vertical partition. Each half of the depropanizer operates independent of the other although both are evacuated by a common vacuum pump. The propane from the discharge of the vacuum pump is compressed to make pressure to prevent propane losses from the plant.

The installed cost of the depropanizer including a distributor plate, support plate and packing is $3,650. The installed cost of the vacuum pump is $2,250. The installed cost of the 2 hp Sears, Roebuck compressor is $250. The total installed cost of the depropanizer system is $6,250.
Primary Compressor

The primary compressor is designed based on a compression ratio of 1.19 and a flow of $9.5 \times 10^5$ lbs/day. The brake hp requirements for these conditions is 850 hp. For this compression a centrifugal compressor - single wheel closed impeller-vertically split case is used. For a two train system with compression occurring in two parallel machines, the installed cost of both machines is $59,500.

Secondary Compressor

The secondary compressor is designed based on a flow of $1.38 \times 10^6$ lb/day and a compression ratio of 2.16. The brake hp requirement is 516 hp. This machine is a centrifugal type with three or four stages in a horizontally split case. The installed cost of this machine is $38,700.

Pumps

The size of the pumps required are computed for the flows shown on the process flow diagram. The pump schedule is as follows. Prices are for 1967 and are obtained from Nelson's Costimating.*

<table>
<thead>
<tr>
<th>Pump</th>
<th>ΔP</th>
<th>Flow</th>
<th>HP</th>
<th>Installed Cost</th>
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</thead>
<tbody>
<tr>
<td>Seawater Feed</td>
<td>45</td>
<td>13,781 lbs/min</td>
<td>58</td>
<td>$3,320</td>
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<tr>
<td>Sluice</td>
<td>100</td>
<td>23,781</td>
<td>221</td>
<td>10,900</td>
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<tr>
<td>Slurry Transfer</td>
<td>110</td>
<td>31,918</td>
<td>328</td>
<td>14,400</td>
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<tr>
<td>Reject Brine</td>
<td>15</td>
<td>7,992</td>
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<td>2,800</td>
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<tr>
<td>Product Water</td>
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<td>2,800</td>
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<tr>
<td>Spare #1</td>
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<td>58</td>
<td>3,320</td>
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<tr>
<td>Spare #2</td>
<td>-</td>
<td>-</td>
<td>328</td>
<td>14,400</td>
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Total Installed Cost $51,940

Say $52,000

*The process pumps all have a spare available in the event of a failure. Spare #1 is a backup pump for either the seawater feed, reject brine, or product water. Spare #2 is a backup pump for either the slurry transfer or sluice pump.
APPENDIX I

Insulation

The insulation area was calculated based on the surface area of the vessels described above. The total area of the process vessels and piping, the latter calculated as 5% of the total tank volume, is 9,960 sq ft. The installed cost of sprayed on foam urethane insulation at $1.00 a sq ft is $10,000.

Propane Storage Tank

The propane storage tank was designed to accommodate 52,000 gallons of propane. This volume is three times the operating inventory of propane in the plant. A spherical tank was chosen to get the maximum storage volume for the least cost. The installed cost includes a refrigeration compressor to maintain the propane at 78°F or lower. Also included are footing and an earthwork dike. The installed cost, including these appurtenances, is $29,750.

General Notes on Vessel Design

1) All of the vessels described above were designed in accordance with Section VIII of the ASME Boiler and Pressure Vessel Code, 1965 edition.

2) A corrosion allowance of 1/8 inch was added to the designed shell thickness in each case.

3) The price for each vessel includes one coat of primer to be applied at the shop.

4) The prices shown for the vessels are all for field erection. The segmented elliptical heads and the vessel skirts will be shop bumped, and the shells will be shop rolled prior to shipment to the field.

5) The prices for all the above pressure vessels were derived from information obtained from Chicago Bridge and Iron Works, Atlanta Branch, and Rockaway Tank Company of Dover, New Jersey.
APPENDIX I

PLANT CAPITAL COST

1,000,000 GPD PLANT

Principal Items of Equipment (PIE) as installed:

<table>
<thead>
<tr>
<th>Item</th>
<th>Installed Cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Maker</td>
<td>$ 61,600</td>
</tr>
<tr>
<td>Melter</td>
<td>27,600</td>
</tr>
<tr>
<td>Wash Columns</td>
<td>157,500</td>
</tr>
<tr>
<td>Propane Water Separator</td>
<td>9,200</td>
</tr>
<tr>
<td>Reject Brine Cyclone</td>
<td>5,300</td>
</tr>
<tr>
<td>Feed Heat Exchangers</td>
<td>88,500</td>
</tr>
<tr>
<td>Propane Condenser</td>
<td>12,000</td>
</tr>
<tr>
<td>Depropanizer</td>
<td>6,150</td>
</tr>
<tr>
<td>Compressors</td>
<td></td>
</tr>
<tr>
<td>Primary</td>
<td>59,500</td>
</tr>
<tr>
<td>Secondary</td>
<td>38,700</td>
</tr>
<tr>
<td>Pumps</td>
<td></td>
</tr>
<tr>
<td>Seawater Feed</td>
<td>3,320</td>
</tr>
<tr>
<td>Slurry Transfer</td>
<td>14,400</td>
</tr>
<tr>
<td>Sluice</td>
<td>10,900</td>
</tr>
<tr>
<td>Reject Brine</td>
<td>2,800</td>
</tr>
<tr>
<td>Product Water</td>
<td>2,800</td>
</tr>
<tr>
<td>Spare</td>
<td>3,320</td>
</tr>
<tr>
<td>Spare</td>
<td>14,400</td>
</tr>
</tbody>
</table>

Sub-total: PIE (Installed) $ 517,990

Services and Auxiliary Costs

<table>
<thead>
<tr>
<th>Item</th>
<th>Cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Insulation</td>
<td>$ 10,000</td>
</tr>
<tr>
<td>Piping</td>
<td>55,700</td>
</tr>
<tr>
<td>Instrumentation</td>
<td>22,300</td>
</tr>
<tr>
<td>Instr. Air Supply</td>
<td>3,000</td>
</tr>
<tr>
<td>Intra-plant Electric System</td>
<td>17,860</td>
</tr>
<tr>
<td>Propane field storage tank</td>
<td>29,750</td>
</tr>
<tr>
<td>Stairs, catwalk @2% (PIE)</td>
<td>10,400</td>
</tr>
</tbody>
</table>

Sub-total: Services 149,010

Total Battery Limits Plant:
(PIE) plus Services = PLANT

$ 667,000
**Site and Improvements Costs**

<table>
<thead>
<tr>
<th>Item</th>
<th>Cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Land and improvements</td>
<td>$15,400</td>
</tr>
<tr>
<td>1 acre land @ $2,000</td>
<td>$2,000</td>
</tr>
<tr>
<td>Level, grade, gravel road</td>
<td>$5,400</td>
</tr>
<tr>
<td>Work shack, toilet</td>
<td>$8,000</td>
</tr>
<tr>
<td>Intake well field</td>
<td>$50,000</td>
</tr>
<tr>
<td>Piping to plant</td>
<td>$4,000</td>
</tr>
<tr>
<td>Brine Disposal line</td>
<td>$6,000</td>
</tr>
</tbody>
</table>

**Sub-total: Site** $75,400

**PLANT and SITE** $742,400

Engineering and Profit @ 15% $111,000

Contingencies @ 5% $37,200

**Total Cost** $890,600

**Indirect Capital Costs**

Consider plant on turn-key basis.

Interest cost during 6 months construction

890,600 x 0.06 x 1/2 x 1/2 $13,350

Acceptance period 60 days

Interest cost: 890,600 x 0.06/6 8,900

Power *: 42,800 KWH/d x 30 x 0.007 9,000

Chemicals: 14,500 gal @ 0.08 1,160

Labor:

Engineer @ 15,000 x 1/6 = $2,500

Operators

4 @ 6,980 x 1/6 = 4,650

Maintenance

1 @ 6,980 x 1/6 = 1,150

Sub-total: Labor $8,300

Overhead @ 20% 1,660

Credit: Water Produced

15d x 1,000 @ 0.75 (11,250)

Sub-total: Construction, Shakedown $31,120

**Plant Cost - Total - Turn Key** $921,720

*Carolina Power and Light Corporation advises that a load of the order of 2,000 kw, unity p.f., with demand charge based on 95% occupancy gives a system power cost of 0.007/kwh.*
APPENDIX I

OPERATING COSTS

1,000,000 GPD PLANT

Operating Costs
Based on one stream day
and 350 operating days
per year

Cost per
Stream Day

(1) Electrical Power
1785 kw x 24 hr/day x $0.007/kwh
$ 299.88

(2) Raw Materials - Propane
Propane loss is taken to be
0.005% of the flowing streams,
or 640 lb/day.

640 lb propane/day x $0.08/gal
propane x 1 gal propane/4.35 lb
propane

11.77

(3) Supplies and Maintenance Materials
0.5% of plant per annum or
0.0015% per day

13.88

(4) Operating Labor
$6,980/man year x \(
\frac{4 \text{ men}}{350}\)

79.77

(5) Maintenance Labor
$6,980/man year x \(
\frac{1 \text{ man}}{350}\)

19.94

Sub-total of Items (4) and (5)

99.71

(6) Payroll Extras
15% of Total of Items (4) and (5)

14.96

Total Essential Operating Costs
Total of Items (1) through (6)

$ 440.20
APPENDIX I

(7) General and Administrative Overhead $ 19.94
20% of Total of Items (4) and (5)

(8) Plant Depreciation 175.10
Fifteen years straight line
$921,720

(9) Taxes and Insurance 55.30
2% of plant per annum or
0.006% per day

Total Operating Cost
Total of Items (1) through (9) $ 690.54

Water Cost = Operating cost/water produced
$690.54/day
103 kgal/day
= $0.69/kgal