Heat Transfer in the Coupled VTE/MSF Test Bed Plant at the Office of Saline Water Test Facility Freeport, Texas

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ABSTRACT

The coupled VTE/MSF seawater desalination plant has been operated over 11,000 hours at hot end temperatures of greater than 250°F. This operation has provided the first information on extended service at high temperatures for the doubly-fluted, enhanced 2-inch O.D., Aluminum-brass (CDA 687) tubing used in the falling film, vertical tube evaporators (VTE).

Fouling of this doubly-fluted VTE tubing has been determined to be strongly dependent on operating temperature level and results from films consisting mainly of the normal oxidation products of the CDA 687 Aluminum-brass in a seawater system containing exposed, mild steel; i.e., oxides of copper and iron. Fouling resistance (FR) as a function of temperature may be expressed as:

$$FR = 1/26685 - 91.84T$$

for the range 160 to 260°F, and the overall heat transfer coefficient of the doubly fluted tubes after 10,000 hours operation (U_f) for the same range as a function of temperature is:

$$U_f = -3273. + 46.74T - .1131T^2.$$

The relation of heat flux (Q/A) to driving force (ΔT) is a valuable tool for desalination plant designers and such a correlation has been developed. It is:

$Q/A = 1/(0.06586 - 0.0019041\Delta T)$ at 250°F brine temperature.

The minor effects of subcooled and superheated feed on heat transfer coefficient of the doubly fluted tubing is reported and indications are that an external feed preheater is not necessary in future design.

No clear correlation could be found between tube performance and liquid loading over the range of 2.2 to 5.1 GPM per tube. Neither could a relation be established for steam quality variations between 90 and 103%.

Preliminary evaluation of the effect of shell-side venting on heat transfer coefficient confirmed the previously recommended 1% of total vapor as a satisfactory venting level.

The use of enhanced tubing in seawater distillation has been demonstrated as a sound and economically attractive process.

It is recommended that the use of enhanced tubing in evaporating-crystallizing equipment be considered by future designers.

NOMENCLATURE

A	Nominal outside tube surface area, sq. ft.
d_i	Net tube inside diameter, in.
FR	Fouling resistance, reciprocal BTU/hr–ft^2–°F
g	Acceleration by gravity, 4.18 × 10^8 ft/hr^2
ΔH	Latent heat, BTU/pound
h_b	Brine boiling film coefficient, BTU/hr–ft^2–°F
h_w	Tube wall coefficient, BTU/hr–ft^2–°F
h_s	Steam condensing film coefficient, BTU/hr–ft^2–°F
k	Thermal conductivity, BTU/hr–ft–°F
L	Tube length, ft.
Q	Heat load, BTU/hr
t	Tube wall thickness, ft.
T	Temperature of evaporating brine, °F
U	Overall heat transfer coefficient, BTU/hr–ft^2–°F
U_c	Overall clean heat transfer coefficient, BTU/hr–ft^2–°F
U_f	Overall fouled heat transfer coefficient, BTU/hr–ft^2–°F
U_t	Overall theoretical heat transfer coefficient, BTU/hr–ft^2–°F
Γ	Liquid loading, lb/hr–ft of nominal outside tube perimeter
ΔT	Temperature difference, condensing minus boiling temperature, °F
ΔT_q	Temperature difference of brine (In–Out of the tube), °F
μ	Viscosity, lb/hr–ft
ρ	Density of liquid, lb/cu. ft.
VTE	Vertical Tube Evaporator–falling film
MSF	Multi-stage Flash Evaporator
HTC	Heat Transfer Coefficient

Conversion to International Metric Equivalents (SI)

From	To	Multiply by
sq. ft	m^2	.0930
$BTU/hr-ft^2-°F$	w/m^2-K	5.67
$BTU/hr-ft^2$	w/m^2	3.152
lb/hr–ft	kg/hr–m	.672
lb/cu. ft	kg/m^3	16.01
BTU	Joules (I)	1054.3

INTRODUCTION

In 1958, Public Law 85–833 was inacted authorizing the construction of several desalination plants for the purpose of demonstrating various seawater conversion processes under actual production conditions. One of these demonstration plants was constructed at Freeport, Texas. This original plant was a 12-effect Vertical Tube Evaporator (VTE), and was put into operation in 1961 under the direction of the Office of Saline Water of the U.S. Department of Interior.

In 1967, the capacity of the Freeport plant was expanded by the addition of a 5-effect module, increasing the plant to a total of 17 effects. This plant was shut down in 1969, and the original 12 effects were dismantled. The 5-effect module was incorporated, as the cold end, into a demonstration plant utilizing an advanced and novel design. A 6-effect VTE was coupled to an 11-stage Multi-Stage Flash plant as the preheater section. Both sections of this coupled plant, the VTE and MSF, utilized enhanced tubing as the heat transfer surfaces. This advanced design plant was put into operation in 1971 and has been operated by the Bechtel Corporation, under contract to OSW, for these last two years.

The current plant has been operated to demonstrate the operability of this advanced design, under actual production conditions, for an extended period. The plant's rated capacity is about 900,000 gallons/day of fresh water. Of this total, approximately 650,000 gallons/day were delivered into the potable water supply system of the city of Freeport, and represented approximately 40% of their total consumption.

The plant was shut down on March 23, 1973, due to a federal budget cut and the Development Program has been terminated after operation of the coupled VTE/MSF plant in excess of 11,000 hours.

PROCESS DESCRIPTION

Figures 1 and 2, Process Flow Diagram, present a schematic of the plant flows, and Figure 3 shows samples of

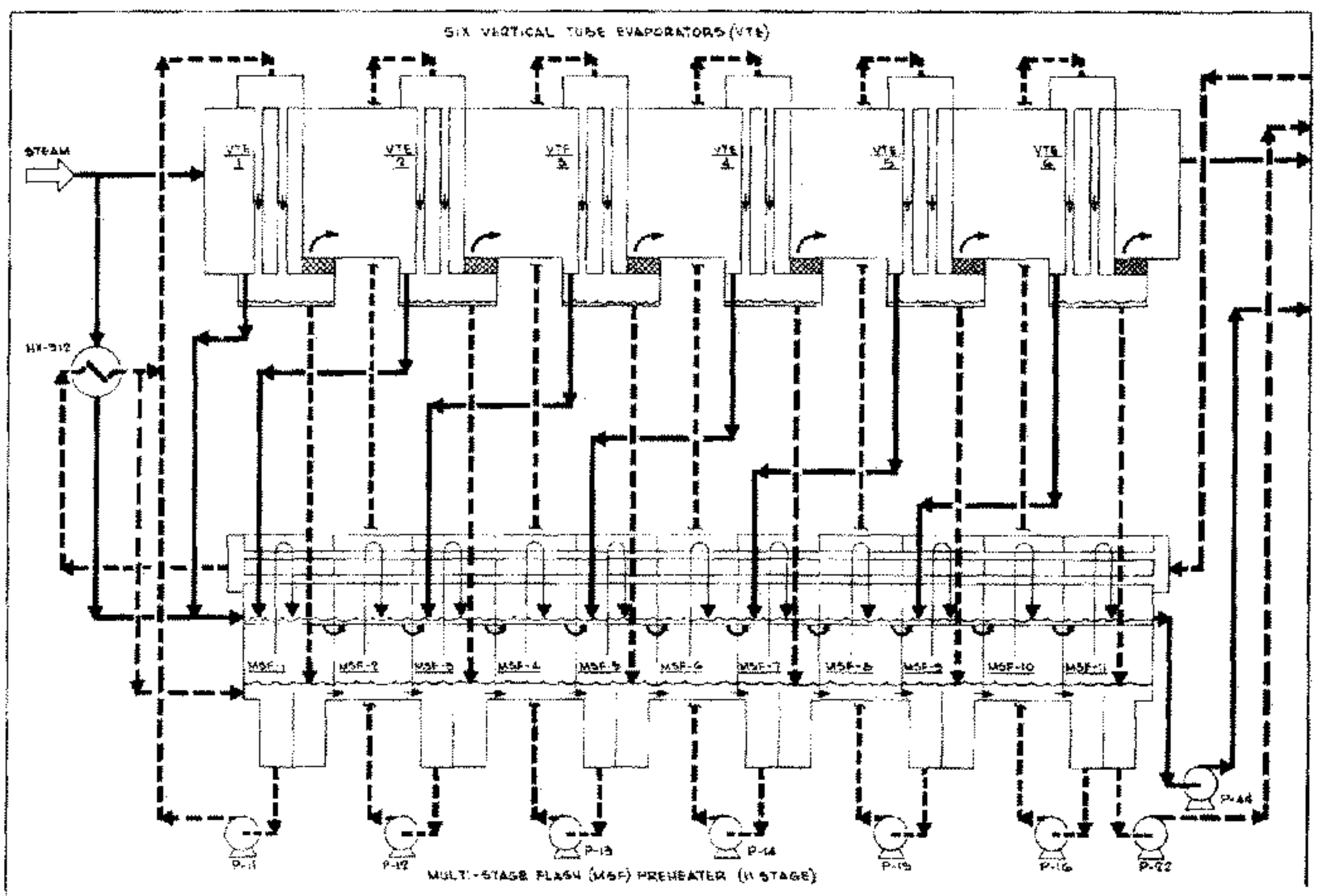

Figure 1. The VTE/MSF Process Flow Diagram of the OSW Freeport Test Bed Plant.

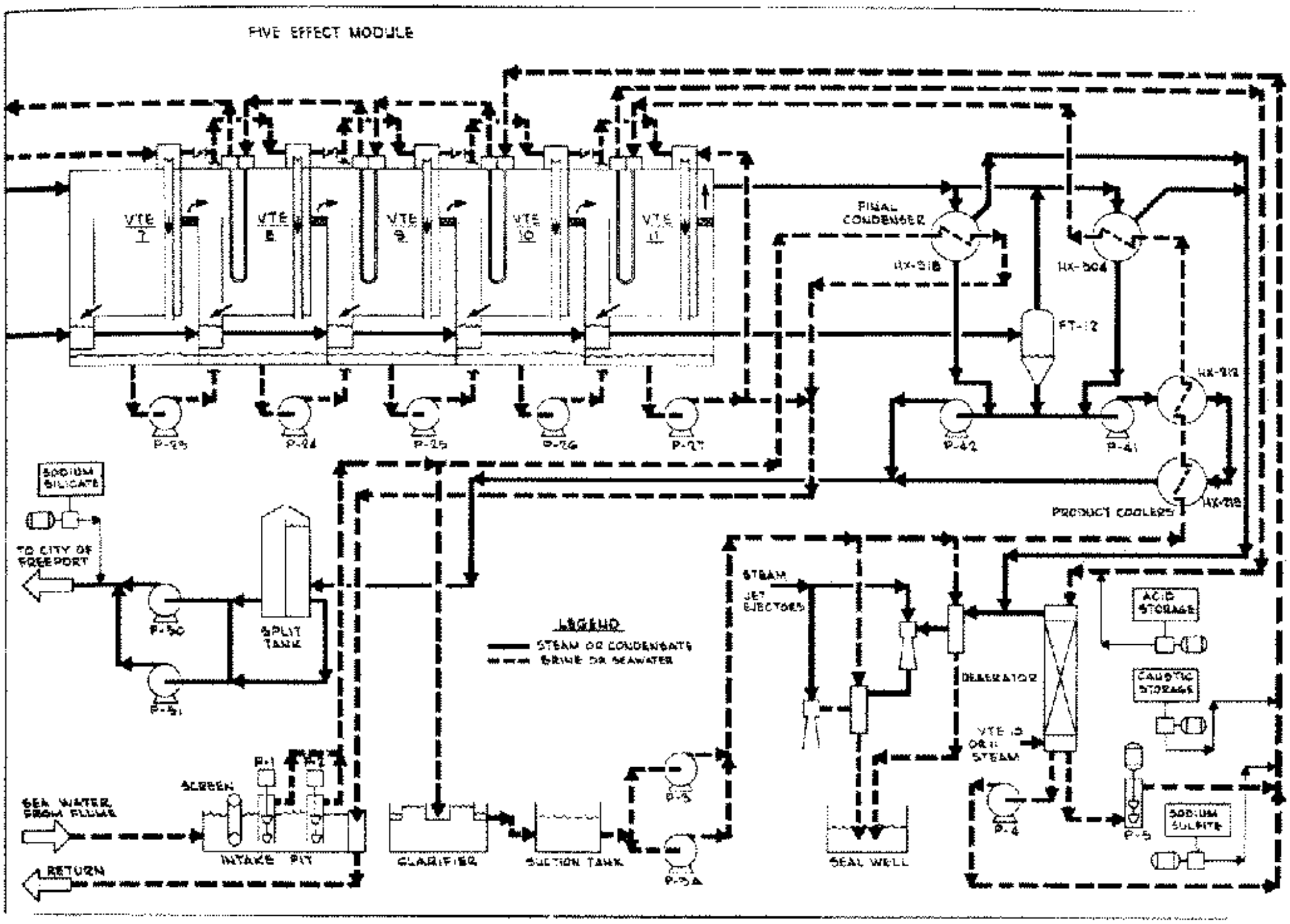

Figure 2. The Five-Effect Module, and Seawater Pretreatment Process Flow Diagram of The OSW Freeport Test Bed Plant.

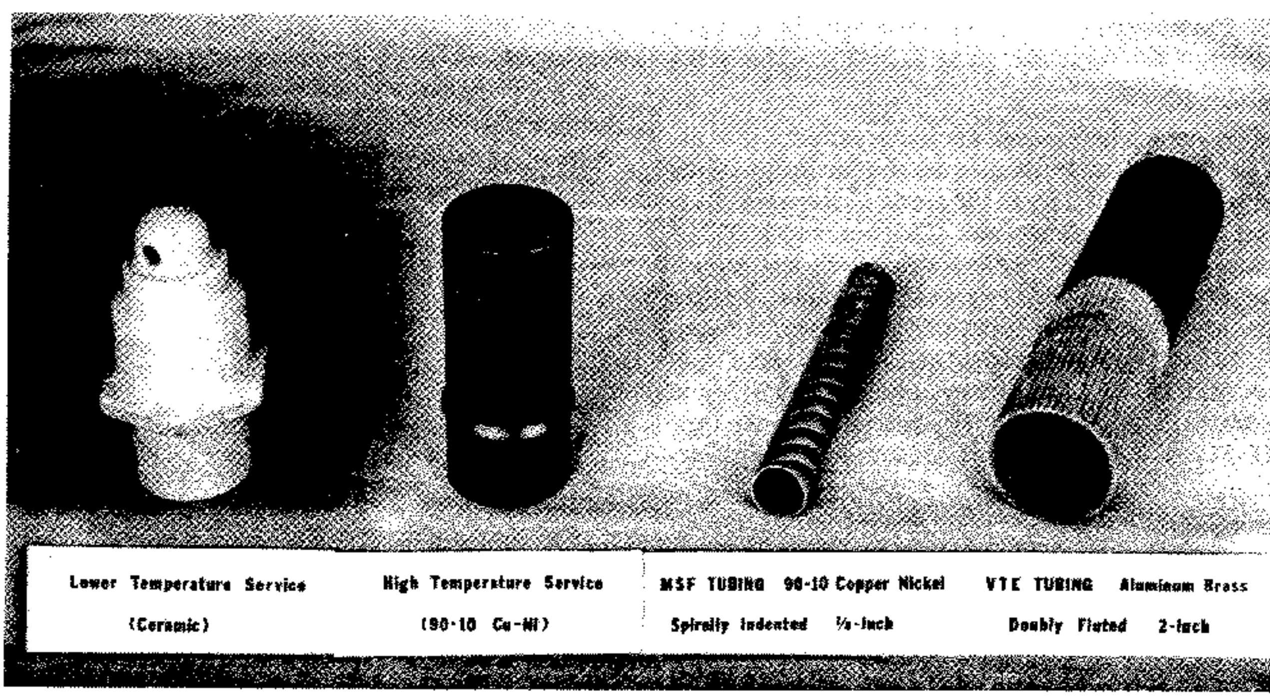

Figure 3. Examples of the Enhanced Tubing and Distribution Nozzles Used in VTE/MSF Section of the Test Bed Plant.

both types of enhanced tubing and the two types of VTE distribution nozzles.

The incoming seawater is warmed to between 100 and 105°F in the product water coolers and passed into the deaerator/decarbonator column after being treated by direct injection of sulfuric acid adjusting the pH to 5.0 to 5.5. This acid treatment neutralizes most of the dissolved alkalinity and the resulting CO_2, the dissolved CO_2, N_2, and O_2 are removed by a vacuum jet system (normally our seawater contains between 2–7 ppm dissolved O_2).

After the deaerator/decarbonator, sodium sulfite is injected into the seawater to scavenge the residual O_2, usually less than 50 ppb, and sodium hydroxide is added to adjust the pH of the feed stream to 6.8–7.0. This feed stream is preheated step-wise, first through "U" tube bundles suspended in the steam chests of the old module, and then through the tube bundle of the MSF section of the coupled module. This bundle consists of 152 tubes, 7/8-inch diameter, CDA 706 (90–10 Cu-Ni), 75-feet long. The tubing is enhanced by spiral indentation and serves as the condenser for the generated vapor in the MSF section as the seawater feed is preheated from about 175 to 235°F. An external feed preheater (HX-312) is used to raise the feed stream to the desired operating temperature, usually 250 to 255°F although operation was successful as high as 265°F without problems.

Now that the feed stream is preheated to operating temperature, three feed modes are possible.

1. All the feed may be introduced into the first MSF stage (MSF-1) from where a portion is pumped as the feed to the first VTE effect (VTE-1) or,
2. All the feed may be introduced to the VTE-1 or,
3. The feed stream can be split with a portion going to VTE-1 and the remainder going to MSF-1. (This is the preferred method.)

In any of the feed modes, the feed is introduced to the VTE tubes through distribution nozzles that distribute the feed brine as a thin film to the inside of the tubes. Each of the six VTE effects contains 172 tubes, 2-inch O.D. X 10-feet long, CDA 687 (Aluminum-brass), and are enhanced by double fluting. Prime steam, introduced on the shell-side of VTE-1, induces a portion of the feed to evaporate. In the VTE bottom sump, this generated vapor separates from the brine and passes through a demister to condense on the VTE-2 tubing, generating vapor from VTE-2 brine, etc. as in a conventional multieffect evaporator.

The brine from the VTE-1 flows by gravity to the MSF-1 where it joins the MSF-1 feed stream. This combined stream is flashed through MSF-2 and MSF-3 to the temperature of VTE-2. A portion of this MSF-3 brine is pumped as feed to VTE-2, etc. with two stages of MSF flash being taken between each VTE.

The condensate from the VTE shell-side is collected, along with that generated in the MSF, in the product water trough section of the MSF. This product water stream is also flash-cooled through the MSF, giving up its heat to preheat the incoming seawater.

PLANT DESIGN DATA

Tables I and II summarize some of the plant's design bases and rated capacity.

Full details of the operation, test results, and data analyses for the last year's operation of the Freeport, Texas VTE/MSF Seawater Distillation Test Bed Plant can be found in the Annual Report (Bechtel, 1973). Furthermore, a summary of the operating results of the VTE and MSF are described in a paper by Houle and Buhrig (Houle, 1973). Following are some data and correlations

TABLE I

Plant design data

	VTE
No. of Effects	Six
Tube Bundle	172 tubes, 10 ft. long, 2 inch nominal OD, doubly fluted, CDA 687 (Aluminum Brass) with flow distributors for falling-film operation
Heat Transfer Area	5,382 sq. ft. (smooth tube basis)
Materials of Construction:	
Brine Service	CDA 706 (90—10 Cu—Ni) lined steel Concrete lined steel, Phenolic lined steel, Glass-reinforced epoxy.
Distillate Service	Epoxy coated steel, steel
	MSF
No. of Stages	Eleven
Tube Bundle	152 tubes, 75 ft. long, 7/8 inch nominal OD, spirally indented "rope" configuration, CDA 706 (90—10 Cu—Ni)
Condensing Surface	2,562 sq. ft. (smooth tube basis)
Materials of Construction:	
Brine Service	Concrete lined steel, Phenolic lined steel, Glass-reinforced epoxy
Distillate Service	Steel, Glass-reinforced epoxy

5-Effect VTE Module

Effect No.	7	8	9	10	11
No. of Tubes	150	369	327	260	222
Length, ft.	11.08	22	22	22	22
Type	Dbl. Fluted	Smooth	Smooth	Smooth	Smooth
Material	CDA 194[1]	316 SS	CDA 687[2]	CDA 706[3]	CDA 706[3]
Smooth Tube OD inch	3-1/8	2	2	2-1/2	3
Smooth Tube Area sq. ft.	1722	4136	3767	3600	3835

[1] 2.6% Fe, Bal. Cu
[2] Aluminum Brass
[3] 90—10 Cu—Ni

TABLE II

Overall plant design operating conditions

Capacity (gross)	310,100 lb/hr (896,000 gal/day)
Capacity (net)	281,200 lb/hr (812,000 gal/day)
Seawater feed	473,800 lb/hr
Brine blowdown	189,400 lb/hr
Concentration ratio	2.5 lb feed/lb blowdown
Total seawater flow	1,420,000 lb/hr
Raw steam supply (including ejector steam)	29,300 lb/hr (150 psig and 550°F)
First effect condensing temp	275°F
Heat rejection condensing temperature	110°F (seawater temp 75°F)
Net economy ratio	9.4 lb product/1000 BTU heat input
Feed treatment	Acidification with H_2SO_4

of particular interest to evaporator designers who might consider future use of enhanced tubing.

FOULING

The term "fouling" may be defined as the heat transfer resistance which results from any undesirable deposit on heat transfer surfaces, during the normal operation of a desalination plant.

The effect of fouling resistance (FR) on design of the VTE bundle is defined by the fundamental equation:

$$\frac{1}{U_f} = \frac{1}{U_c} + FR \qquad (1)$$

which relates the overall heat transfer coefficient U_f of tubes which have been in service for a long time to the performance U_c of the same bundle when the heat transfer surfaces are new.

Fouling resistance, as used in design equations, is a very important economic factor and affects the initial equipment cost, operating cost, and periodic cleaning cost.

There has been very little data published from which fouling resistance for seawater distillation could be calculated or the fouling vs time curve obtained, i.e., the duration of operation before an asymptotic or stable value will be reached.

The VTE/MSF Test Bed Plant at Freeport, having over 10,000 hours on-stream operation time, is the only plant capable of providing information about fouling of doubly fluted tubes in prolonged operation at temperatures over 220°F. The fouling resistance data previously available has been obtained from only single tube tests or from OSW pilot plants and the VTE-X module at the OSW San Diego Test Facility (Distillation Digest 4, 1971, p. 283). The only information on long-term fouling of Aluminum-brass tubes at high temperatures has been based on tests of single tubes from the second effect of the old VTE Freeport Test Bed Plant. The history of the overall heat transfer coefficient for the old 12-effect VTE Freeport plant is lost as the overall coefficients were not available for the plant operational period from 1961 to 1964.

While the fouling process of doubly fluted tubes is obviously a time function, the magnitude of fouling is related to the operating parameters of the plant and construction materials.

Table III presents the overall heat transfer coefficients for each of the six effects of the doubly fluted, CDA Alloy 687 (Aluminum-brass) tube bundles. These coefficients represent values of prolonged "10,000 Hours" performance at the typical operating brine temperatures. These temperatures do not represent any specific runs, but are an average of operating temperatures through all the history of a given effect.

The low overall heat transfer coefficient (U_f) is a result of fouling during 10,000 hours of on-stream operation, and as can be seen in Figure 4 is strongly dependent on the operating temperature. For comparison, Figure 4 presents also, clean heat transfer coefficient (U_c) obtained after retubing of the first effect (VTE-1) with new Aluminum-brass bundle. In addition, for comparison, Figure 4 includes the plot of VTE-X heat transfer coefficient vs brine temperature. The data for VTE-X second effect Alu-

TABLE III

"10,000 Hour" HTC's in VTE's

VTE's	1	2	3	4	5	6
Temperature, T, °F	251.5	235.2	220.9	206.7	191.0	173.8
U_f, BTU/hr-ft^2-°F	1327	1450	1580	1642	1503	1450

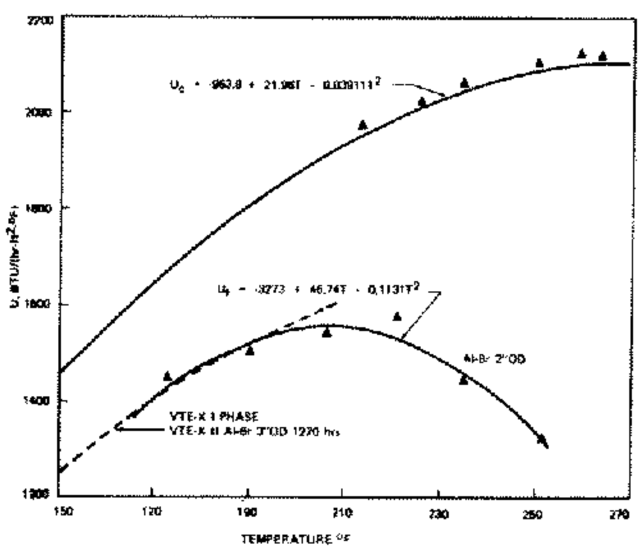

Figure 4. The "10,000 Hours" Heat Transfer Coefficient (U_f) of the Aluminum-brass, 2-inch O.D., Doubly Fluted Tubes as a Function of Operating Temperature.

minum-brass, 3-inch O.D. tubes are results of Phase I VTE-X Development Program (Burns & Roe, 1971).

The remarkable similarity of results for the common temperature range between 160 and 215°F is a confirmation of our data. However, because the VTE-X operated only 1270 hours in Phase I, Phase II results show an increase in fouling resistance, partially because the second effect of VTE-X operated some of its time in the upper temperature range.

With the help of multiple regression analysis, we correlated the results for clean and fouled bundles. The overall heat transfer coefficient U_f for 2-inch, Aluminum-brass tubes is given by the following equation:

$$U_f = -3273. + 46.74T - .1131T^2 \quad (2)$$

for operating temperature range from 160°F to 260°F. The clean tube equation for U_c vs brine temperature is shown for the new tubes which have been naturally oxidized in an ambient atmosphere, and therefore, represents the performance of commercially clean bundles.

$$U_c - -953.9 + 21.96T - .03911T^2 \quad (3)$$

From equation (1), fouling resistance FR data is obtained by solving equation (3) for temperatures from Table III and substituting the actual U_f values. Table IV presents fouling resistance for each of the six effects as a result of 10,000 hours service at the typical operating brine temperature.

TABLE IV
Fouling Resistance (FR) in "10,000 Hours Service"

VTE's	1	2	3	4	5	6
Temperature, T, °F	251.5	235.2	220.9	206.7	191.0	173.8
FR x 10^{+4} hr-ft^2-°F/BTU	2.7707	2.0213	1.3039	1.2685	1.0973	0.9375

The general relation of the fouling resistance with the brine operating temperature is correlated by the following hyperbolic function:

$$FR = 1/26685 - 91.84T \quad (4)$$

valid for the tested temperature range 160°F to 260°F. The hyperbolic function (4) results from a least-squares fit of Table IV data by its linear transform. The above relation is shown graphically as Figure 5.

It must be remembered that fouling resistance is relative to commercially clean tubes and not related to theoretical performance of doubly fluted tubes, and that it is based on nominal area.

It is apparent, from the foregoing, that considerable fouling resistance is encountered in the operation of doubly fluted tubes in the normal continuous seawater distilla-

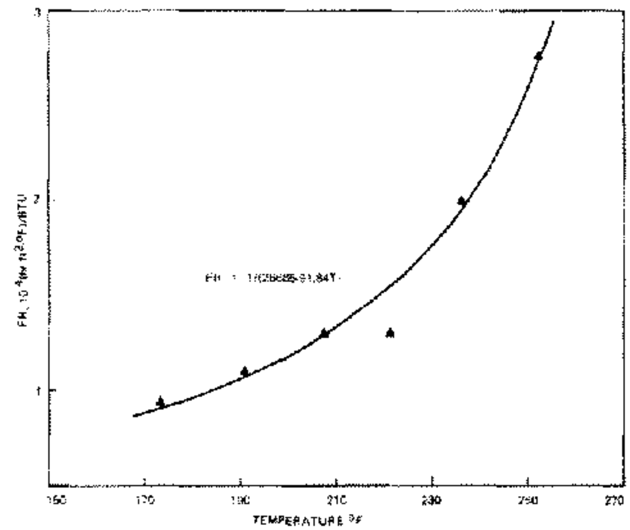

Figure 5. The "10,000 Hours" Fouling Resistance Versus Temperature, Based on Comparison with Commercially New Aluminum-brass Tube Bundle Performance.

tion process. The fouling films on both the tube O.D. and I.D. were analyzed and the results are as follows:

1. Cuprite Cu_2O and cupric oxide CuO are the chief constituents in the range of 40–90%.
2. Hematite Fe_2O_3 and magnetite Fe_3O_4 in the range of 5–30%.
3. Small amounts of ZnO (a normal corrosion product of the Aluminum-brass tubes) and $Al_2O_3 - SiO_2$ (as mud or silt trapped on the rough tube I.D. surface).

These conclusions were based on analyses of the film using one or more of the following methods:

1. DTA
2. Emission Spectrography
3. Energy Dispersion X-Ray
4. X-Ray Diffraction
5. X-Ray Fluorescence

The origin of the fouling resistance is from oxidation of the heat transfer surfaces, typical to the Aluminum-brass alloy, and oxidation products of other construction materials.

Most of iron oxide fouling on the outside of the doubly fluted tubes is evidently a result of the inability of the "Amercoat" coating (peeling) to protect the steam and vapor chests, and from other unprotected steel steam and vapor lines.

The fouling on the tube I.D. is primarily a result of oxidation and corrosion of Aluminum-brass material. A lesser cause of fouling may be attributed to the suspended debris entering the plant in the seawater feed.

It must be remembered that this high fouling resistance occurred despite the fact that most of the carbon steel in the plant was protected from brine with concrete lining or other protective coatings. Moreover, the plant operated with proper scale control systems, and fouling due to alkaline scale or calcium sulfate scale was negligible and that scale was eliminated as a possible cause of the high fouling rates, and finally that venting, deaeration, and decarbonation was comparable to any industrial plant.

The most important operating parameter which has to be considered as an integral part of the fouling resistance, is the operating temperature. Several series of development runs have been performed to demonstrate this effect once more.

Figure 6 shows a family of curves expressing the relation of heat transfer coefficients versus operating temperature separately for each effect.

These data have been obtained by multiple scanning over a period of one or two days to eliminate the operating time effect. Therefore, the change in heat transfer coefficients is a direct result of physical property changes in the boiling and condensing film.

Inspection of the curves show that the heat transfer coefficient for the sixth effect (VTE-6) is better for the corresponding temperature range than for the performance of the fifth effect (VTE-5). Furthermore, the performance of the fifth effect is better than the performance of the fourth effect (VTE-4) . . . and so on.

For example, if we could operate, for a short period of time, all six effects each at a temperature of 250°F, the resulting performance would be the highest for VTE-6, and descend in order for VTE-5, VTE-4, VTE-3, VTE-2,

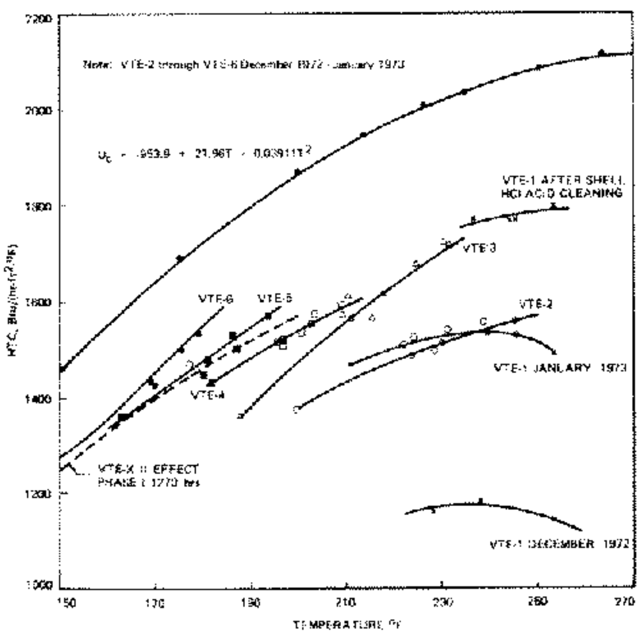

Figure 6. Heat Transfer Coefficient as a Function of Evaporating Brine Temperature Showing The Relation For Each of the Six Effects, (2-inch, O.D., Aluminum-brass Doubly Fluted Tubes).

and finally VTE-1. The FR, conversely is the highest for VTE-1 and diminishes towards VTE-6.

Conclusions

1. The primary contributors to fouling resistance are oxidation and corrosion products.

2. Fouling resistance is not a constant value and is dependent on temperatures at which oxidative fouling occurs.

3. The temperature at which a given tube bundle operates has the strongest effect on the value of the fouling resistance.

4. A total fouling resistance of .00009 to .000275 should be used for Aluminum-brass tubes to seawater at temperatures from 170 to 250°F (Figure 5), to guarantee long term (10,000 hours) favorable results. The fouling resistance values are based on comparison with commercially clean tubes.

5. The shell-side fouling resistance is greatly responsible for the overall fouling and is, at least, as large as the tube-side fouling.

6. There are likely asymptotic values for fouling resistance, but the resulting low overall heat transfer coefficient requires, from the operating standpoint, cleaning of the heat transfer surfaces before the stable values can be reached.

Recommendations

It is recommended that in the future development of the VTE/MSF process, the effect of operating temperature on the oxidation fouling be considered serious enough to warrant development of new preventative techniques of on-stream "vapor-phase cleaning" and/or "passivating" the heat transfer surfaces.

Systematically designed experiments, which would isolate effects of known parameters, should be included in test plants such as Orange County. Otherwise, continuous changing conditions from high to low temperature will mask the actual results of fouling.

TEMPERATURE EFFECT ON HEAT TRANSFER COEFFICIENT

It is generally recognized that the overall heat transfer coefficient (HTC) of doubly fluted tubes is affected by temperature level. The predictive equations for both the condensing steam-side and boiling brine-side coefficients are directly proportional to the property grouping $(k^3\rho^2g/\mu^2)^{1/3}$. As could be expected, the overall heat transfer coefficient for clean tubes increases with increased operating temperature. This effect is more pronounced for low temperature ranges, below 220°F, where the change in physical properties, particularly in viscosity, is significant. Several tests were performed to determine the effect of temperature on the overall heat transfer coefficient. By controlling pressure of live steam to the first effect, we were able to vary the evaporating temperature. The boiling temperature for the first effect (VTE-1) varied from 265 to 210°F and corresponding evaporating temperatures for the sixth effect (VTE-6) varied from 185 to 150°F.

The results of these runs are shown in Figure 7. These runs were performed after retubing of the VTE-1 bundle, and after an attempt was made to improve the heat transfer coefficients of the remaining effects with carbon dioxide vapor phase cleaning. The family of curves represents the performance of each of the six effects, and the relative position of the curves depends on the fouling resistance of any given effect. The same general effect of temperature is shown in Figure 6, where the heat transfer coefficients are lower due to the higher fouling resistance.

It is interesting to note the effect of temperature on heat transfer coefficient of VTE-1 for runs performed in December, 1972 and January, 1973. The first effect bundle with considerable fouling resistance, shows maximum value of heat transfer coefficient with temperatures. This is an indication that fouling affects heat transfer in ways other than as a thermal insulating layer.

The effect of temperature on the overall heat transfer coefficient of commercially new tubes can be expressed by the empirical equation:

$$U_c = -953.9 + 21.96T - .03911T^2 \tag{5}$$

where T is evaporating brine temperature in °F.

The following equations are recommended by F.C. Standiford 4 (1971, p. 417) for film coefficients, based on the nominal tube diameter.

$$\text{Steam, } h_s = 1840\,(k_\rho^3{}^2 g/\mu^2)^{1/3}\,(\mu\Delta HL)^{1/3}(Q/A)^{-.833} \tag{6}$$

$$\text{Tube Wall, } h_w = 1.273\ k/T \tag{7}$$

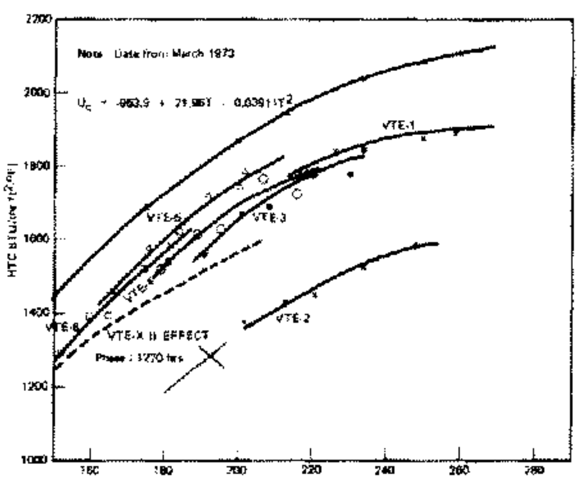

Figure 7. Heat Transfer Coefficients as a Function of Evaporating Brine Temperature Obtained after Retubing the First Effect and Vapor Phase Cleaning the Remaining Five Effects with *in-situ* Generated CO_2.

$$\text{Brine Evaporating, } h_b = .040\,(k^3\rho^2 g/\mu^2)^{1/3}(4\Gamma/\mu)^{1/3} \tag{8}$$

The brine film coefficient "fit" better with the results of the VTE-X by the use of a new constant as suggested by R. Khan:

$$h_b = 0.256\,(k^3\rho^2 g/\mu^2)^{1/3} \tag{9}$$

In this case, the best fit with actual data for commercially clean tubes can be achieved with a constant 0.0176 in the expression of the boiling film coefficient.

Table V shows the effect of brine temperature and other parameters on the actual overall heat transfer coefficient, and compares the results obtained from recommended equations. The suggested correlation for steam-side condensing coefficient, h_s, is in our opinion too strong a function of the flux $(Q/A)^{-.833}$. In the derivation of equation (6), Standiford used the experimental results, described by Thomas (1967), and combined these results with the "practical" Nusselt relation for smooth tubes. Thomas' experimental results have been obtained for a considerably higher range of fluxes from 2×10^4 to 10^5 BTU/hr-ft²-°F, while typical fluxes expected for the distillation plant are 1.0×10^4 to 3.0×10^4 BTU/hr-ft²-°F. Moreover, Thomas' data was obtained with forced circulation cooling, and therefore, with varying fluxes along the height of the tube.

Equations (8) and (9) for the boiling brine coefficient, H_b, have been developed from:

1. Oak Ridge National Laboratory
2. Houston Research Institute
3. VTE-X Experimental Module, San Diego, California data. In our opinion, h_b, should be corrected to reflect the increased boiling heat transfer coefficient with an increase in ΔT, where ΔT is the temperature difference between condensing steam and boiling brine temperature. This relation is discussed in the following section.

In general, all of the "theoretically" recommended equations do not represent the actual performance, and in the absence of a real model, an empirical equation such as (5) should be used to predict practical results of clean doubly fluted tubes.

HEAT FLUX VERSUS ΔT

The relation between heat flux, Q/A and ΔT, the temperature difference between the steam chest and the brine sump, is a very important factor in the design of VTE/MSF plants.

The designer of desalination plants, having specified overall economy has to decide on the most economical distribution of total available temperature difference. He can design with equal ΔT and a different heat transfer surface (HTS) per each effect, or attempt to design with equal HTS and varying ΔT. With the selection of ΔT, the maximum value of flux for a given temperature level is established, and the relation between ΔT and flux (Q/A)

TABLE V

Effect of various parameters on actual overall heat transfer coefficient

	Run No. 4-14	Run No. 4-13	Run No. 4-12	Run No. 4-10	Run No. 4-17
Evaporating brine temp. VTE–1, °F	263.3	259.5	234.3	226.1	213.5
Condensing steam temp. VTE–1, °F	280.6	276.4	247.8	239.2	225.0
Heat Flux, BTU/hr X ft^2	32970	32000	25080	24080	20300
Liquid Loading, lb/hr X ft	3890	3900	3930	3930	3970
Concentration Factor	0.580	0.638	0.696	0.638	0.638
Actual Coefficient, BTU/hr X ft^2 X °F	1909	1890	1858	1840	1781
Steam Side Property, $(k^3 \rho^2 g/\mu^2)^{1/3}$ BTU/hr–ft^2–°F	7270	7150	6560	6320	6120
Brine Side Property, $(k^3 \rho^2 g/\mu^2)^{1/3}$ BTU/hr–ft^2–°F	6750	6680	6100	5910	5660
Modified Reynolds Number	28810	27370	24180	23120	21400
$Re^{1/3}$	30.66	30.19	28.92	28.48	27.78
Condensing Coefficient, h_s, BTU/hr X ft^2 X °F	8130	8330	9900	10250	11230
Boiling Coefficient: h_b by Standiford	8280	8050	7060	6730	6290
h_b by Khan, BTU/hr X ft^2 X °F	5300	5150	4520	4310	4020
Tube Wall Resistance, hr X ft^2 X °F/BTU	.000064	.000064	.000064	.000064	.000064
Theoretical Clean U_T: U_T by Standiford	3250	3240	3260	3220	3200
U_T by Kahn	2660	2640	2590	2540	2490
U_C by Bechtel	2119	2096	2058	2036	1964
hr X ft^2 X °F/BTU Used to obtain actual FR by Standiford	0.000216	0.000221	0.000231	0.000233	0.000249
FR by Kahn	0.000148	0.000151	0.000152	0.000150	0.000160
FR by Bechtel	0.0000519	0.0000520	0.0000523	0.0000524	0.0000523

is of great interest. Such a relation has been developed, and is shown as Figure 8. Data for these curves were obtained from test results for the first effect bundle (VTE-1), and should be of value to designers of future desalination plants.

All of these data have been correlated by a least-squares fit resulting in the following hyperbolic type equation:

$$Q/A = 1/(0.06586 - 0.0019041\Delta T) \qquad (10)$$

The index of determination of this function is 0.9865, and the percent difference in correlation has a maximum range of −3.3 to 3.0.

An important conclusion of this correlation is that the boiling and condensing mechanisms are independent of feed brine quality in the range of fluxes from 2.2×10^4 to 3.6×10^4 BTU/hr-ft^2-°F, and the brine-side forced convection vaporization mechanism does not change instantaneously into nucleate boiling. This correlation (10) has been used to establish the relation between the overall heat transfer coefficient and ΔT. The results are expressed by the following equation, and are shown in Figure 9.

$$U = \text{Flux}/\Delta T = Q/A\Delta T = 1/(.06586\Delta T - .0019041\,\Delta T^2) \qquad (11)$$

Inspection of Figure 9 reveals a very interesting fact. The overall heat transfer coefficient declines with an increased temperature differential. At a certain point, any further increase in temperature difference across the tube causes partial restoration of higher heat transfer performance. It may be concluded that higher ΔT's reduce the condensing coefficient and increase the boiling coefficient. At a certain point, the improvement in the boiling coefficient begins to exert a greater influence than the declining condensing-side coefficient on the overall performance.

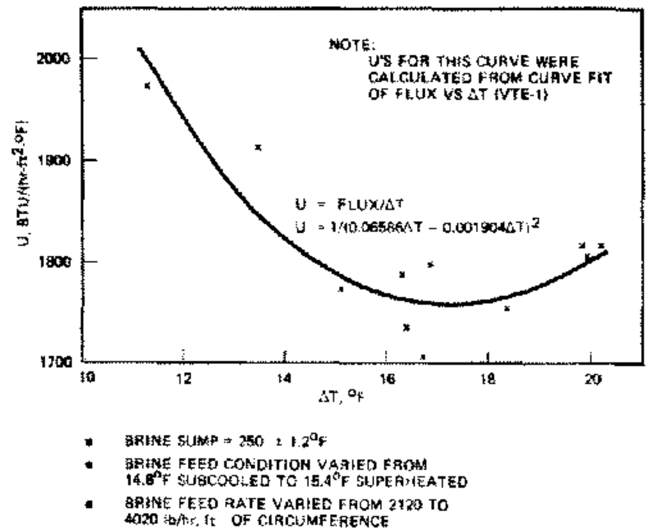

Figure 9. Heat Transfer Coefficients as Function of Overall Temperature Difference for the 2-inch, Aluminum-brass, Doubly Fluted Tubes.

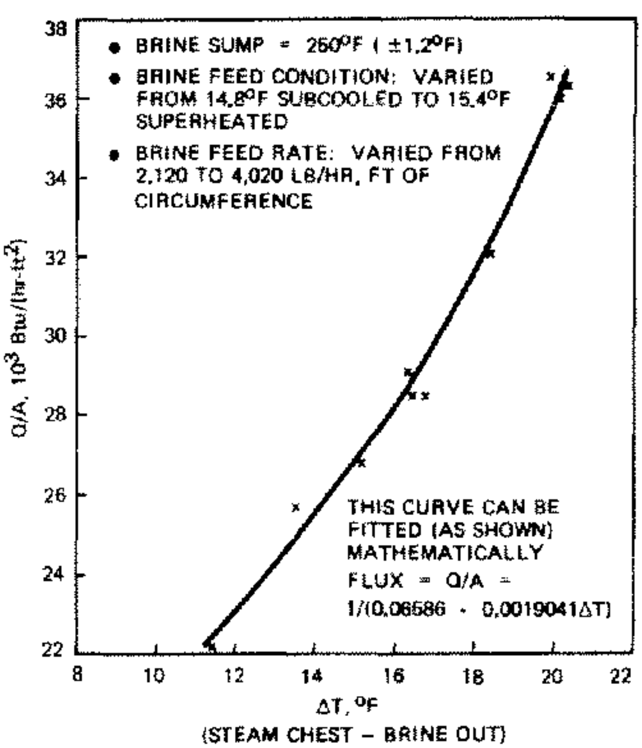

Figure 8. The Relation of Flux with the Overall Temperature Difference Between Steam and Evaporating Brine in VTE-1.

FEED QUALITY, SUBCOOLING—SUPERHEAT

The effect of feeding subcooled or superheated brine to the VTE-1 tubes is shown in Figure 10. These data were obtained by adjusting the quality of feed with adjustment of the brine heater HX-312 steam pressure. The VTE-1 tube inlet temperature was superheated by 15.4°F as compared with the temperature of the VTE-1 brine sump, and then reduced to subcooled condition in several steps. Maximum subcooling was achieved by the complete elimination of preheating in the brine heater. At that condition, the feed was subcooled by 14.8°F.

In contrast to past assumptions that the quality of feed (1) does not affect the doubly fluted tube performance, or (2) only reduces the heat transfer coefficient with increased subcooling, the results shown in Figure 10 demonstrate that supersaturation improves overall heat transfer, partly as a result of an increase in the shear and turbulence at the top of the doubly fluted tubes on the evaporating-side, and partly, as a result of reduced fluxes on the steam-side. When the feed is subcooled prior to its introduction to the tubes, the heat transfer coefficient at some level of subcooling also starts to improve over conditions surrounding the saturation point.

This phenomena can be explained in terms of the in-

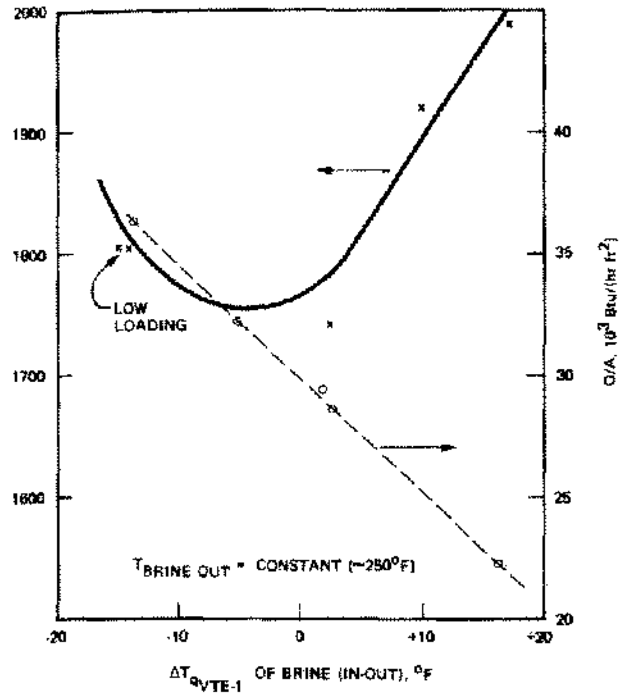

Figure 10. The Heat Transfer Coefficient and Flux Relation with the Brine Quality Entering the Tubes; Tested in Subcooled and Superheated Feed Conditions for VTE-1.

crease in temperature difference across the tubes, and consequently, in the higher coefficient of the boiling heat transfer. In this instance, although the condensing coefficient is reduced because of increased fluxes, the increase in the boiling-side coefficient is considerably greater, resulting in an overall increase in performance.

The balance of both the brine and the steam-side coefficients determine the exact location of a minimum heat transfer coefficient as a function of feed quality; a point not necessarily at saturated feed.

We have suspected the possibility of brine feed loading having a major impact on the improvement in the boiling heat transfer coefficient with a simultaneous increase in subcooling. Therefore, we performed special runs at a reduced loading rate.

The brine feed rate per tube was reduced from 3910 lb/hr-ft to 2120 lb/hr-ft of nominal circumference, and the subcooling rate was kept at −14° and −14.8°F levels. The resulting overall heat transfer coefficient changed only from 1830 to 1803 BTU/hr-ft^2-°F, as is shown in Figure 10; not a significant change.

ΔT used in heat transfer calculations has been defined as the condensing-side temperature less the sump evaporating brine temperature. Regardless of feed quality, no correction in ΔT has been applied. All of the above conclusions have been based on actual performance of the first effect VTE-1 with live steam pressure being adjusted each time the quality of the feed was changed in order to keep the boiling temperature constant at 250°F.

We have concluded that an increase in subcooling increases the overall ΔT and fluxes for the existing tube bundles as has been shown in Figures 8 and 9. To avoid misunderstanding, this does not mean, however, that if a bundle is designed for saturated brine and a large ΔT, the performance will be significantly different.

In other words, no matter what the quality of feed is, the amount of heat transferred by the tube will be basically dependent on the overall temperature difference.

The practical importance of these conclusions is related to the overall design philosophy.

If, with fluxes of 34–38 X 10^3 BTU/hr-ft^2-°F and subcooling of 14–15°F, the first effect operation is stable, the necessity of designing the brine heater is eliminated by the 4–5% increase in overall heat transfer coefficient.

On the other hand, if we utilize the advantage of superheated feed, achieving 8–10% improvement in the overall VTE heat transfer coefficient with 10°F of superheat, the concept of combining the VTE with MSF can be reevaluated. That is, in a normal VTE/MSF plant, two or three stages of MSF are coupled for each VTE effect to reduce the thermodynamic irreversibility in preheating incoming feed, but savings realized in reducing heat transfer surface of doubly fluted tubes due to utilization of superheated feed offset the cost of preheating in a conventional VTE plant.

BRINE LOADING

The effect of liquid loading, or feed rate, on the performance of doubly fluted tubes is part of the overall design consideration since the selection of liquid loading relates directly to pumping power. For this reason, the effect of liquid loading on overall heat transfer coefficient was studied and the results are shown as Figure 11 and 12.

Equations (8) and (9) for the tube-side boiling film coefficient are proportional to a modified Reynolds number to the .333 power. The Reynolds number is $Re = 4\Gamma/\mu$, where Γ is the liquid loading in pounds per foot of wetted perimeter of the nominal outside tube circumference, and μ is the viscosity of the brine, lb/hr X ft.

VTE-1

The effect of liquid loading on the overall heat transfer coefficient is shown in Figure 11. For the first effect, the variation in liquid loading from 2200 to 5100 lb/hr X ft, which corresponds approximately 2.2 to 5.1 GPM per tube, did not produce the expected increase in heat transfer coefficients.

The lowest liquid loading rate is limited by the minimum wetting rate and the danger of overconcentration (scaling). The upper limit is controlled by flooding the tube, and an increase in the tube-side pressure drop.

The past circulation rate limit of about 4000 lb/hr X ft for VTE-1 was overcome by modification in the plant design, i.e., raising of VTE-1 demisters and installing an

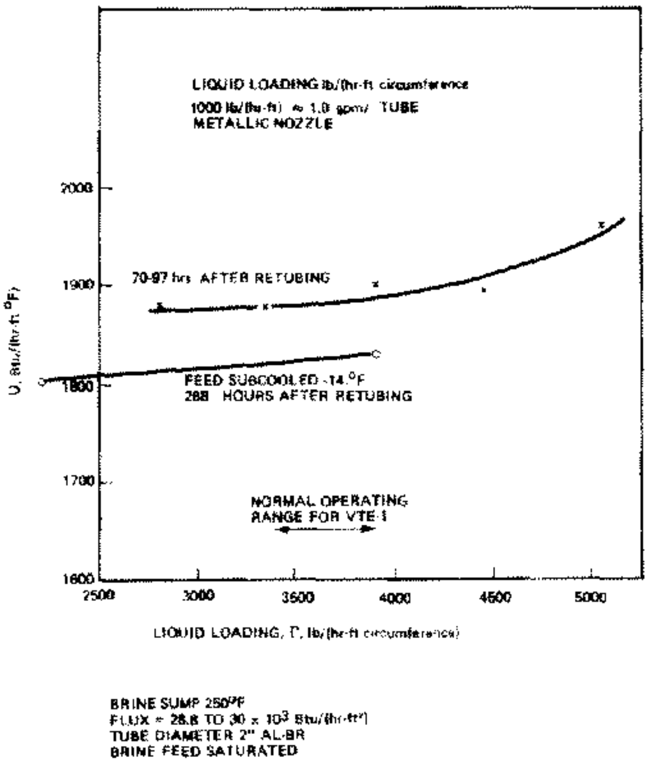

Figure 11. The Heat Transfer Coefficient as a Function of Varying Tube Loading to the VTE-1 Tube Bundle.

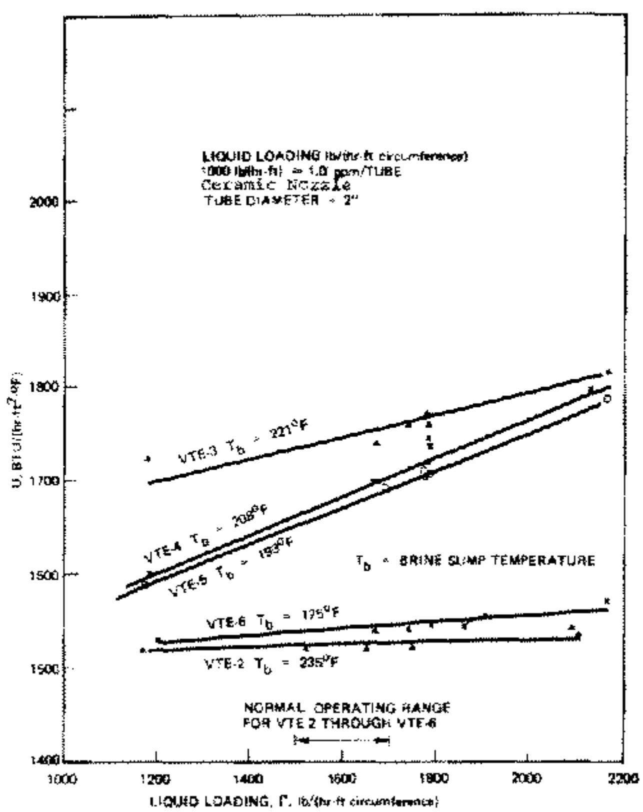

Figure 12. The Effect of Brine Tube Loading on The Heat Transfer Coefficients in Effects 2 through 6.

inlet flash baffle in MSF-1, which allowed an increased liquid loading rate without loss of product purity. The careful control of the relation between the temperature of the brine heater HX-312 and the VTE-1 steam chest pressure was an important factor in permitting the increased brine flow from VTE-1 sump to return to MSF-1 without excessive carryover.

It is obvious that plant hydraulics and not the tube loading criteria was responsible for past limitation.

Finally, based on our experience, we conclude that none of the above mentioned limits were reached, and our normal operating range of 3400 to 3900 lb/hr X ft of liquid loading should be used in future design of the first effect VTE-1 operations in the 250 to 260°F temperature range.

It is believed that lower VTE-1 liquid loadings in the 1500 to 1700 lb/hr X ft range as used in VTE-2 through VTE-6 could be used without danger of overconcentration, but this was not proven in actual operation.

VTE-2 through VTE-6

For effects VTE-2 through VTE-6, as is shown in Figure 12, no clear relationship was established between heat transfer coefficient and the liquid loading of brine flow down each of the tubes.

These effects operated with liquid loadings as low as 1088 lb/hr X ft and as high as 2296 lb/hr X ft. It is important to note that the normal operating range for effects 2 through 6 was at liquid loadings between 1500 and 1700 lb/hr X ft, and no detrimental effect on the tube performance was observed during a total of 11,134 operating hours.

The designer may safely use 1500 lb/hr X ft or approximately 1.5 GPM per 2-inch O.D. tube and realize the savings in the pumping power over the previously suggested 2 GPM per tube.

STEAM QUALITY

No clear relationship was established between the overall heat transfer coefficient and the quality of steam.

Runs with wet steam with a quality below 90%, and with superheated steam at 103% quality showed no significant effect on the performance of the VTE-1 bundle.

No enhancement in performance due to superheat, or any detrimental effect due to an increase in the condensate loading with wet steam, could be confirmed.

VENTING

The reduction in heat transfer, caused by the presence of a non-condensable gas in condensing steam, has long been known. In order to give up its latent heat, the vapor must diffuse through the non-condensables on the vapor-liquid interface. This provides an additional thermal resistance, and thus reduces the overall driving force, ΔT.

The non-condensables are present in the VTE plant due to incomplete deaeration and decarbonation, and due to air in-leakage. The non-condensable effect on heat transfer is predicted well by the Colburn-Hougen relation. The combined analogy between heat transfer and mass transfer was successfully utilized in the models proposed by Eissenberg (Eissenberg, 1969) and Van Winkle (Van Winkle, 1971) and tested in the VTE-X module.

In the VTE/MSF Freeport plant, the normal venting procedures included cascading vents from VTE-1 and VTE-2, venting VTE-2 to the atmosphere, and cascading the vents of the subsequent eight effects downstream to the vacuum system. At a venting rate from 0.8 to 1.5% of the total steam rate, the performance of the plant was stable.

The VTE-1 vent was closed for 100 hours in several runs without significant effect on VTE-1 performance. The stable pressure in VTE-1 Steam chest and the constant rate of steam condensation suggested that the live steam was highly deaerated. Determination of oxygen concentration in the condensate of VTE-1 confirmed this fact.

The increase in venting rates of VTE-1 above 2% of the total steam mass flow did not significantly change the VTE-1 heat transfer coefficient. By closing the vents in all six VTE effects, it was observed that after two hours, the non-condensables caused a continuous decrease in the unit capacity, and the pH of the product dropped from 8.2 to 7.4, explainable by the absorption of CO_2.

To demonstrate the effect of non-condensables, the following tests were run: Run 3–16 with VTE-1 vent closed was performed to obtain the base-line values. Then the vents in VTE-2 and VTE-3 were closed, and after 6.5 hours of stable operation, H&MB run 3–17 was taken. The vents in effects VTE-2 and VTE-3 were opened, and after one hour, H&MB run 3–18 was taken.

Table VI compares runs 3–16, 3–17, and 3–18 and their results. It shows the maximum percentage decrease in the heat transfer coefficient of effects VTE-2 and VTE-3 during the test. It is interesting that after one hour, the coefficients in run 3–18 did not recover the values of run 3–16, and that longer venting time was required. Also, the condensate subcooling suggests the non-condensable layer effect on the condensing film.

Comparison of results in VTE-1 indicates the possibility of non-condensable binding on the boiling side. The decrease in coefficient by 5% in run 3–17 compares with runs 3–16 and 3–18 and could be a result of non-condensable buildup *inside* the doubly fluted tubes.

The detrimental effect of gases on tube-side performance had been previously indicated when the brine waterbox vent in VTE-1 was found plugged. A recovery of over 10% in the heat transfer coefficient for VTE-1 resulted from unplugging the waterbox vent.

Results of tests are limited to preliminary evaluation and, while incomplete due to the expiration of the Development Program, the following general conclusions are offered:

TABLE VI

Effect of noncondensable venting on "U"

Heat Transfer Coefficient HTC, BTU/hr X ft² X °F						
Run Number	VTE-1	VTE-2	VTE-3	VTE-4	VTE-5	VTE-6
3-16[1]	1123	1417	1692	1685	1584	1513
3-17[2]	1065	1126	1446	1658	1560	1482
3-18[3]	1104	1376	1631	1750	1570	1513
(3-16) minus (3-17)	58	291	246	27	24	31
% decrease in "U"	5.2	20.5	14.5	1.6	1.5	2.1

Condensate Subcooling T Vapor-T Condensate °F						
3-16	.4	2.2	1.9	2.2	2.3	5
3-17	.8	7.1	4.2	2.1	2.1	4.6
3-18	.7	3	2.1	1.5	2.2	4.7

Vapor Flow-Klb/hr and Venting Rate—Klb/hr												
3-16	24.16	0	23.75	−.35	20.89	.33	21.50	.23	20.64	.32	21.52	.35
3-17	22.97	0	22.57	0	18.92	. 0	19.90	.16	19.96	.18	20.85	.26
3-18	23.92	0	23.71	−.19	20.49	.22	20.98	.17	20.53	.19	21.53	.27

*Base Line Values

**After VTE-1, -2, and -3 shell side vents closed for 6.5 hours

***After VTE-2 opened to atmosphere and VTE-3 cascaded for one hour

1. The vent rate is important in maintaining the heat transfer coefficient.

2. The venting rate of 1% of the total vapor rate is sufficient to achieve stable performance in the cascading vent system.

3. The vent location does not influence the performance, and top or bottom vent lines are equally effective.

4. The mass velocities of 4200–4800 lb/hr X ft^2 in the inlet to the first vapor pass, and 2200–2500 lb/hr X ft^2 in the inlet to the second vapor pass are considered adequate for sweeping the non-condensable gases and to maintain turbulent flow conditions on the steam-side.

5. The heat transfer surface in the second pass after the venting baffle is 26% of the total, and the equilateral triangle pitch, S, of 2.5 inches with 2-inch O.D. tubes (S/D-1.25), are good design criteria for sizing VTE bundles for a 1 MGD capacity plant.

6. Decrease in heat transfer coefficient without adequate venting is significant, and a maximum decline of over 20% in the overall performance can be expected.

SUMMARY

The above results show that the use of vertical tube evaporators with doubly fluted tubes represents today a well-demonstrated and economically attractive process for desalination of seawater.

To fully understand the advantage of the enhanced tubing, the results above can be compared with smooth tube performance.

The heat transfer coefficient is enhanced by 180–200% at a 250°F brine temperature, and at lower temperatures (175°F), the enhancement over smooth tubing is over 350%.

The attractiveness of doubly fluted tubes is further increased when we consider the low cost penalty for enhancement. Based on January, 1973 price information, the cost penalty over smooth tubes is 12 to 35% depending on the type of enhancement. However, this penalty is insignificant compared to the capital cost savings in being able to use smaller evaporators and less tubes. Obviously, the overall advantages over smooth tubing shows that economic application of enhanced tubing can be successfully utilized in advanced desalting technology. The enhanced tubes are available today in a variety of materials and profiles, and in our opinion can be utilized in the forced circulation and calandria type evaporating-crystallizing operations.

It is recommended therefore that designers of the evaporating-crystallizing equipment include in their investigations and design consideration the unique advantages of the enhanced tubing herein described.

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