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OFFICE OF SALINE WATER

**A SIMPLIFIED METHOD FOR MAKING PRELIMINARY THERMAL
AND ECONOMIC ANALYSES OF VERTICAL-TUBE AND
MULTISTAGE-FLASH MULTIEFFECT EVAPORATOR PLANTS**

E. C. Hise

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OFFICE OF SALINE WATER

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AUGUST 1967

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ABSTRACT

A simple method is presented for performing preliminary overall thermal and economic analyses of vertical tube and multistage-flash evaporator plants. The analyses are based on the terminal conditions of the plant and average values for heat transfer coefficients, boiling point elevations, heats of vaporization, etc. The method is not intended for effect-by-effect or stage-by-stage calculations. However, after the overall conditions have been established, the detailed calculations can be made with the same equations or by other means. Equations are developed for calculating performance ratio, heat requirements, surface area requirements, heat cost per unit product, and surface area cost per unit product. Equations are given for determining the optimum performance ratio for any given set of heat and surface area costs.

INTRODUCTION

This paper presents a simple method for performing both thermodynamic and economic analyses of evaporator plants. This method is not intended to replace other methods for detailed design calculations but rather to permit one to make quickly a parametric survey to select a process configuration best suited to a given set of conditions, to determine the major process quantities, and, perhaps most important, to permit one to easily understand the interrelationship of the various factors in an evaporator process. The method does contain approximations in that it uses average values for boiling point elevation, heat transfer coefficient, heat of vaporization, and installed surface cost. Obviously averages that have been established for a given range of operating conditions should be used with care when extrapolating very far beyond that range. One is therefore cautioned to examine all values used for applicability to the conditions of use.

The method of analysis of the multiple effect vertical tube evaporator (VTE) utilizes the perhaps novel but valid model that the multieffect evaporator is composed of an evaporator portion, whose function is to condense steam and evaporate brine, and a heater portion, whose function is to cool product and heat feed, and that these two functions can be and most easily are evaluated separately and then combined. The heater is a special case of the multistage flash evaporator (MSF) and therefore any MSF can be analyzed with the equations presented for the heater. The analysis is extended in Appendix II to show that the multieffect VTE inherently makes more efficient use of its heat transfer surface than any MSF having less than an infinite number of stages.

PLANT MODEL

The multiple effect vertical tube evaporator plant, hereafter called plant, is considered to be composed of a multieffect vertical-tube section, hereafter called evaporator and a once-through multistage feed heater section hereafter called heater. The plant model chosen as the basis for this study is illustrated in Figure 1.

The method of analysis is based on the premises:

1. A plant attains its highest thermodynamic efficiency when there is no flow of heat between the evaporator and the heater. This may or may not represent the most favorable distribution of heat transfer surface and overall plant economy.
2. The evaporator and the heater can be analyzed separately and the efficiency and economics of the plant is the properly combined sum of the two.
3. Departures from the model may be desirable, can be evaluated with respect to the model, and an analysis can be closely approximated by this method.

Figure 1
Plant Model

In the plant model the evaporator and heater may or may not be physically separate. There is an isothermal interchange of material (no flow of heat) between the evaporator and heater. The heat that is added by process steam to the first effect of the evaporator passes essentially undiminished or unaugmented through each effect to the final condenser. The heat that is added by process steam to the brine heater section of the heater raises the feed to its maximum temperature. This temperature increase from nonregenerative heating is called brine heater rise (BHR). All other heating of the feed is regenerative and is accomplished by an equivalent cooling of the brine and the product water. Brine feed to each evaporator effect is taken from the flashing brine stream of the heater stage where the brine temperature is equal to the evaporating temperature of the effect. The brine passes through the evaporator effect tubes with a negligible temperature change. Brine and product from each evaporator effect are sent to the heater stages where the heater stage vapor temperatures are equal to the evaporating brine and product vapor temperatures.

Now it can be seen that, if there is a flow of heat (Q_1) from an effect of the evaporator to a stage of the heater, the amount of product made in that effect and subsequent effects will be reduced by the amount of $Q_1/1000$ lbs. (Here we assume that the average latent heat of vaporization of the feed is 1000 Btu/lb.) However, the feed (f , in lb/hr) in that stage of the heater will receive a temperature increase greater than the temperature decrease of the brine and product in that stage by the amount of Q_1/f degrees. The feed brine will then require less heat addition in the amount of $(Q_1/f) \times f$ or Q_1 . That is, if heat flows from an effect to a stage at a given rate, then the heat requirement of the heater is reduced by that rate but the production of the plant is reduced by that rate times the total number of effects from that effect to the last effect. Therefore, there is a significant loss in performance ratio. (Performance ratio is defined as pounds of product per 1000 Btu supplied.) If, conversely, heat (Q_1) flows from a stage of the heater to an effect of the evaporator the amount of product made in that and subsequent effects will increase by $Q_1/1000$ lbs per effect and the feed (f) in that stage of the heater will receive a temperature increase less than the temperature decrease of the brine and product in that stage by the amount of Q_1/f degrees. The feed brine will then require more heat in the amount of $(Q_1/f) \times f$ or Q_1 . That is, if heat flows from a stage to an effect at a given rate, then the heat requirement of the heater is increased by that rate but the production of the plant is increased only by that rate times the number of effects from that effect to the last effect rather than by all the effects as would be the case if the heat entered the first evaporator effect. Again there is a loss in performance ratio. Therefore any departure from the model results in a loss of performance ratio and premise (1) is demonstrated. That is not to state that the product cost is necessarily lowest when the thermodynamic efficiency is highest. This balance will be explored in the economic analysis.

THERMAL ANALYSIS

Evaporator

Heat Input

In the model, the process heat that must be added to the first effect of the evaporator is designated the heat rate (Q_{ev}) and is equal to the total evaporator product rate (P_{ev}), (not the plant product rate) times the heat of vaporization divided by the performance ratio (PR_{ev}). This is from the definition of performance ratio. The average heat of vaporization over the temperature range of operation is assumed to be 1000 Btu/lb. This can be written as

$$Q_{ev} = \frac{P_{ev} \times 1000}{PR_{ev}} \quad (1)$$

Performance Ratio

The evaporate produced per effect (P_{ef}) is equal to the heat rate (Q_{ev}) divided by the heat of vaporization or $P_{ef} = Q_{ev}/1000$, and the total evaporator product is the number of effects times the product from each effect or $P_{ev} = N_{ef}P_{ef}$ (since heat is supplied to only the first effect, $Q_{ef} = Q_{ev}$). Combining these with Eq. (1) yields

$$PR_{ev} = N_{ef} \quad (2)$$

Thus the performance ratio is numerically equal to the number of effects. It is assumed that the condensate from the first effect is returned to the power plant and that the product comes from $N_{ef}-1$ effects plus the final condenser or a total of N_{ef} effects.

Combining Eqs. (1) and (2) we have

$$Q_{ev} = \frac{1000 \cdot P_{ev}}{N_{ef}} \quad (3)$$

Heat Transfer Surface

The amount of heat transfer surface required per effect (A_{ef}) is the heat rate divided by the effect temperature differential (ΔT_{ef}) and the heat transfer coefficient ($\bar{U}$) or

$$A_{ef} = \frac{Q_{ev}}{\Delta T_{ef} \bar{U}} \quad (4)$$

The total heat transfer surface required for the evaporator (A_{ev}) will equal the area of one effect times the number of effects,

$$A_{ev} = A_{ef} N_{ef} \quad (5)$$

This can also be determined from the overall evaporator parameters.

Combining Eqs. (3) and (4) gives

$$A_{ef} = \frac{1000 P_{ev}}{N_{ef} \Delta T_{ef} \bar{U}} \quad (6)$$

The average ΔT_{ef} is the total available temperature difference over the evaporator or the flashing range (FR) divided by the number of effects, minus the temperature losses (BPE). BPE as used here is the sum of the boiling point elevation due to salt concentration and the temperature losses due to the vapor pressure drop in going from one effect to the next. The elevation due to the salt can be estimated from concentration-temperature curves such as shown in Figure 2. The equation for the average ΔT_{ef} is

$$\Delta T_{ef} = \frac{FR}{N_{ef}} - BPE. \quad (7)$$

Combining Eqs. (5), (6), and (7)

$$A_{ev} = \frac{1000 P_{ev} N_{ef}}{\bar{U} (FR - N_{ef} BPE)} \quad (8)$$

For Eq. (8) it is satisfactory to use the average of the $\bar{U}$ and BPE over the operating range. It is interesting to note that for a given product rate and flashing range, the surface area per effect increases with the number of effects due to the BPE penalty recurring in each effect as indicated in Eq. (7). The distribution of the total surface between the effects is not implied in the calculations. It is common practice and a good initial assumption to assign equal areas of heat transfer surface to each effect. This makes ΔT_{ef} increase as the evaporator temperature decreases because the heat transfer coefficient usually decreases with temperature.

SEA WATER ... BOILING-POINT RISE

OSW 11.51

Example

An average chemical boiling point rise can be estimated by locating the terminal temperatures and concentrations as indicated by points A and B. These show a rise of 1.3°F and 1.8°F, respectively and can be averaged to 1.55°F.

Temperature losses due to the pressure drop in the vapor flow through pipes, demisters, and bundles must be estimated from the envisioned design and added to the chemical rise. If more accurate information is not available, an estimate of 0.5°F should be satisfactory.

Figure 2

Example of Boiling Point
Rise

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Heater

Heat Input

The heat that must be supplied to the feed heater Q_h equals the brine heater rise (BHR) times the feed rate (f) (the specific heat of the brine is assumed to be 1.0)

$$Q_h = \text{BHR } f. \quad (9)$$

The feed rate can be determined from

$$f = P \frac{\text{CR}}{\text{CR} - 1}, \quad (10)$$

which comes from the definition of concentration ratio, $\text{CR} = f/(f-P)$.

Performance Ratio

Because the heater is a special case of the multistage flash plant, a more rigorous analysis is required for the heater than the approximations that are usually adequate for the multistage flash. The difference arises because the vertical tube evaporator superimposes a high concentration ratio, usually of the order of 2.5 to 3, upon the heater in a once-through cycle whereas the multistage flash is capable of a concentration ratio of the order of 1.18 in a once-through cycle. Therefore the brine stream available for flashing into product is decreasing much more rapidly in the heater than in the multistage flash so the heater will make less product per unit heat input than a multistage flash operating in a comparable range. Therefore, the statement

$$\text{PR} = \frac{\text{FR}}{\text{BHR}} \quad (11)$$

(which is approximately true for a flash evaporator) is not an adequate approximation for a heater. Equation (11) is based on the assumption that the flashing brine stream flow rate throughout the plant is equal to the feed rate. This, of course, is not true because the average brine flow rate is the feed rate reduced by the average product rate. A much closer approximation for a multistage flash evaporator is

$$\text{PR} = \frac{\text{FR}}{\text{BHR}} \left(1 - \frac{\text{FR}}{2000} \right) \quad (12)$$

where the term in parenthesis is about 0.92 to 0.93 for the usual case. The derivation of Eq. (12) is given in Appendix I.

In the heater, the brine stream to feed stream ratio departs even farther from unity (a much higher percentage of the surface is devoted to recovering heat from the product) and the performance ratio now is most easily expressed in terms of the plant concentration ratio (CR). The amount of product made in the heater (P_h) is the average brine stream (b_{av}) times its total temperature loss (FR) divided by the heat of vaporization,

$$P_h = \frac{b_{av} \text{ FR}}{1000} \quad (13)$$

The average brine stream is the sum of the brine entering the process (which is the feed) and the brine leaving the process (the blowdown) divided by two. Or, expressed in terms of feed rate and concentration ratio, the average brine stream is

$$b_{av} = \frac{f(CR + 1)}{2CR} \quad (14)$$

Equation (14) can be written to be used to determine the feed rate,

$$f = \frac{2b_{av} CR}{CR + 1} \quad (14a)$$

or in terms of product rate and concentration ratio

$$f = P_p \left(\frac{CR}{CR - 1} \right) \quad (14b)$$

From Equations (11) and (14) the performance ratio is

$$PR_h = \frac{FR}{BHR} \left(\frac{CR + 1}{2CR} \right) \quad (15)$$

The value of CR must be the true value on a once-through basis, not that which is obtained by recycling. The derivations of Eqs. (14) and (15) are given in Appendix I.

Heat Transfer Surface

Although the number of stages does not appear in the definition of performance ratio or in the equation to be developed for required heat transfer surface, it is implied in the BHR and LMTD terms because the flashdown per stage is an inverse function of the number of stages. The BHR is the sum of the temperature difference between the feed leaving the first heater stage and the condensing temperature in that stage (TTD), the boiling point elevation in that stage, and the stage flashdown (FD).

That is,

$$\text{BHR} = \text{TTD} + \text{BPE} + \text{FD}. \quad (16)$$

The total number of stages in the heater (N_h) is

$$N_h = \frac{\text{FR}}{\text{FD}} = \frac{\text{FR}}{\text{BHR} - \text{TTD} - \text{BPE}}. \quad (17)$$

Now, for a heater with a flashing range of 160° and a CR of 3 to have a PR_h of 10, considering Eq. (15), it is necessary for the BHR to equal about 10.5°F . With a BPE in the first stage of about 1.1° and if we permit a TTD of 6° , the flashdown will equal $10.5 - 1.1 - 6$ or 3.4° . The restriction against flow of heat between the evaporator and the heater means that the sum of $\text{TTD} + \text{BPE} + \text{FD}$ must be a constant throughout the plant. The BPE increases to 2.0° or even 2.5° in the final stages so that if FD is to remain constant, TTD must decrease to 5.1° or even 4.6° . The number of stages required then would be 47 and the LMTD would range from about 7.7 to 6.1° . Of course, reducing the TTD would increase the FD and decrease N_h .

The heat transfer surface required for the heater can be calculated from the standard heat transfer equation, $A_h = Q / U\Delta T$. Since $Q = \text{FR} \cdot f$ and ΔT is the average of the stage LMTD values, we have

$$A_h = \frac{\text{FR} f}{U_{av} \text{LMTD}_{av}}. \quad (18)$$

Plant Thermal Analysis

As stated in the second premise, the performance ratio of the plant is the proper sum of the performance ratios of the evaporator and the heater. From the definition of PR then

$$\text{PR} = \frac{P_{ev} + P_h}{(Q_{ev} + Q_h)/1000}. \quad (19)$$

and by definition $P_{ev} = \frac{(\text{PR}_{ev} Q_{ev})}{1000}$ and $P_h = \frac{(\text{PR}_h Q_h)}{1000}$ so

$$\text{PR} = \frac{\text{PR}_{ev} Q_{ev} + \text{PR}_h Q_h}{Q_{ev} + Q_h}. \quad (20)$$

The plant heat and area requirements are simply the sums of the evaporator and heater requirements.

Application of Thermal Analysis

To perform a heat and materials balance for the design of a plant, it is necessary to select a value or range of values for some parameters. Certain values may be fixed or limited by conditions such as coolant temperature fixed by seawater temperature, maximum brine temperature fixed by feed treatment, and maximum concentration ratio limited by the economics of boiling point elevation and by the problem of scale formation. The limits of other values will become quickly apparent during the calculations. The calculations are sufficiently simple and quick that a range of values for some parameters may be chosen and a parametric survey made. A suggested procedure is:

1. Select a product rate. I find a product rate of one pound per unit time most convenient for comparisons.
2. Select trial values for $T_{\max}$, $T_{\min}$, CR, TTD, N_E and FD.
3. For the selected $T_{\max}$, $T_{\min}$, and CR determine from Fig. 2 the BPE's for the terminal conditions of the evaporator and heater.

4. Calculate for the heater:

BHR	Eq. (16)
f	Eq. (14a)
Q_h	Eq. (9)
PR_h	Eq. (15)
P_h	Eqs. (13) and (14)

5. Calculate for the evaporator

P_{ev}	$P_D - P_h$
Q_{ev}	Eq. (1)

6. Calculate for the plant

PR	Eq. (19) or (20)
----	------------------

The material balance for the evaporator involves simply supplying the evaporator bundle with an appropriate brine rate and removing the brine and product. This rate is a function of the tubing and heat flux and is not within the scope of this paper.

The overall material balance for the heater is already done since P is assigned, f has been calculated, and the blowdown is the difference. If a stage to stage material balance is desired, it will require calculation of the brine flashed in each stage and the amount of vapor condensed in each effect. Bear in mind that credit for product made in a stage can be taken only for flashed brine and not for reflash product.

The temperature at which the feed must be introduced to the last stage feed heater bundle is the brine temperature of the last stage minus the BHR.

An analysis of a design that deviates from the model in that there is flow of heat between the evaporator and the heater can be accomplished by listing the heat duty in each effect and the amount that each effect deviates from the first effect. A decrease in heat duty means an equivalent loss of product in that and all subsequent effects (unless and until there is a compensating flow of heat from the heater to the evaporator) and a comparable decrease in performance ratio below that of the model. It indicates that the heater downstream of that point is less efficient than optimum. An increase of heat duty in an effect means an equivalent loss of product in all preceding effects, a comparable decrease in performance ratio, and indicates that the heater upstream of that point is less efficient than optimum. This discussion refers to heat flow between the evaporator and the heater designed into the plant and recognizes that minor deviations are unavoidable in operation and must be provided for.

ECONOMIC ANALYSES

These analyses were made on the premise that the objective is a minimum product cost.

The two major items of product cost are heat and installed surface. These are inversely related through performance ratio and are the most important items in an optimization.

There are several lesser items of cost that are inversely related to heater surface through concentration ratio and some optimization is possible. These items are chemical treatment and the deaerator. In the special case with certain enhanced surfaces, where the heat transfer coefficient of the evaporator surface is related to the concentration, another variable is introduced to the optimization of concentration ratio.

The final condenser, the seawater intake, and the pumps and pump power can be given some independent consideration on the basis of their inverse relationship to performance ratio and heat rejection temperature.

The remaining cost items can be considered to be independent of process design. These are such things as site work, cranes, instrumentation, electrical, and auxiliary buildings.

Optimum Performance Ratio

Just as it has been shown that the evaporator and heater should be thermally analyzed independently, it can be shown that they should be economically analyzed independently. The evaporator should utilize its heat input and surface to produce the most economical water and the heater should utilize its heat input and surface to produce the most economical brine supply to the evaporator. If each unit is optimum, then the sum of

the two is optimum. A departure from optimum for one unit cannot be compensated for by making the other unit "more optimum." We will therefore examine the two independently.

Evaporator

Optimization of the performance ratio of the evaporator consists simply of finding the minimum sum of heat cost, and surface cost per unit of product.

The heat cost per unit product (H) is the cost of steam in cents per million Btu ($\$/M$) times the heat of vaporization (assumed to be 1000 Btu/lb) times the product unit in lb (we will use 1000 gal as our unit) divided by the performance ratio.

$$H_{ev} = \frac{(\$/M) (1000) (1000) (8.345)}{PR_{ev}} = \frac{(8.345 \times 10^6) (\$/M)}{PR_{ev}} \quad (21)$$

and is shown graphically in Fig. 3 for assumed steam costs of 6, 10 and 22 cents per million Btu.

The cost of the surface then must also be expressed in terms of cents per unit production which we will again take as 1000 gal. This is obtained by multiplying the area required per 1000 gal per hr times the area cost per square foot per hour.

The area requirement can be obtained by using a product rate (P_{ev}) of 1000 gal/hr (8345 lbs/hr) in Eq. (8) yielding (this P_{ev} must be the evaporator product not the plant product):

$$A_{ev} \text{ per 1000 gal per hr} = \frac{(8.345 \times 10^6) (N_{ef})}{U (FR - N_{ef} BPE)} \quad (22)$$

and the area cost per sq ft per hr is the unit cost in cents per sq ft ($\$/sq ft$) times the amortization rate (r) divided by the hrs per year and the on-stream factor (F_s).

$$\text{Surface Cost per hr} = \frac{(\$/sq ft) (r)}{(365)(24)(F_s)} \quad (23)$$

so the surface cost in cents per 1000 gal (S_{ev}) is the product of Eqs. (22) and (23)

$$S_{ev} = \frac{(952) (N_{ef}) (\$/sq ft) (r)}{(U_{av}) (FR - N_{ef} BPE_{av}) (F_s)} \quad (24)$$

The term $\phi_{sq\ ft}$ must include all capital costs related to an effect - namely, the vertical tubes and the tube sheets, the entrainment separators, the shell, the effect pump and the related piping, but not the feed heater costs, divided by the total heat transfer area.

The term r varies with the type financing available. For this example, r was chosen as 0.058 to cover 30-year amortization, interest, insurance, and interim replacements.

The term F_s represents the fraction of time the plant is in operation or on-stream factor and the value used depends upon the ground rules or degree of optimism chosen. For this example, a value of 0.90 is used.

The flashing range FR is $T_{max} - T_{min}$.

The $\bar{U}$ is the average for the surface chosen over the range of conditions chosen.

BPE should be taken as an average over the operating range.

The calculation should be performed for enough N_{ef} 's to be able to draw a smooth curve in the range of interest.

Heater

The heat cost per unit product (H_h) is calculated the same as for the evaporator as given in Eq. (21).

The area requirements per 1000 gal of product per hr made in the heater can be obtained from Eq. (18) modified by expressing the feed (f) in terms of 1000 gal per hr of product (P) made in the heater. See Appendix II for the derivation.

$$A_h \text{ per 1000 gal per hr} = \frac{(16.7 \times 10^6) CR}{(\bar{U}_{av}) (LMTD_{av}) (CR + 1)} \quad (25)$$

and the surface cost in cents per 1000 gal (S_h) is the product of Eqs. (23) and (25).

$$S_h = \frac{(1910) (\phi_{sq\ ft}) (r) (CR)}{(\bar{U}_{av}) (LMTD_{av}) (CR + 1) (F_s)} \quad (26)$$

Unfortunately for the cause of simplicity, it is not convenient to express Eq. (26) directly in terms of performance ratio because of the interrelationship of stage flashdown, LMTD, and performance ratio. However, if a number of stages is assumed thus establishing a stage flashdown, then the performance ratio and S_h can be calculated for each assumed LMTD and a family of curves representing S_h vs. PR for different numbers of stages can be plotted.

SAMPLE CALCULATIONSOptimum Performance Ratio

The following parametric study, using hypothetical inputs for heat costs, surface cost, and overall coefficient, illustrates the optimization procedure.

Evaporator

$$S_{ev} = \frac{952 (N_{ef}) (\phi_{sq \text{ ft}}) (r)}{(\bar{U}_{av}) (FR - N_{ef} \text{ BPE}) (F_s)} \quad (24)$$

where $\phi_{sq \text{ ft}} = 500$ cents per square foot complete installed cost including overhead, etc.

$$r = 0.05185$$

$$F_s = 0.90$$

$$\bar{U} = 600$$

$$FR = 160^\circ$$

$$\text{BPE}_{av} = 2^\circ \text{ including losses}$$

$$N_{ef} = \text{PR}_{ev}$$

then

$$S_{ev} = 45.6 \frac{\text{PR}_{ev}}{(160 - 2\text{PR}_{ev})}$$

	<u>PR_{ev}</u>					
	9	11	13	15	17	23
<u>S_{ev} cents per 1000 gal</u>	2.89	3.64	4.42	5.26	6.15	9.2

The cost of heat for the evaporator is determined from Eq. (21).

$$H_{ev} = \frac{(8.345 \times 10^6) (\phi_M)}{\text{PR}_{ev}}$$

$$@ 6\text{¢/MBtu} = \frac{50}{\text{PR}} \text{ ¢/1000 gal}$$

$$@ 10\text{¢/MBtu} = \frac{83.5}{\text{PR}} \text{ ¢/1000 gal}$$

$$@ 22\text{¢/MBtu} = \frac{183}{\text{PR}} \text{ ¢/1000 gal}$$

Heater

$$S_h = \frac{(1910) (\phi_{sq \text{ ft}}) (r) (CR)}{(\bar{U}_{av}) (LMTD_{av}) (CR + 1) (F_s)} \quad (26)$$

where $\phi_{sq \text{ ft}} = 500$ cents per sq ft complete installed cost including overhead, etc.

$$r = 0.5185$$

$$F_s = 0.90$$

$$\bar{U}_{av} = 700$$

$$FR = 160^\circ F$$

$$BPE = 2^\circ F$$

$$CR = 3$$

$$S_h = \frac{(1910) (500) (0.5185) (3)}{(700) (3 + 1) (0.90)} \cdot \frac{1}{LMTD} = \frac{59}{LMTD}$$

The values of S_h are given in Table I in cents per 1000 gallons for the LMTD's associated with various number of stages and performance ratios.

Overall Plant

Overall optimum performance ratio from the optimum PR_{ev} and PR_h as indicated on Fig. (3) for a 40 stage evaporator and 6¢/MBtu heat, are

$$PR_{ev} = 13 \quad PR_h = 8$$

$$P_h = \frac{b_{av} FR}{1000} = 0.16 \times 8345 \text{ lb/hr} = 1333 \text{ lb/hr} \quad (13) \text{ \& (14)}$$

$$Q_n = BHR \cdot f = 13.3 \times 8345 \times 1.5 = 167 \times 10^3 \text{ Btu/hr} \quad (9) \text{ \& (10)}$$

$$p_{ev} = 8345 - 1333 = 7012 \text{ lb/hr}$$

$$Q_{ev} = \frac{1000 P_E}{N_E} = \frac{7012 \times 1000}{15} = 540 \times 10^3 \text{ Btu/hr} \quad (1)$$

$$PR = \frac{7012 + 1333}{167 + 540} = \frac{8345}{707} = 11.8 \quad (19)$$

Table 1. Cost of Heater Surface

PR_h	BHR (°F)	TTD or T_1 (°F)	T_2 (°F)	LMTD	S_H (¢/1000 gal)
[1]	[2]	[3]	[4]	[5]	[6]
30 stages, $FD = 160/30 = 5.33^\circ/\text{stage}$					
6	17.8	10.78	15.8	12.9	4.56
8	13.3	6.0	11.3	8.4	7.02
10	10.65	3.32	8.65	5.6	10.50
12	8.9	1.6	6.9	3.65	16.20
40 stages, $FD = 160/40 = 4^\circ/\text{stage}$					
6	17.8	11.8	15.8	13.7	4.31
8	13.3	7.3	11.3	9.1	6.49
10	10.65	4.65	8.65	6.45	9.15
12	8.9	2.9	6.9	4.6	12.80
50 stages, $FD = 160/50 = 3.2^\circ/\text{stage}$					
6	17.8	12.6	15.8	14.3	4.13
8	13.3	8.1	11.3	9.5	6.21
10	10.65	5.45	8.65	6.8	8.68
12	8.9	3.7	6.9	5.12	11.51

- [1] Assumed.
 [2] From Eq. (15).
 [3] $TTD = BHR - FD - BPE = T_1$.
 [4] $T_2 = BHR - BPE$.
 [5] $LMTD = (T_2 - T_1) / (\ln (T_2 / T_1))$
 [6] $S_H = \frac{59}{LMTD}$ From Page 15.

Figure 3
Optimization of Performance Ratio

Effect of Concentration Ratio

Heater Surface Cost

We can also determine the effect of changing the concentration ratio on the cost of heater surface from Eq. (26),

$$S_h = \frac{(1910) (\$_{sq ft}) (r) (CR)}{(U_{av}) (LMTD_{av}) (CR + 1) (F_s)}$$

If we assume 40 stages in the heater and the other factors the same as in the previous example, we have the costs for a concentration ratio of 3 at various values of PR_h already given in Table 1. For a concentration ratio of 2, the curves in Fig. 2 indicate an average boiling point elevation of 1.22°F . In the previous example we assumed other losses to be 0.4°F and with this same assumption, we will use $BPE = 1.62^\circ\text{F}$ for $CR = 2$. Table 2 gives the costs for $CR = 3$ calculated previously and the values determined for $CR = 2$ with the new BPE value. The results are shown in Fig. 5 along with a line representing the cost of heat assuming 6 cents per MBtu steam.

Evaporator Surface Cost

The first example gave us values for the cost of evaporator surface as a function of performance ratio for a concentration ratio of 3. Changing the concentration ratio changes only the value for the BPE. Thus, for a $CR = 2$ we will use $BPE = 1.62^\circ\text{F}$ in place of 2.0°F in Eq. (24). These results, along with those determined for $CR = 3$, are given in Table 3. This data is also shown in Fig. 5.

It should be noted that, although decreasing the CR causes an increase in PR_h the value of PR_p may decrease or increase. This is true because the ratio of P_h to P_{ev} increases with decreasing CR. Therefore, if $PR_h < PR_{ev}$, the PR_p decreases and if $PR_h > PR_{ev}$, the PR_p increases.

Table 2. Cost of Heater Surface as a Function of Concentration Ratio

Basis: 40 stages

PR_H	Cost of Surface (cents per 1000 gal)	
	CR = 2	CR = 3
6	3.72	4.31
8	5.50	6.49
10	7.62	9.15
12	10.32	12.80

Table 3. Cost of Evaporator Surface as a Function of Concentration Ratio

PR_E	Cost of Surface (cents per 1000 gal)	
	CR = 2	CR = 3
9	2.72	2.89
11	3.53	3.64
13	4.27	4.42
15	5.05	5.26
17	5.85	6.15
23	3.55	9.2

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Figure 4

Effect on Concentration Ratio on Heater and Evaporator Costs

NOMENCLATURE

b	brine flow rate	lb/hr
BHR	brine heater rise	°F
BPE	boiling point elevation	°F
CR	concentration ratio	feed rate/blowdown rate
f	feed rate	lb/hr
FD	flashdown	°F
FR	flashing range	°F
F _s	Stream factor (fraction of time plant is on-stream)	
H	heat cost	¢/1000 gallons
N	number (of stages or effects)	
P	product rate	lb/hr
PR	performance ratio	lb of product/1000 Btu
Q	heat rate	Btu/hr
r	amortization rate	
S	cost of heat transfer surface	¢/1000 gallons
T	temperature	°F
TTD	terminal temperature difference	°F
U	overall heat transfer coefficient	Btu/hr-ft ² -°F

Subscripts

ev	evaporator
ef	effect
h	heater
av	average
p	plant

APPENDIX I. DERIVATIONSEquation (10)

The feed rate (f) is equal to the product rate (P) plus the blowdown,

$$f = P + \text{blowdown} \quad (\text{A-1.1})$$

By definition, the concentration ratio (CR) is

$$\text{CR} = f/\text{blowdown}. \quad (\text{A-1.2})$$

Combining these gives

$$\text{CR} = f/(f - P) \quad (\text{A-1.3})$$

or

$$f = P(\text{CR})/(\text{CR} - 1). \quad (10)$$

Equation (12)

In the body of this report, Eq. (11) comes from

$$\text{PR} = \frac{\text{FR}}{\text{BHR}} \cdot \frac{b_{av}}{f} \quad (11a)$$

and we considered that $b_{av} = f$ which results in Eq. (11),

$$\text{PR} = \frac{\text{FR}}{\text{BHR}}. \quad (11)$$

A closer approximation of the average brine flow rate takes into account the product removed. This can be written

$$b_{av} = \frac{f + (f - P)}{2}. \quad (\text{A-1.4})$$

The product flow rate (P) is approximately the feed rate times the flashing range divided by the average heat of vaporization.

$$P = \frac{f \cdot \text{FR}}{1000}. \quad (\text{A-1.5})$$

Combining Eqs. (A-1.4) and (A-1.5) gives

$$b_{av} = f \left(1 - \frac{\text{FR}}{2000} \right) \quad (\text{A-1.6})$$

Substituting this expression for b_{av} in Eq. (11a) gives

$$PR = \frac{FR}{BHR} \left(1 - \frac{FR}{2000}\right) \quad (12)$$

Equation (14)

The average brine flow rate can be defined exactly in terms of the feed and blowdown:

$$b_{av} = (f + \text{blowdown})/2 \quad (A-1.7)$$

and the concentration ratio is defined as

$$CR = f/\text{blowdown} \quad (A-1.8)$$

Combining (A-1.7) and (A-1.8) gives

$$b_{av} = \frac{f(CR + 1)}{2CR} \quad (14)$$

Equation (15)

We saw above that Eq. (11) could be written

$$PR_h = \frac{FR}{BHR} \cdot \frac{b_{av}}{f} \quad (11a)$$

Substituting Eq. (14) in this gives

$$PR_h = \frac{FR}{BHR} \left(\frac{CR + 1}{2CR}\right) \quad (15)$$

Equation (25)

Equations (13) and (14) are

$$P_h = \frac{b_{av} FR}{1000} \quad (13)$$

$$b_{av} = \frac{f(CR + 1)}{2CR} \quad (14)$$

Combining these and solving for f gives

$$f = \frac{2000 \cdot CR \cdot P_h}{FR(CR + 1)} \quad (A-1.9)$$

Eq. (18) is

$$A_h = \frac{FR \cdot f}{U_{av} LMTD_{av}} \quad (18)$$

and substituting (A-1.9) in this gives

$$A_h = \frac{2000 \cdot CR \cdot P_h}{U_{av} LMTD_{av} (CR + 1)} \quad (A-1.10)$$

For a product rate (P_h) of 1000 gal/hr or 8345 lb/hr

$$A_h / 1000 \text{ gal/hr} = \frac{16.7 \times 10^6 CR}{U_{av} LMTD_{av} (CR + 1)} \quad (25)$$

APPENDIX II. COMPARISON OF EFFICIENCY OF VTE AND MSF

If a VTE and an MSF process have the same thermal efficiency (PR), then the heat utilization efficiency (HE) of the vte will always be greater. The HE is the ratio of the available temperature differential (ΔT_a) to the total temperature differential (ΔT_t).

$$HE = \frac{\Delta T_a}{\Delta T_t} \quad (A-2.1)$$

For the MSF, the performance ratio is approximately

$$PR = \frac{FR}{BHR} \quad (A-2.2)$$

The ΔT_t in any stage is the sum TTD + FD + BPE which is also equal to the BHR

$$\Delta T_t = TTD + FD + BPE = BHR \quad (A-2.3)$$

The ΔT_a in a stage is the log mean average of the TTD and the (TTD+FD) which of course must be less than (TTD+FD) for a finite number of stages.

$$\Delta T_a < (TTD+FD) \quad (A-2.4)$$

Combining (A-2.3) and (A-2.4)

$$\Delta T_a < (\Delta T_t - BPE) \quad (A-2.5)$$

For the VTE, the performance ratio is

$$PR = N_{ev} \quad (A-2.6)$$

The ΔT_t for an effect is the flashing range divided by the number of effects

$$\Delta T_t = \frac{FR}{N_{ev}} \quad (A-2.7)$$

Combining (A-2.6) and (A-2.7)

$$PR = \frac{FR}{\Delta T_t} \quad (A-2.8)$$

and the ΔT_a is

$$\Delta T_a = (\Delta T_t - BPE) \quad (A-2.9)$$

Comparing Eqs. (A-2.2) and (A-2.3) shows that for a equal performance ratio and flashing ranges that the ΔT_+ of the two plants is the same. Comparing Eqs. (A-2.5) and (A-2.9) shows that the ΔT_a for the MSF will always be less than for the VTE, and therefore the HE will always be less for the MSF.

Furthermore, if the two plants have the same concentration ratio, the BPE for the MSF will be greater since it must recycle. Figure 6 compares the BPE for the two plants at CR = 2.

Numerical Example

Assume for both plants FR = 150° PR = 15 CR = 2.

For the MSF the average BPE is about 1.6° from Figure 6. From Eqs. (A-2.2) and (A-2.3), $\Delta T_+ = \text{BHR} = \frac{150}{15} = 10$. If we assume 50 stages, then FD = 3°

and from Eq. (A-2.3), the TTD = $10 - (3 + 1.6) = 5.4^\circ$

Then the $\Delta T_a = \text{LMTD}$ of 5.4° and 8.4° or 6.8° and the heat utilization efficiency $\text{HE} = \frac{\Delta T_a}{\Delta T_+} = \frac{6.8}{10} = 68\%$.

For the VTE the average BPE is about 1.3° from Figure 6. From Eq. (A-2.7), $\Delta T_+ = \frac{\text{FR}}{\text{PR}} = \frac{150}{15} = 10^\circ$. From Eq. (A-2.9) $\Delta T_a = \Delta T_+ - \text{BPE} = 8.7^\circ$.

$\text{HE} = \frac{8.7}{10} = 87\%$.

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SEA WATER BOILING-POINT RISE

Figure 5

Boiling Point Elevation
Penalty as a Function of
Temperature and Concentra-
tion for MSF and VTE
Processes

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