

6.6 EVAPORATOR DESIGN

A properly designed evaporator must, at a minimum:

- effectively transfer heat at a high rate with minimum surface area so that it is economic for installation, operations and maintenance
- effectively separate vapour from liquid concentrate
- meet solvent evaporation capacity
- meet product quality (concentration)
- be energy efficient by effective use of steam in multiple-effect evaporation or vapour recompression, wherever possible
- minimize fouling of heat transfer surface
- be constructed of materials that are adequate to minimize corrosion

In case of crystallising evaporators, the requirement/need for producing crystals of a definite uniform size usually limits the choice to forced circulation evaporators.

There are three principal elements in evaporator design— heat transfer, vapour-liquid separation and energy utilisation. Among these, heat transfer is the most important factor since the heat transfer surface represents the largest part of evaporator cost.

6.6.1 Single Effect Evaporation

The surface area of a continuous single effect evaporator is obtained from the heat transfer equation as

$$A = \frac{q}{U \Delta T_{eff}} \quad (6.1)$$

Where q is obtained from energy balance neglecting condensate subcooling and radiation losses. It is essential to have a clear understanding of the basis of definitions of ΔT_{eff} and its corresponding U .

The effective temperature driving force for heat transfer (ΔT_{eff}) is given as –

$$\Delta T_{eff} = T_{con} - T \quad (6.2)$$

Where T_{con} is the temperature at which the heating vapour (steam) condenses and T is the bulk liquid temperature, which is estimated by adding the boiling point elevation (BPE) of the liquid leaving the effect to the boiling point (T_s) of pure solvent (water in most cases) at the operating pressure (P) in the effect, i.e.

$$T = T_s + BPE \quad (6.3)$$

BPE is a function of the solute concentration x in the thick liquor leaving and is available from literature and T_s is found from the steam table as saturation temperature corresponding to pressure P .

Feed to the evaporator can be at a lower temperature than the effect but quickly attains the bulk liquid temperature (T) due to well mixed liquid phase in evaporators. The heating steam enters at a temperature T_{steam} with a few degrees of superheat and quickly cools to its saturation (condensing) temperature (T_{con}). Therefore the effective temperature driving force for heat supply to the effect is considered as $\Delta T_{eff} = T_{con} - T$, and the heat transfer coefficient (U) is defined with respect to this temperature driving force.* In an evaporator, the average boiling point of liquid is higher than the boiling point corresponding to the pressure in the vapour space due to the effect of hydrostatic head. This reduces the ΔT across the heating surface, thereby causing a decrease in evaporator capacity.

Figure 6.17 shows the terms of material and energy flows around a single effect with the notations F , L , V , S corresponding to feed, liquor, vapour and steam mass flow rate respectively. Solute concentration (x, x_F in % w/w), enthalpy per unit mass (h) and temperature (T) and pressure (P) values are marked therein. Subscripts L , V , $steam$ and con denote thick liquor, vapour leaving, supply steam and condensate generated. h_{steam} is obtained from steam table at steam supply pressure P_{steam} and temperature T_{steam} . Enthalpy of feed and thick liquor are respective functions of solute concentration (x_F and x) and temperature. These are found from enthalpy charts (see Figure 6.22) for different solute-solvent systems. In absence of such charts or experimental data, information can be obtained from solution rules with or without considering heat of solution.

***Footnote:** Earlier practice was to define U corresponding to the apparent temperature difference $\Delta T_{app} = (T_{con} - T_s)$. One must be careful while employing data reported in literature and check whether the value of reported U is based on true or apparent temperature difference as $\Delta T_{app} \geq \Delta T_{eff}$.

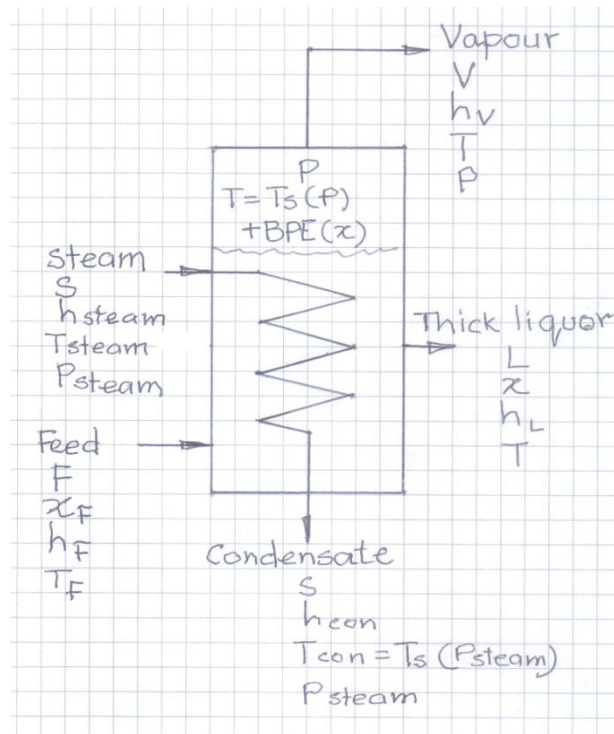


Figure 6.17: Terms related to material and energy flows around a single effect

Using the notations specified in Figure 6.17, the mass and energy balance equations are -

$$F = L + V \quad (6.4)$$

$$Fx_F = Lx_L \quad (6.5)$$

$$Fh_F + Sh_{steam} = Lh_L + Vh_v + Sh_{con} \quad (6.6)$$

Equipment costs are usually correlated as function of heating surface area, material of construction and evaporator type. Other things

Overall heat transfer coefficient

being equal, the evaporator type is selected on the basis of highest heat transfer coefficient under desired operating condition. U is governed by condensing steam side coefficient and boiling liquid coefficient. The steam side

coefficient increases with increase in steam temperature due to decreased viscosity of condensate film and decreasing ΔT since this slows down condensation resulting in thinner liquid films. The boiling liquid side coefficient is influenced by velocity and viscosity of the evaporating liquid and cleanliness of the heating surface. Reduction in temperature difference between the heating fluid and the product requires higher heat transfer coefficient and can be achieved by mechanical agitation.

Usually overall heat transfer coefficients are used in evaporator design. The curves of Figure 6.18 as represented by Eqn. 6.7 can be used for vertical tube evaporators.

$$U = 5.6783 \times \left[b + \frac{(a-b)}{1 + (0.5556 \times \Delta T_{eff} / c)^d} \right] \quad (6.7)$$

Where U and ΔT_{eff} are in W/m^2K and $^{\circ}C$ respectively and the constants a , b , c and d for different boiling points in the effect are given in Table 6.4

Table 6.4: Constants to be used in Eqn. 6.7 for predicting U

Boiling Point($^{\circ}C$)	a	b	c	d
100	0.4530	720.2609	12.7327	0.9524
75	-0.4856	684.4337	22.3611	1.1038
60	0.2090	509.0018	26.5472	1.2798
50	0.3641	687.8854	83.2172	1.0207

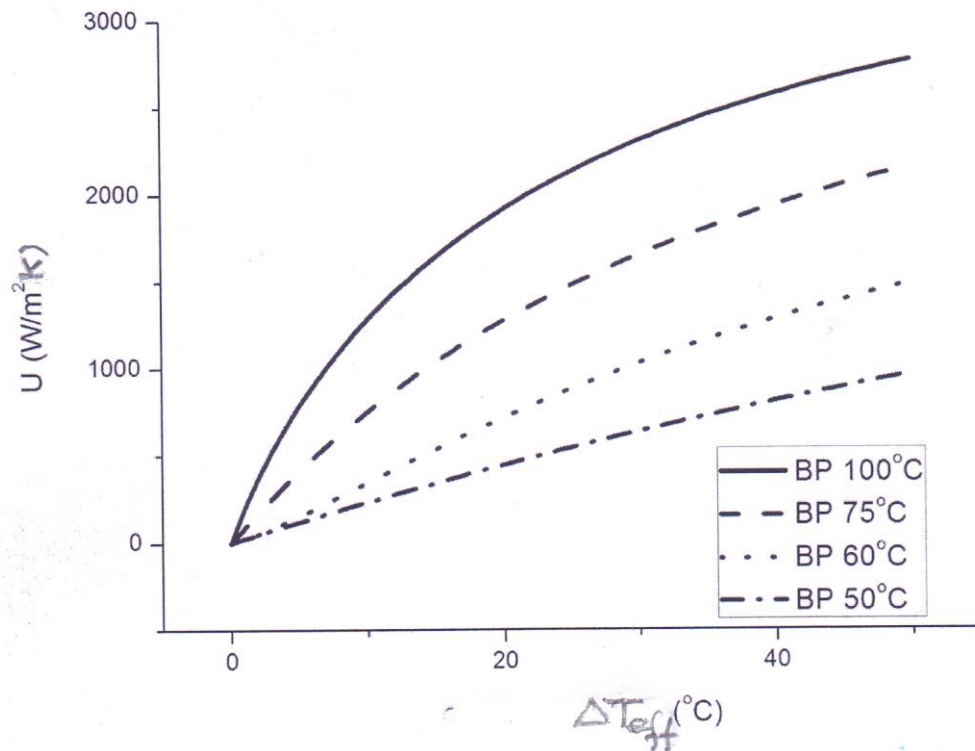


Figure 6.18: Relationship between boiling point (BP), temperature drop (ΔT_{eff}) and overall heat transfer coefficient in vertical tube evaporators

These curves are derived from the experimentally generated curves by Badger and Shepard (AIChE Trans. 13(I):101-137(1920)) for 30" diameter evaporator with 24 numbers of 2" diameter 48" long tubes.

While ΔT_{eff} and condensing temperature (T_{con}) are fixed by operating conditions and not by evaporator construction, the removal of non-condensable gases depends on the evaporator construction. Removal of non-condensable gases in vertical tube evaporator is not very efficient. Non-condensable gases are transferred to the condensate steam chest in horizontal tube, long tube vertical and forced circulation evaporators.

6.6.2 Multiple Effect Evaporation

The equations for single effect are valid for the individual effects of multiple effect evaporator. The rate of heat transfer for any effect 'i' is thus given by

$$q_i = U_i A_i \Delta T_{eff,i} \quad (6.8)$$

Assuming that the feed enters the first effect at its boiling point and the condensate leaving effect 2 is at temperature of vapour obtained from boiling liquid in first effect (T_1), the entire heat used in generating vapours in first effect must be given out when the same vapour condenses in second effect and so on.

Mathematically,

$$q_i = q_{i-1} = q_{i+1} \quad (6.9)$$

or

$$U_i A_i \Delta T_{eff,i} = U_{i-1} A_{i-1} \Delta T_{eff,i-1} = U_{i+1} A_{i+1} \Delta T_{eff,i+1} \quad (6.10)$$

Where $\Delta T_{eff,i}$ is given by Eqn. 6.2.

In ordinary practice, to obtain economy of construction, all A_i s are preferred to be equal, which gives –

$$U_i \Delta T_{eff,i} = U_{i-1} \Delta T_{eff,i-1} = U_{i+1} \Delta T_{eff,i+1} \quad (6.11)$$

Therefore, the temperature difference across the heating surface in each effect ($\Delta T_{eff,i}$) of a multiple effect evaporator is inversely proportional to the heat transfer coefficient.

Optimum number of effects in a multiple effect system

Assuming that the latent heat is not significantly influenced by pressure, it can safely be considered that the latent heat required to evaporate 1 kg steam is same in each effect and for every 1 kg steam supplied to the first effect, N kg water will be evaporated in an N -effect evaporator while the evaporator capacity remains substantially constant.

Total amount of heat required to achieve the evaporator capacity (total evaporation per hour) for N effects is

$$q_{total} = \sum_{i=1}^N q_i = \sum_{i=1}^N U_i A_i \Delta T_{eff,i} \quad (6.12)$$

For all effects having equal area ($A_i = A_{i-1} = A_{i+1}$) and an average coefficient U_{av}

$$q_{total} = U_{av} A \Delta T_{ov} \quad (6.13)$$

Where ΔT_{ov} , the overall temperature difference in the evaporator system is given by

$$\Delta T_{ov} = T_{Steam} - T_v \quad (6.14)$$

T_{Steam} is the heating steam supply temperature and T_v refers to the vapour saturation temperature from the last (N^{th}) effect (including *BPE*).

For a single effect evaporator of area A , operating at ΔT_{ov} with a heat transfer coefficient U_{av}

$$q = U_{av} A \Delta T_{ov} \quad (6.15)$$

Thus for constant ΔT_{ov} , the total capacity of the system remains substantially unchanged by varying the number of effects while the capital cost increases. Thus a N effect evaporator will cost about N times a single effect evaporator. Therefore, the choice of the proper number of effects is dictated by an economic balance between savings in steam obtained by multiple effect evaporation and added investment costs due to area (effect) addition.

The total annualised cost, computed as the sum of the annualised fixed cost (calculated based on the prevailing interest rate) and the annual cost of steam, water and labour is the economic parameter that is optimised. The number of effects that correspond to the minimum annualised fixed cost is the optimum number of effects for the design case. Figure 6.19 shows a typical graph for finding the optimum.

It is needless to say that if the optimum number of effects thus obtained is a fraction, it is rounded off to the next higher number

***Footnote:** Annualised fixed cost (Rs/year) = (Interest % per annum/100) \times Fixed cost (Rs)

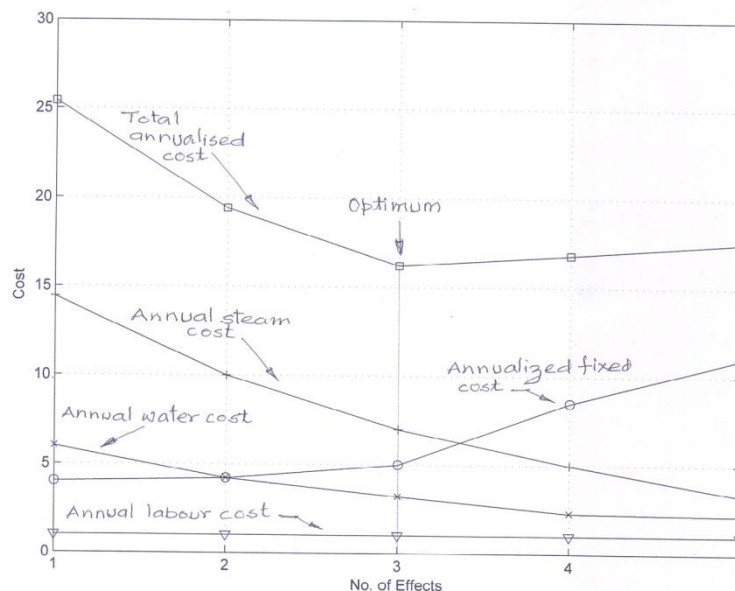


Figure 6.19: Plot for arriving at the optimum number of effects

6.6.3 Design data

Elevation of boiling point (BPE)

Usually evaporators deal with concentrated solutions whose specific heat, latent heat of vaporisation and boiling point deviate from water. There may be other thermal phenomena

(eg. heat of crystallisation) that need to be considered. The elevation in boiling point for different solute percentages at different pressures is estimated from Duhring plot and the change in other thermodynamic properties are estimated from enthalpy-concentration plots.

Duhring plots: These are plots of boiling point of pure solvent (usually water) in the abscissa and the corresponding boiling point of the solution in the ordinate *at the same pressure*. Different lines are for solutions at different concentrations. The points on a particular line are for a solution of same concentration but different pressures. These are straight lines for practical purposes that make interpolation and extrapolation easier. Boiling point elevation (*BPE*) of a solution at a particular concentration is obtained from the difference between the ordinate and abscissa at that concentration where the abscissa corresponds to the pure solvent boiling point at the effect pressure. The linearity of the lines suggests that the boiling point elevation is primarily a function of solute concentration and does not change much with pressure and temperature. This is a fact that is often used for simplifying calculations. Although linear, the lines are not parallel and in general have a steeper slope for more concentrated solution, i.e. boiling point elevation (*BPE*) increases faster with increasing pressure for more concentrated solution than for dilute solutions. Figure 6.20 (a) and (b) are Duhring plots for sodium chloride and caustic soda solutions.

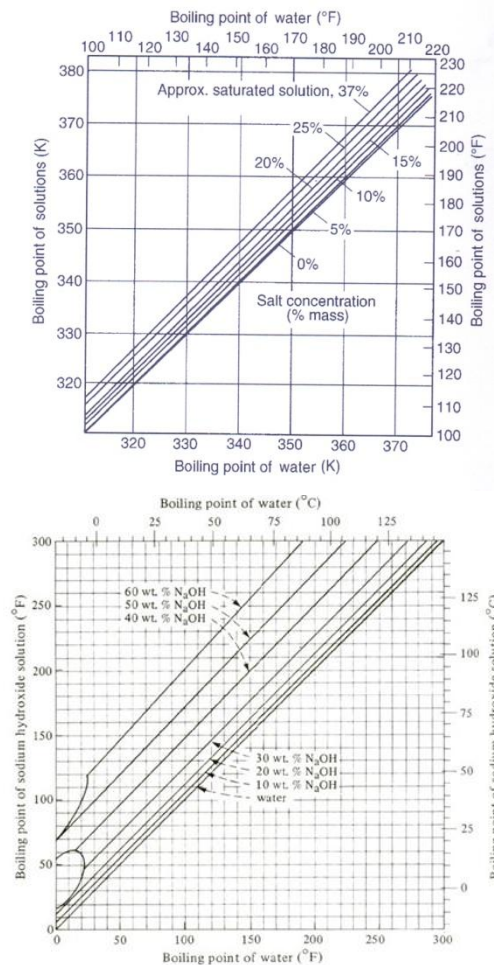


Figure 6.20: Duhring Plot for (a) sodium chloride solution, (b) caustic soda solution

Boiling point elevation is usually small and can be neglected for (i) very dilute solutions and (ii) solutions of materials with extremely high molecular weight.

Information on boiling point elevation not only provides an estimate of the boiling point of concentrated solutions but also gives an idea whether low pressure steam can provide a proper temperature difference (ΔT) for concentrating solutions under atmospheric pressure.*

***Footnote:** When the boiling point elevation is not (accurately) known, it has been a customary practice to use an apparent temperature difference, $\Delta T_{\text{apparent}} (=T_{\text{saturated steam}} - T_{\text{saturated vapour}})$.

Boiling Point Elevation in Multiple Effect Evaporators

The effect of boiling point elevation is more pronounced in multiple effect evaporators. This is illustrated in Figure 6.21 for a single, double and a triple effect evaporator operating under the same terminal conditions, i.e. same steam pressure to the first effect and same saturation temperature of vapour to condenser. Total height of each column in the figure represents ΔT_{ov} , the total temperature spread from steam temperature to saturation temperature of vapour in last effect. The shaded portions represent loss of ΔT due to boiling point elevation in each effect. The figure reveals that in the extreme case of a large number of effects or very high BPE, the sum of the BPEs may be equal to or even more than ΔT_{ov} , the total temperature drop available. Operation is impossible under this condition.

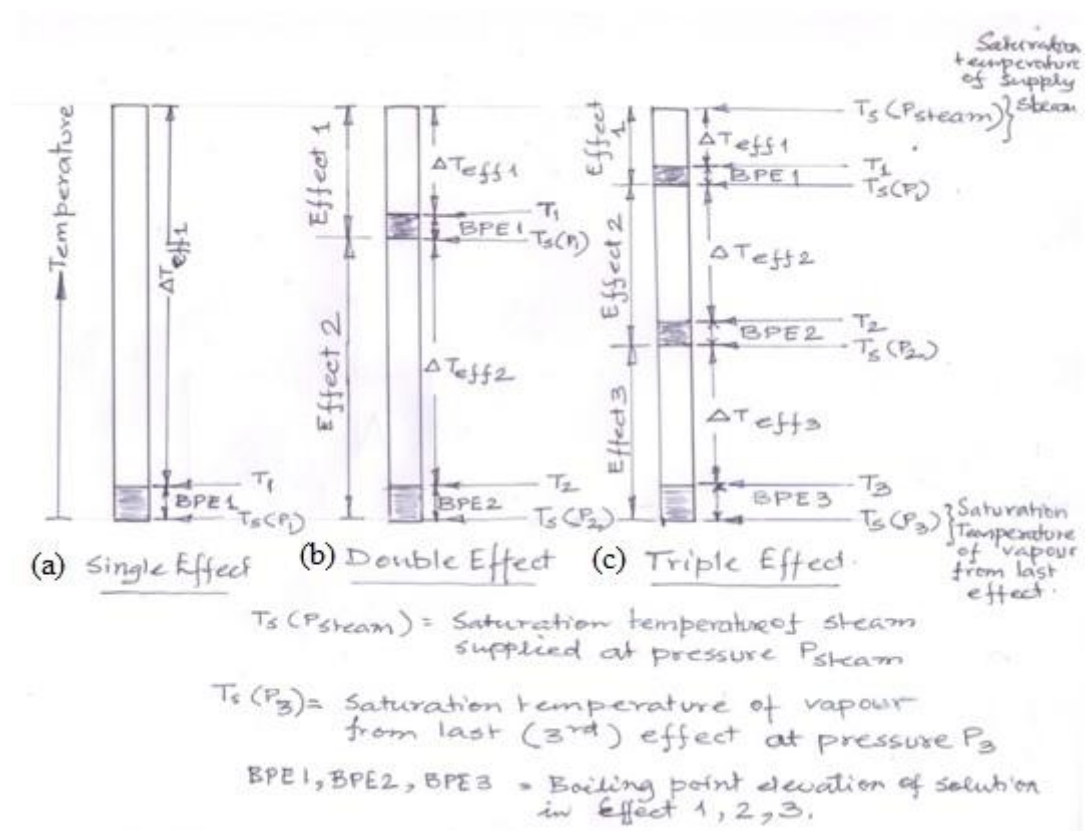


Figure 6.21 Comparison of temperature driving forces in (a) single (b) double and (c) triple effect evaporator with same terminal conditions

Enthalpy plots

Evaporator design calculations involve enthalpy balance around the effects that require solution enthalpy to be evaluated. Enthalpy of the solution is a function of solute concentration and solution temperature and pressure. This data is presented as enthalpy plots e.g. in Figure 6.22, where lines of constant solution boiling temperature are drawn on graphs with solute concentration in abscissa and enthalpy of solution in ordinate. Pressure is not included as a parameter in the enthalpy-concentration plot as the solution boiling point and its solute concentration defines it implicitly. The pure solvent boiling point corresponding to the solution (boiling) temperature and solute concentration can be found from the Dühring plot for the system and the corresponding pressure is the vapour pressure of pure solvent. In some cases, the same information may also be available as separate plot of solution vapour pressure.

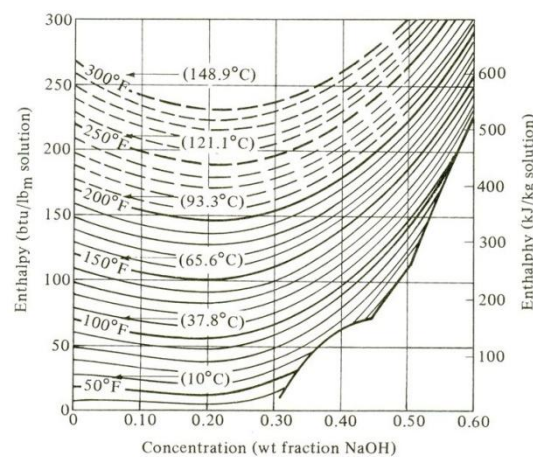


Figure 6.22: Enthalpy Plot for caustic soda solution

In absence of enthalpy-concentration charts, the enthalpy of feed and thick liquor can be calculated from specific heats of components. Heats of dilution are usually neglected. The latent heat of vaporisation of water from an aqueous solution may be taken from steam tables at the actual boiling temperature of the solution rather than the equilibrium temperature of pure water.

$$T_{steam} \text{ \& } T_{con}$$

Among pressure and temperature, pressure can be measured more easily and in the temperature range involved in evaporators, it can be measured more accurately than temperature. Therefore, the pressure in the steam chest is measured and the saturated temperature is obtained from steam tables. The same applies to the vapour space. The saturation temperature is obtained from steam tables corresponding to the measured pressure in the vapour space and the boiling point elevation added to this gives the actual temperature of the steam space.

Steam pressure

Although steam at a higher pressure leads to a higher ΔT and a consequent decrease in the heat transfer area (and cost) of an evaporator, it is rarely used as heating medium for evaporation because –

- High pressure steam is much more valuable as a source of power than a source of heat. It is much cheaper to generate power using high pressure steam and use the exhaust steam in evaporator than to use boiler steam directly.
- Steam at higher pressure has a higher enthalpy but a lower latent heat of vaporisation. So low pressure steam can deliver more latent heat per kg steam.
- The construction of an evaporator to hold high pressure steam would be more expensive.

The disadvantage of low pressure steam is that it exists at a lower temperature and so provides lower ΔT between steam and evaporating liquid. Usually the steam pressure is between 1 atm to 0.7 atm(g) and in some case can be up to 2.5 to 3 atm(g).

Pressure in the vapour space

It is important to note that evaporators do not necessarily have to work under vacuum. A vacuum is necessary solely for the purpose of obtaining a larger ΔT_{ov} , even when the concentrated solution exhibits boiling point elevation. Vacuum also ensures that heat sensitive products and feeds do not decompose or alter during evaporation and entails cheaper construction (lower capital cost).

One may use steam at 10 – 15 kg/cm²(g) and take off vapour at 3.5 to 5.5 kg/cm²(g) if there is a use for such vapour in the plant but such cases are not common

Influence of feed, steam and condensate temperature

The feed temperature usually has no effect on evaporator calculation. This arises because the volume of liquid in evaporator is very large compared to the amount/rate of feed addition and the liquid is always at the final concentration at the boiling point of the final solution. Moderate amounts of superheat in the steam used for heating and subcooling of the condensate also has negligible effect on the mean temperature of steam. Therefore, temperature of feed, temperature of condensate and (in most practical cases) any possible superheating of steam are neglected in evaporator design calculations and a practical working temperature difference is the temperature difference between temperature of saturated steam and boiling liquid.

However, if the feed is at a temperature much below its boiling point and the amount of evaporation is small, there would be areas where the liquid temperature is below the temperature of the thick liquor. Since it is impossible to estimate the extent of these areas and their mean temperature in advance, the only practical temperature that can be used in calculations is the boiling point of thick liquor. Therefore, it is desirable to use a feed preheater to heat the feed to nearly its boiling point in the evaporator.

6.6.4 Design algorithm for Multiple-Effect Evaporator

The evaporator schematic is depicted in Figure 6.23 with key variables shown. Steam is supplied to the first effect on the left and the vapour boil up in effects are used to heat the

subsequent effect. The final vapour stream is condensed. To compute the energy balances for the units, information on enthalpy and boiling point elevation are required. Heat transfer coefficients or correlations are required for each effect. The pressure, at which the third effect is operated, often at vacuum, must be known.

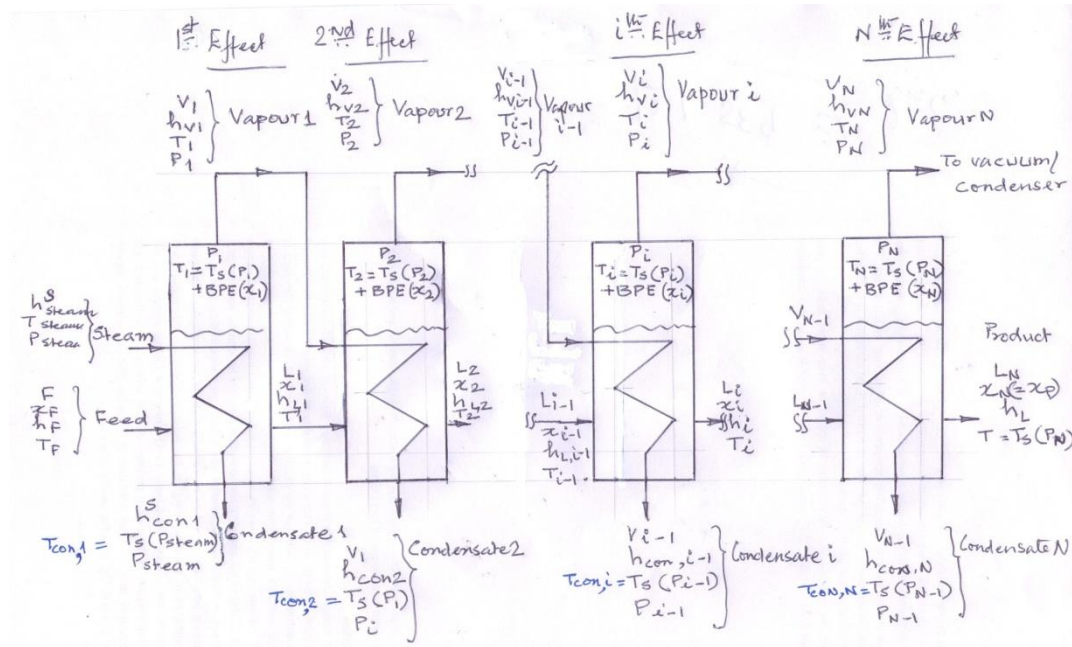


Figure 6.23: Generalised configuration of N-effect forward feed evaporator

[Note for redrawing – all terms L,V, subscripts etc on drawing, except “Feed”, “Vapour”, “Steam”, “Effect”, “Product” and “Condensate” need to be in *italics*]

The following convention is used in Figure 6.23 -

- F : feed mass flow rate entering Effect 1
- x_F : mass fraction of solute in F
- T_F : temperature of feed stream
- h_F : enthalpy of F at T_F
- S : mass flow rate of heating steam entering 1st Effect
- P_{steam}, T_{steam} : temperature and pressure of S
- $h_{con,steam}$: enthalpy of condensate (saturated) water at pressure P_{steam}
- $h_{con,i}$: enthalpy of condensate (saturated) water at pressure P_i
- L_i, V_i : liquid and vapour mass flow rate leaving i^{th} Effect
- T_i, P_i : temperature and pressure in i^{th} Effect
- h_{Li}, h_{Vi} : enthalpy of L_i and V_i at T_i, P_i
- x_i : mass fraction of solute in L_i
- x_P : mass fraction of solute in product liquid (L_N) leaving N^{th} Effect (last effect)
- BPE_i : boiling point elevation in i^{th} Effect with respect to saturation temperature of pure solvent ($T_s(P_i)$) in the same Effect

Design input

Process design of evaporator with N effects starts with known

- (i) feed rate (F), feed conditions (x_F , T_F) and desired product concentration (x_P)
- (ii) Pressure (P_N) in N^{th} stage (last effect).
- (iii) System properties – calculation procedure and / or tables for saturation temperature of solvent (water: steam tables), BPE, enthalpy of solution and solution properties (may be available in the form of Figure 6.22 as already discussed).
- (iv) Steam supply pressure and temperature (P_{steam} and T_{steam}). Usually T_{steam} will have a few degrees of superheat.
- (v) Evaporator type & configuration and the estimated heat transfer coefficient of each effect (U_1 , U_2 , ..., U_N). This is usually known from past experience of design and operation. It is important to note that for evaporators, U_i is based on the temperature difference of condensing vapour/steam in the steam chest ($T_s(P_{i-1})$) and the (saturation) temperature of vapour leaving i^{th} effect. In case of the 1st effect, $T_s(P_{i-1}) = T_s(P_{steam})$.

Process design output for the chosen type of multiple-effect evaporator includes–

- (i) Heating steam requirement – estimated from mass and energy balance
- (ii) Heating surface requirement – estimated from heat transfer equations
- (iii) Estimated temperature in each effect
- (iv) Amount of vapour leaving the last effect and going to the condenser

The design solution involves trial and error.

Design objective

Meeting the final product concentration with high steam economy is the foremost objective. However, the multiple effect evaporator design also aims at arriving at a design with (nearly) same heat transfer area for each effect. This is an indirect way to economise on fabrication cost and minimise inventory cost by holding common spares.

Design deliverables

To summarise, process design output gives the heat transfer areas (A_i , $i=1, \dots, N$) and also includes values of all variables shown on Figure 6.24, whose values are not specified as design input.

Design algorithm

- i) Note the inputs: N , F , x_F , T_F , P_N , U_1 , U_2 , ..., U_N
- ii) Find the value of $T_s(P_N)$ from steam table
- iii) $L_N = F \times x_F / x_P$
- iv) $V_{total} = F - L_N$

v) Assume: V_i , $i=1,...,N-1$; Calculate $V_N = V_{total} - \sum_{i=1}^{N-1} V_i$; Initial guess can be $V_i = (F - L_N) / N$, for $i=1,...,N$.

vi) $L_1 = F - V_1$ and $L_i = L_{i-1} - V_i$ for $i=2,...,N-1$

vii) $BPE_i = BPE(x_i)$, for $i=1,...,N$ – this can be found from the Duhring plots (Figure 6.20) that are linear and pass through the origin.

viii) $\Sigma BPE = \sum_{i=1}^N BPE(x_i)$

ix) $\Delta T_{ov} = T_s(P_{steam}) - T_s(P_N)$

and for all the effects put together $\Delta T_{eff, total} = \Delta T_{ov} - \Sigma BPE$

x) Assume: $\Delta T_{eff, i}$, $i=1,...,N-1$; $\Delta T_{eff, N} = \Delta T_{eff, total} - \sum_{i=1}^{N-1} \Delta T_{eff, i}$;

Initial guess can be from assuming same heat transfer rate in each effect which

$$\text{gives } \Delta T_{eff, i} = \frac{\Delta T_{eff, total}}{\sum_{i=1}^N (1/U_i)} \times (1/U_i)$$

xi) $T_N = T_s(P_N) + BPE(x_N)$;

$$T_{con,1} = T_s(P_{steam})$$

$$T_{con,i} = T_s(P_{i-1}) = T_i + \Delta T_{eff, i} \text{ and } T_{i-1} = T_{con,i} + BPE(x_{i-1}), i=N,...,2$$

xii) $S = \{V_1 \times h_{v,1}(T_1) + h_{con, steam} - F \times h_F(T_F)\} / h_{steam}(T_{steam}, P_{steam})$

$$h_F = h_F(T_F)$$

$$h_{v,i} = h_s(T_{con,i+1}) + C_{p, steam} \times (T_i - T_{con,i+1}), i=1,...,N-1$$

$$h_{v,N} = h_s(T_{con,N-1}) + C_{p, steam} \times (T_N - T_s(P_N))$$

$h_{con,i} = h_w(T_{con,i})$, $i=1,...,N-1$, [$h_w(T_{con,i})$ stands for saturated water enthalpy at $T_{con,i}$ obtained from steam table.]

$h_{L,i} = h_L(x_i, T_i)$, $i=1,...,N$, [Enthalpy of solution with solute concentration x_i at T_i from table/ chart similar to Fig. 6.22 applicable for the system.]

xiii) Calculate heat load and area required for each effect

$$q_1 = S \times (h_{steam}(T_{steam}, P_{steam}) - T_{con,1})$$

$$q_i = V_{i-1} \times h_{v,i-1}(T_{con,i}), i=2,...,N$$

$$A_i = q_i / (U_i \times \Delta T_{eff, i})$$

If $\max(\text{ABS}(A_i - \sum_{i=1}^N (A_i) / N)) > 0.05, i=1, \dots, N$, then

$$\text{re-estimate } \Delta T_{eff,i} = \frac{\Delta T_{eff}}{\sum_{i=1}^N (1 / U_i)} \times (q_i / U_i), \text{ go to step } xi$$

- xiv) Recalculate V from the following set of N equations formed by energy balance around each effect -

1st Effect

$$V_1 = \{F \times (h_F - h_{L,1}) + S \times (h_{steam}(T_{steam}, P_{steam}) - h_{con,1})\} / (h_{V,1} - h_{L,1})$$

2nd to N^{th} Effect

$$V_i = \{L_{i-1} \times (h_{L,i-1} - h_{L,i}) + V_{i-1} \times (h_{V,i-1} - h_{con,i})\} / (h_{V,i} - h_{L,i}), i=1, \dots, N$$

- xv) Check validity of V_i estimates in step v

If $\max(\text{ABS}(V_i - \sum_{i=1}^N (V_i) / N)) > 0.01, i=1, \dots, N$, then re-estimate $V_i =$ values from step (xiv), $i=1, N-1$,

$$V_N = V_{total} - \sum_{i=1}^{N-1} V_i,$$

go to step v.

- xvi) Report –

$$\text{Overall steam economy} = (F - L_N) / S$$

$$1^{\text{st}} \text{ Effect steam economy} = V_1 / S$$

$$i^{\text{th}} \text{ Effect steam economy} = V_i / V_{i-1}, i=2, \dots, N$$

$$\text{Capacity: Feed processed per unit mass steam required} = F/S$$

$$\text{Product produced per unit mass steam required} = L_N/S$$

6.7 DESIGN ILLUSTRATION

Design Example 1: Problem statement

Design a forward feed triple effect evaporator to concentrate 14% w/w caustic soda solution to a product with 40% w/w NaOH. Feed liquor is available at 6 kg/sec at 75°C. Last effect of the evaporator can be connected to an existing vacuum system and operated at a pressure of 7 kPa (abs). Steam with negligible superheat is available at 120°C. Estimated overall heat transfer coefficients for the effects are 3000, 2000 and 1250 W/(m²C), defined with respect to the difference in temperature of the liquid in the effect and the condensing temperature of steam/vapour heating it.

Process Design deliverables

- Steam consumption (S kg/sec)
- Heat load (q W); Heat transfer area (A m²); Operating temperature (T_i °C) and pressure (P_i kg/cm²(g)) and Condensate temperature ($T_{con,i}$ °C) in each effect.
- Flows rate of all streams (S, V, L kg/sec);
Concentration of NaOH in each liquor stream (x fraction w/w)
- Enthalpy of all streams ($h_s, h_v, h_{con}, h_F, h_L$ kJ/kg)

Data

$N=3$; $F=6$ kg/sec; $x_F=0.14$; $x_3=0.4$; $T_F=75^\circ\text{C}$;

$P_3=7\text{kPa}=7/101.3-1.03323=-0.964$ kg/cm² (g)

BPE table for NaOH

x, Fraction, w/w	0	0.1	0.2	0.3	0.35	0.4	0.45	0.5	0.55	0.6	0.65	0.7
BPE, °C	0	2.2	7.8	13.9	19.4	26.1	36.1	41.7	48.9	55.6	66.7	76.7

$T_{steam}=120^\circ\text{C}$;

$U_1=3000$; $U_2=2000$; $U_3=1250$;

P_{steam} =Saturation pressure corresponding to ($T_{steam} = 120^\circ\text{C}$) = 0.9926 kg/cm² (g)

$T_{con,4}$ = Saturation temperature of steam at pressure $P_3=38.4^\circ\text{C}$

$L_3=Fx_F/x_3=2.1$; $V_{total} = F-L_3=3.9$

Trial 1: V_1, V_2, V_3 are assumed equal to one third of V_{total}

$V_1=V_{total}/3=3.9/3=1.3$; $V_2=1.3$; $V_3=1.3$;

Step - A

$L_1 = F - V_1 = 6 - 1.3 = 4.7$; $L_2 = L_1 - V_2 = 4.7 - 1.3 = 3.4$;

$x_1 = Fx_F/L_1 = 0.1787$; $x_2 = Fx_F/L_2 = 6(0.14/3.4) = 0.2471$;

Estimating BPE corresponding to x_1, x_2 and x_3 in the effects from BPE table, in °C: 6.49, 10.4 and 26.1.

Total BPE=42.99

Total $\Delta T_{av} = 120 - 38.4 = 81.6^\circ\text{C}$

$\Delta T_{eff, total} = 81.6 - 42.99 = 38.61^\circ\text{C}$

Estimating individual contribution by assuming same q and A_i in all effects

$$\Delta T_{eff,1} = 38.61 \times (1/3000) / (1/3000 + 1/2000 + 1/1250) = 7.89$$

$$\Delta T_{eff,2} = 38.61 \times (1/2000) / (1/3000 + 1/2000 + 1/1250) = 11.82$$

$$\Delta T_{eff,3} = 38.61 - 7.89 - 11.82 = 18.9$$

$$T_3 = T_{con,4} + BPE_3 = 38.4 + 26.1 = 64.5$$

$$T_{con,3} = 64.5 + 18.9 = 83.4$$

$$T_2 = T_{con,3} + BPE_2 = 83.4 + 10.4 = 93.8$$

$$T_{con,2} = 93.8 + 11.82 = 105.62$$

$$T_1 = T_{con,2} + BPE_1 = 105.62 + 6.49 = 112.11$$

$$T_{con,1} = \text{Saturation temperature of steam} = 120^\circ\text{C}$$

Estimation of Effect pressure (g)

$$P_1 = \text{saturation pressure corresponding to } 93.8^\circ\text{C} = 0.2263 \text{ kg/cm}^2(\text{g})$$

$$P_2 = \text{saturation pressure corresponding to } 83.4^\circ\text{C} = 0.4792 \text{ kg/cm}^2(\text{g})$$

Enthalpies

$$h_{steam}(T_{steam}, P_{steam}) = 2705.9; \quad h_{con,1} = h_{sat \text{ steam}}(T_{con1}) = 503.9$$

$$h_{V,1} = h_{steam}(T_1, P_1) = 2697.1; \quad h_{con,2} = h_{sat \text{ steam}}(T_{con2}) = 442.9$$

$$h_{V,2} = h_{steam}(T_2, P_2) = 2669.5; \quad h_{con,3} = h_{sat \text{ steam}}(T_{con3}) = 349.3$$

$$h_{V,3} = h_{steam}(T_3, P_3) = 2676.1;$$

Enthalpy of solution found by neglecting heat of solution and considering NaOH (Specific heat of solid = 1.4915 kJ/kg) to have enthalpy 'zero' at 0°C .

$$h_F = h_F(75, 0.14) = 285.7;$$

$$h_{L,1}(112.11, 0.1787) = 416.2;$$

$$h_{L,2}(93.8, 0.2471) = 330.5;$$

$$h_{L,3}(83.4, .4) = 200.5;$$

$$F_{steam} = (V_1 h_{V,1} + L_1 h_{L,1} - F h_F) / (h_{steam} - h_{con,1}) = 1.703$$

$$q_1 = F_{steam} \cdot (h_{steam} - h_{con,1}) = 1.701 \times (2705.9 - 503.9) = 3735.5$$

$$q_2 = V_1 \cdot (h_{V,1} - h_{con,2}) = 2931.5$$

$$q_3 = V_2 \cdot (h_{V,2} - h_{con,3}) = 3016.4$$

$$A_1 = q_1 / [U_1 (T_{con,1} - T_1)] = 0.1580$$

$$A_2 = q_2 / [U_2 (T_{con2} - T_2)] = 0.1240$$

$$A_3 = q_3 / [U_3 (T_{con,3} - T_3)] = 0.1276$$

$$A_{avg} = (A_1 + A_2 + A_3) / 3 = 0.1365$$

$$\% \text{ deviation from } A_{avg} = -5.2433, \quad 3.0596 \text{ and } 2.1837$$

Recalculating $\Delta T_{eff,1}$, $\Delta T_{eff,2}$ and $\Delta T_{eff,3}$

$$\sum_{i=1,3} (q_i / U_i) = 5.124$$

$$\Delta T_{eff,1} = \Delta T_{eff, \text{total}} \times (q_1 / U_1) / 5.124 = 9.3829$$

$$\Delta T_{eff,2} = \Delta T_{eff, \text{total}} \times (q_2 / U_2) / 5.124 = 11.0451$$

$$\Delta T_{eff,3} = \Delta T_{eff, \text{total}} \times (q_3 / U_3) / 5.124 = 18.1835$$

Recalculated A values with new values of $\Delta T_{eff,1}$, $\Delta T_{eff,2}$, $\Delta T_{eff,3}$: 0.1315, 0.1329 and 0.1328. These are within 5% from A_{avg} .

Re-estimating V values

$$V_1 = (q_1 + F \cdot h_F - L_1 \cdot h_{L,1}) / h_{steam}(T_1, P_1) = 1.3000$$

$$V_2 = (q_2 + L_1 \cdot h_{L,1} - L_2 \cdot h_{L,2}) / h_{V,2} = 1.3269$$

$$V_3 = (q_3 + L_2 \cdot h_{L,2} - L_3 \cdot h_{L,3}) / h_{V,2} = 1.2731$$

Recalculated V_i values differ from corresponding V values by more than 1%. Hence new assumed values for are V are replaced by V_i values and calculations from Step A are repeated till the V_i and V values differ within 1% and the design calculation converges.

The final converged results are –

Parameter	Effect 1	Effect 2	Effect 3
Feed/ Liquor entering, kg/sec	6	4.7	3.3731
Steam/ heating vapour entering, kg/sec	1.6901	1.3	1.3269
Solute in Feed/ Liquor entering, (wt fraction)	0.14	0.1787	0.2490
Vapour leaving, kg/sec	1.3	1.3269	1.2731
Liquor leaving, kg/sec	4.7	3.3731	2.1
Solute in Liquor leaving, (wt fraction)	0.1787	0.2490	0.40
BPE, °C	6.49	10.51	26.1
Effect Temperature, °C	110.7	93.3	64.5
Condensing temperature in steam chest, °C	120	104.2	82.8
Condensing temperature of vapour leaving Effect 3, °C			38.4
U, W/m ⁻² .sec ⁻¹	3000	2000	1250
q,W	3704.7	2936.3	3080.9
Area, m ²	0.1333	0.1346	0.1345
Enthalpy, kJ/kg			
Feed/Liquor entering	285.67	410.98	328.31
Liquor leaving	410.98	328.31	200.46
Vapour leaving	2695.9	2668.8	2675.2
Steam/heating vapour entering	2705.9	2695.9	2668.8
Condensate leaving	503.8	437	346.8