



DESIGN PRACTICES

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Revision Memo

- 12/98** The highlights of this revision include:
- Figure 8** (fluidity curve) added for calculation of overall tray efficiency for heavy hydrocarbon fractionators.
 - Contingency debit for efficiency of valve and V-grid trays reduced to that of sieve and bubble cap trays.
 - Reference added for towers with high viscosity liquids.
 - More extensive description for key components and estimating slope of equilibrium line.
 - More details of estimating the efficiency of bubble cap tray given.
 - References updated for computer programs and PRO/II's PRINT TSLOPE option for printing slope of equilibrium line.
 - Reference made to checking bubble cap, valve, and V-grid trays for entrainment and spray regime operation using sieve tray program.
 - Applicability range description improved.



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SCOPE

This section covers the theory and calculation methods for estimating mass transfer tray efficiency for cross flow type trays, including sieve, bubble cap, valve and Nutter V-grid trays. It also contains guidelines for maximizing tray efficiency in heavy hydrocarbon stripping towers, but is not applicable to reaction/distillation systems. The method is designed for most distillation systems, absorbers, and strippers.

Tower pumparound or other heat transfer sections should be sized according to the heat transfer criteria contained in [Section III-F](#). Dualflow and jet tray efficiency information is contained in [Sections III-N](#) and [III-D](#), respectively.

Equations used only in the text for explanation purposes are labeled alphabetically. Those used in the text as well as in the calculation procedure carry the equation number shown on the TRAY EFFICIENCY CALCULATION FORM (either CUSTOMARY or METRIC). Where both a Customary and Metric value are shown, the metric value will be in parentheses ().

REFERENCES

1. Colwell, C. J., *Tray Efficiency Maximization for Heavy Hydrocarbon Strippers*, ER&E Report EE.20ERL.69, June 27, 1969.
2. Colwell, C. J., *Sieve Tray Efficiency Calculation Methods Have Been Improved*, ER&E Report EE.67E.72, May 1972.
3. Colwell, C. J., *Sour Water Stripper Tray Efficiencies*, 58905, December 18, 1973.
4. Colwell, C. J., *Tray Efficiency Prediction in the Weeping Region is Now Possible*, ER&E Report EE.152E.79, Dec., 1979.
5. Lockett, M. J., *Distillation Tray Fundamentals*, Cambridge University Press, Cambridge, England (1986).
6. Colwell, C. J., *Designing Towers for High Liquid Viscosity*, 96CET41, March 21, 1996.

DEFINITIONS

Tray Geometry - See [Sections III-A](#), [B](#), and [E](#) for definitions and discussions of such parameters as hole diameter, hole area, bubble area, weir length, weir height, and number of passes.

Point Efficiency, E_{OG} - The point efficiency E_{OG} is a measure of the effectiveness of vapor-liquid contacting at a **given point** on a tray. It expresses the actual separation in a differential volume element of tray liquid ("point"), compared to the theoretically obtainable separation at equilibrium:

$$E_{OG} = \frac{\bullet y_n - \bullet y_{(n-1)}}{\bullet y_n^* - \bullet y_{(n-1)}} \quad \text{Eq. (A)}$$

where:

- E_{OG} = Point efficiency, dimensionless
- $\bullet y_n$ = Mole fraction of a component in the vapor leaving a given point on tray n
- $\bullet y_{(n-1)}$ = Mole fraction of a component in the vapor entering a given point on tray n (coming from tray n-1)
- $\bullet y_n^*$ = Mole fraction of a component in the vapor which would be in equilibrium with the liquid at a given point on tray n

- ♦ **Design Point Efficiency, E_{OG}^*** - For **design purposes**, a 10% contingency is applied to the calculated point efficiency. Thus:

$$E_{OG}^* = 0.9 E_{OG} \quad \text{Eq. (10c1)}$$

The contingency factor of 0.9 is intended primarily for new tower designs, to insure that the design fractionation will be achieved or exceeded in 9 out of 10 cases. In many situations, such as revamps, towers with good efficiency backup data, or towers in which the degree of fractionation is not critical, the designer has the option of selecting a more appropriate contingency factor from [Table 3](#). Where fractionation is critical or for a revamp where tray type is being changed (for example, sieve trays being replaced by valve trays) consultation with a FRACTIONATION SPECIALIST is advisable.

Tray Efficiency, E_{MV} - The tray efficiency E_{MV} is a measure of the contacting effectiveness of an entire tray. It is defined in terms of average compositions to and from the whole tray:

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DEFINITIONS (Cont)

$$E_{MV} = \frac{y_n - y_{(n-1)}}{y_n^* - y_{(n-1)}} \quad \text{Eq. (B)}$$

where: E_{MV} = Tray (Murphree, vapor) efficiency, dimensionless
 y_n = Average mole fraction of a component in the total vapor leaving tray n
 $y_{(n-1)}$ = Average mole fraction of a component in the total vapor entering tray n (coming from tray n-1)
 y_n^* = Mole fraction of a component in the vapor which would be in equilibrium with the liquid leaving tray n

E_{MV} is related to E_{OG} (or to E_{OG}^*) by [Eq. \(11d1\)](#).

Overall Efficiency, E_O - The overall efficiency, E_O , is a measure of the effectiveness of an entire tower or section of a tower. This efficiency is the one most often used by designers in determining the number of actual trays to provide. E_O is simply the total number of calculated theoretical trays required, divided by the total number of actual trays required for the separation. E_O is related to E_{MV} by [Eq. \(13b1\)](#).

Equilibrium Slope - The slope, m , of the equilibrium curve is defined as the slope of the y^* vs. x plot for a given component in the tower. This parameter is required to determine both the point efficiency and the effect of liquid mixing.

Key Components - Before making efficiency calculations, the designer must specify the light and heavy key components in a multicomponent distillation system. These two components characterize the split between components in the fractionation operation. The light key component is defined as the least volatile component in the tower or tower section whose concentration increases up the tower. The heavy key component is the most volatile component whose concentration decreases up the tower.

For strippers, the designer often only needs to specify one component, the light key. For absorbers, often only the heavy key component should be specified.

Mixing Pools - Lateral liquid mixing on a tray is characterized by a number of hypothetical mixing pools, n , which represents an equivalent number of completely mixed stages through which the liquid flows. See [Eq. \(11b1\)](#).

Flow Path Length - The number of mixing pools is related to the distance through which liquid flows as it crosses a tray. The flow path length, l_{fp} , is normally taken as the distance between the edge of the downcomer and the outlet weir. See [Figure 12](#) in [Section III-A](#). The width of any blanked perforated area should be subtracted from this distance. In 2, 3, and 4-pass trays, where the flow path length may not be the same for inboard and outboard passes, the minimum flow path length should be used in the efficiency calculations.

Clear Liquid Height - Clear liquid height, h_c , along with average froth height, h_f , and froth density, ψ , are intermediate parameters in the calculation of liquid and vapor residence times on the tray. Clear liquid height is defined as the liquid height that would result if the froth on the tray were collapsed into a deaerated layer. Procedures to estimate these parameters are provided in [Section III-B](#).

Average Froth Height - The average froth height, h_f , is a measure of the **effective** contacting height of the vapor-liquid mixture on a tray. This value is usually somewhat lower than the spray height (i.e., the highest level attained by any of the liquid droplets on the tray).

Froth Density - Froth density, ψ , is defined as the volume fraction of the vapor-liquid mixture occupied by liquid. It is related to clear liquid height (h_c) and average froth height (h_f) by the following equation:

$$\psi = \frac{h_c}{h_f} \quad \text{Eq. (C)}$$

GENERAL

BASIC INFORMATION REQUIRED

The calculation method is based on predicting point efficiencies, which are then scaled up to tray efficiencies and overall tower efficiencies. In order to predict point efficiency anywhere in a tower, the designer should preferably have a plate-to-plate printout, some physical property information, and a detailed tray design summary (such as that obtained from Program 1133, Sieve Tray Design) or from hand calculations based on the CALCULATION FORMS in [Section III-B](#) or [III-E](#).

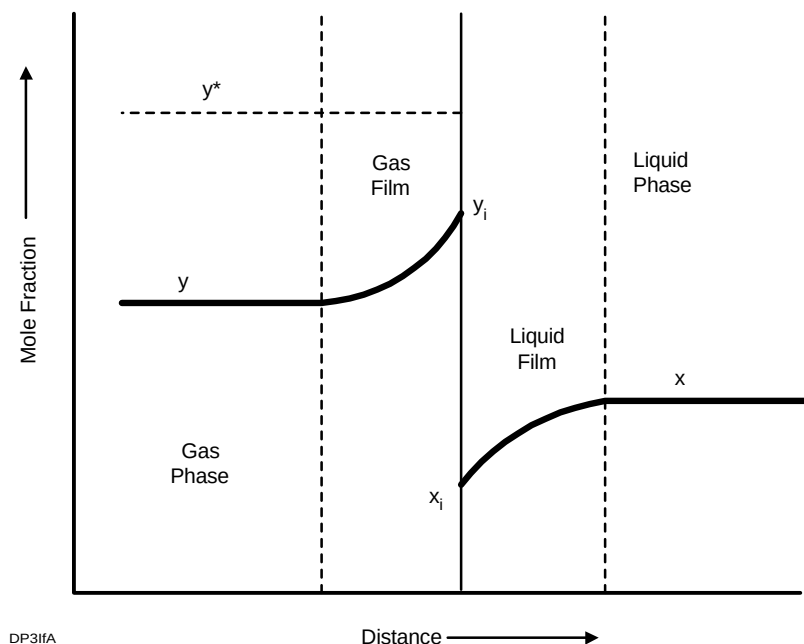
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MASS TRANSFER THEORY

The efficiency prediction method given in this section is based on the two-resistance theory of mass transfer, which postulates rate mechanisms through two small films, one on each side of the vapor-liquid interface. This is shown in **Figure A** below.

FIGURE A
TWO RESISTANCE THEORY



TRANSFER UNITS

The transfer mechanism in each of these films is characterized by two basic rate parameters known as the vapor and liquid phase mass transfer units (N_G and N_L , respectively). They are composite functions of the mass transfer coefficients, interfacial area, and residence times according to the following equations:

$$N_G = k_G a t_G \quad \text{Eq. (8a1)}$$

$$N_L = \frac{k_L a t_L}{\psi} \quad \text{Eq. (8b1)}$$

where: N_G = Vapor phase transfer unit, dimensionless
 N_L = Liquid phase transfer unit, dimensionless
 k_G = Vapor phase mass transfer coefficient, cm/s (mm/s)
 k_L = Liquid phase mass transfer coefficient, cm/s (mm/s)
 a = Interfacial area, cm²/cm³ (mm²/mm³)
 t_G = Vapor residence time, s
 t_L = Liquid residence time, s

The overall mass transfer unit N_{OG} is a measure of the combined mass transfer in both the vapor and liquid phases. In terms of effective resistances to mass transfer, **Eq. (10b1)** shows the total resistance to mass transfer, characterized by $1 / N_{OG}$, to be the sum of the vapor phase resistance, $1 / N_G$, and the liquid phase resistance, λ / N_L :

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$$\frac{1}{N_{OG}} = \frac{1}{N_G} + \frac{\lambda}{N_L} \quad \text{Eq. (10b1)}$$

where: N_{OG} = Overall (vapor) mass transfer unit, dimensionless
 N_G, N_L = As defined earlier for **Eq. (8a1)** or **(8b1)**

$$\lambda = m \frac{G_m}{L_m}, \text{ and} \quad \text{Eq. (9b1)}$$

m = Slope of the equilibrium curve

G_m = Vapor rate, moles/h (kmol/s)

L_m = Liquid rate, moles/h (kmol/s)

The point efficiency (E_{OG}), in turn, is directly related to the overall transfer unit:

$$N_{OG} = -\ln(1 - E_{OG}), \text{ or} \quad \text{Eq. (D)}$$

$$E_{OG} = 1 - (e)^{-N_{OG}} \quad \text{Eq. (10b3)}$$

CALCULATING LIQUID AND VAPOR PHASE RESISTANCES

To obtain N_{OG} , it is necessary to calculate N_G and N_L . This is accomplished by first calculating mass transfer coefficients, interfacial area, residence times, and froth density and then applying **Eqs. (8a1)** and **(8b1)** discussed earlier. Correlations for these values are provided as follows:

Clear Liquid Height, Average Froth Height, Froth Density - These parameters are required as intermediate steps for determining residence times. They are available from an 1133 Sieve Tray Computer program printout or from hand calculations using the Calculation Form of **Section III-B** or **III-E**. A more rigorous clear liquid height method is used for efficiency calculations in the Sieve Tray Design computer program (1133) and is outlined in **Table 3A** or **3B** of **Section III-B**. Values obtained via the rigorous method may differ by as much as 10% from those obtained by the approximate method contained on the Calculation Form of **Section III-B**.

Residence Times - Liquid (t_L) and vapor (t_G) residence times are related to clear liquid height and average froth height by the following equations:

Customary

$$t_L = \frac{37.4 A_b h_c}{Q_L}$$

$$t_G = \frac{h_f}{12 V_b}$$

Metric

$$t_L = \frac{A_b h_c}{Q_L} \quad \text{Eq. (4b1)}$$

$$t_G = \frac{(0.001) h_f}{V_b} \quad \text{Eq. (4c1)}$$

where: t_L = Liquid residence time, s
 t_G = Vapor residence time, s
 h_c = Clear liquid height, in. (mm)
 h_f = Average froth height, in. (mm)
 A_b = Bubble area, ft² (m²)
 Q_L = Liquid rate, gpm (dm³/s)
 V_b = Vapor velocity based on bubble area, ft/s (m/s)

Efficiency Prediction When Liquid Residence Time is Low - When liquid moves rapidly across the tray, there is a tendency to sweep vapor flow away from the contacting space available at the liquid inlet side of the tray. Thus, the space above the tray bubbling area, A_b , is not fully utilized, and tray efficiency is reduced as a result. This effect is accounted for by incorporating a factor that reduces the bubble area used in the efficiency prediction calculations. The factor is applied only if the ratio of vapor to liquid residence time, t_G / t_L exceeds 0.5. When this condition occurs, the corrected bubble area, (A_b^*), is calculated from the following equation:

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$$A^*_b = A_b \left[1 - \left[\frac{1-\psi}{2} \right] \frac{t_G}{t_L} \right] \quad \text{Eq. (5b1)}$$

Mass Transfer Coefficients - The two equations below provide empirical correlations for gas and liquid phase mass transfer units. These relationships are also given in **Figures 1 and 2**.

$$k_G = C_2 [V_b]^{0.76} [h_{w0}]^{-0.21} \left[\frac{A_o}{A_b} \right]^{-0.084} \quad \text{Eq. (6a1)}$$

If tray has no outlet weir, set $h_{w0} = 0.25$ in. (6 mm)

$$k_L = \frac{C_3}{[\rho_L]^{1/3} [\mu_L]^{0.42}} \quad \text{Eq. (6b1)}$$

where:

- C_2 = 0.649 (Customary); 31.6 (Metric)
- C_3 = 0.158 (Customary); 4.0 (Metric)
- k_G = Gas phase mass transfer coefficient, cm/s (mm/s)
- k_L = Liquid phase mass transfer coefficient, cm/s (mm/s)
- V_b = Vapor velocity based on bubble area, ft/s (m/s)
- h_{w0} = Outlet weir height, in. (mm)
- A_o / A_b = Ratio of total hole area to bubble area, expressed as a fraction
- ρ_L = Liquid density, lb/ft³ (kg/m³)
- μ_L = Liquid viscosity, cP (mPa·s)

The k_L equation [Eq. (6b1) above] was obtained by inserting an approximate liquid diffusivity relationship (obtained from Fractionation Research, Inc.) given in Eq. (E) below into the more rigorous k_L correlation shown in Eq. (F).

Customary

$$D_L = \frac{12.1 \times 10^{-4}}{\rho_L [\mu_L]^{0.5}}$$

$$k_L = 4.53 \left[\frac{\rho_L}{\mu_L} \right]^{1/6} [D_L]^{0.5}$$

Metric

$$D_L = \frac{1.94}{\rho_L [\mu_L]^{0.5}} \quad \text{Eq. (E)}$$

$$k_L = 2.85 \left[\frac{\rho_L}{\mu_L} \right]^{1/6} [D_L]^{0.5} \quad \text{Eq. (F)}$$

where: D_L = Liquid molecular diffusivity, cm²/s (mm²/s)

Because of this approximation, Eq. (E) should not be used for k_L on systems with greater than 75% liquid phase control. For these cases, D_L may be obtained from Section III-G, Appendix II and k_L determined rigorously from Eq. (F) above.

Interfacial Area - This parameter has been correlated empirically for sieve trays. See Figure 3 or Eq. (7a1) below:

$$a = C_4 (1 - C_5 F_b) \quad \text{Eq. (7a1)}$$

where:

- a = Interfacial area, cm²/cm³ (mm²/mm³)
- C_4 = 4.54 Customary; 0.454 Metric
- C_5 = 0.1 Customary; 0.082 Metric
- F_b = $V_b [\rho_v]^{0.5} = F$ -factor based on bubble area, ft/s (lb/ft³)^{0.5}, (m/s) (kg/m³)^{0.5}
- V_b = Vapor velocity based on bubble area, ft/s (m/s)
- ρ_v = Vapor density, lb/ft³ (kg/m³)

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For aqueous systems, or when the surface tension exceeds 40 dyne/cm (mN/m), a factor (R_a) to correct interfacial area may be needed. This applies when:

Customary

$$F_0 \leq 14$$

Metric

$$F_0 \leq 17$$

where: $F_0 = V_0 (\rho_v)^{0.5}$ = F-factor based on hole area, ft/s (lb/ft³)^{0.5}, (m/s) (kg/m³)^{0.5}

If true, then the correction factor, R_a , is calculated by:

Customary

$$R_a = 1 - 0.06 (14 - F_0)$$

Metric

$$R_a = 1 - 0.05 (17 - F_0)$$

Eq. (7b1)

The interfacial area, a , from **Eq. (7a1)** is then multiplied by R_a . That is,

$$a_{(\text{corrected})} = a (R_a)$$

EFFICIENCY RELATIONSHIPS

After the point efficiency E_{OG} has been determined from **Figure 4** or **Eq. (10b3)**, the designer is ready to determine the tray efficiency E_{MV}^* and the overall efficiency E_O . On the actual tray, the composition of the liquid phase changes as the liquid flows from the inlet to the outlet side of the tray. The contacting effectiveness of the tray depends on the magnitude of this liquid concentration gradient. This, in turn, is characterized by a number of mixing pools, n , arranged in series, through which the liquid flows. The tray efficiency is related to the point efficiency in terms of the number of mixing pools by **Eq. (11d2)**, which is also presented graphically in **Figure 5**.

$$E_{MV} = \frac{\left[1 + \frac{\lambda E_{OG}^*}{n} \right]^n - 1}{\lambda} \quad \text{Eq. (11d2)}$$

where: E_{MV} = Tray (Murphree, vapor efficiency), dimensionless, not yet corrected for vapor mixing or weeping
 λ = Defined by **Eq. (9b1)**
 E_{OG} = Point efficiency, dimensionless. For **design**, use E_{OG}^* . See **Eq. (10c1)** discussed under **DEFINITIONS**.
 n = Number of mixing pools

Customary

Metric

$$n = 465 \frac{[l_{fp}]^2}{D_E t_L} + 1$$

$$= \frac{[l_{fp}]^2}{2 D_E t_L} + 1 \quad \text{Eq. (11b1)}$$

where: l_{fp} = Length of liquid flow path on the tray, ft (mm) (calculated from information given in tray geometry printout, Deck 1133, or the Calculation Form in **Section III-B**)
 t_L = Average liquid residence time, s [see **Eq. (4b1)**]
 D_E = Liquid eddy diffusivity, cm²/s (mm²/s)
= 0.635 (1/ ψ)³ h_c (Customary)
= 2.5 (1/ ψ)³ h_c (Metric)

Eq. (11a1)

And where:

ψ = Froth density, dimensionless
 h_c = Clear liquid height, in. (mm)

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The limiting flow situations for the mixing pool concept are: (a) perfect mixing on the tray, in which case $n = 1$ and $E_{MV} = E_{OG}$, and (b) plug flow across the tray, in which case $n = \infty$ and E_{MV} exceeds E_{OG} by the greatest amount theoretically possible. Experimental data indicate that where the flow path length is greater than about 2.5 ft (750 mm), the liquid is essentially in plug flow. The effect of mixing in the vapor phase on the tray efficiency is smaller than the effect of liquid mixing, but should nevertheless be taken into account. **Figure 6** gives the correction for vapor mixing, which should be applied to the E_{MV} calculated by **Eq. (11d1)**, to give E_{MV}^* , the corrected tray efficiency.

Once E_{MV}^* has been calculated, the overall efficiency, E_O , can be determined from **Figure 7** or **Eq. (13b1)**, which is based on the assumption that both E_{MV}^* and λ are constant. Actually E_{MV}^* and λ are usually not constant throughout the tower, but **Eq. (13b1)** gives acceptable results for most cases when E_O is calculated at an average tray in each tower section (see *Basic Design Considerations*).

$$E_O = \frac{\ln [1 + E_{MV}^* (\lambda - 1)]}{\ln \lambda} \quad \text{Eq. (13b1)}$$

where: E_O = Overall efficiency, dimensionless
 Other terms are as previously defined

Eq. (13b1) is presented graphically in **Figure 7**.

Efficiency Prediction in the Weeping Region - Calculation of efficiency under weeping conditions requires an estimate of fractional weepage, f_w , from **Section III-B**. A fundamental correction is then applied to account for the effect of bypassing partially uncontacted liquid from one tray to the tray below. Also, the number of mixing pools, n , is reduced. Note that some efficiency debit is already incurred within the basic mass transfer parameters via reduced clear liquid height and vapor/liquid residence times. This is because clear liquid height is set only by the portion of liquid flowing over the weir and does not include the portion that weeps. The effect of liquid bypassing is applied to the tray efficiency, E_{MV} , before the vapor mixing correction and final overall efficiency, E_O , are calculated. The following equations express how the weeping-corrected tray efficiency, E_{MVa} , relates to E_{MV} :

$$E_{MVa} = \frac{\left[\left[1 + \frac{f_w E_{MV}}{2} \right]^2 + 2 f_w (\lambda - 1) E_{MV} \right]^{0.5} - \left[1 + \frac{f_w E_{MV}}{2} \right]}{f_w (\lambda - 1)} \quad \text{Eq. (11e1)}$$

when: $\lambda = 1$

$$E_{MVa} = \frac{E_{MV}}{1 + \frac{f_w E_{MV}}{2}} \quad \text{Eq. (11e2)}$$

when: $\lambda = 0$

$$E_{MVa} = E_{MV}$$

In addition, the number of mixing pools, n , must be adjusted for weeping as follows:

$$n_{(\text{corrected})} = [n_{(\text{predicted})} - 1] [1 - f_w] + 1 \quad \text{Eq. (11c1)}$$

Since weeping correlations are not available for bubble cap, valve, or V-grid trays, efficiency estimates should neglect weeping and the tray designed only within acceptable turndown limits.

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APPLICATION

The mass transfer efficiency prediction method detailed in this section is restricted to single or multipass sieve, bubble cap, valve, and Nutter V-grid trays with downcomers, and requires that the calculations are based on a reliable plate-to-plate fractionation calculation. The method has been tested for nonhydrocarbon, as well as hydrocarbon systems over a wide range of tray design parameters and tower loadings.

The efficiency calculation procedure is not applicable to systems with combined reaction/distillation which should be discussed with your FRACTIONATION SPECIALIST. For atmospheric and vacuum pipestills and cat and coker fractionators, efficiencies can be obtained directly from **Figure 8**. For absorbers or regenerators with amine or FLEXSORB™ solvents in gas treating processes to remove H₂S, CO₂, or mercaptans, contact the Gas Processes Section of ER&E. For sour water strippers see Reference 3.

For sidestream strippers for flash point control or bottoms strippers in pipestills, four to six actual trays are typically used. Refer to the appropriate process manual for design criteria. For example, the Atmospheric Pipestill Guide (EEPE-0006, February, 1996) and the Fuels Vacuum Pipestill Guide (EETD-0076, February, 1996).

Table 1 recommends acceptable ranges for various design parameters, which should be followed for calculating design efficiencies. The designer should consult the FRACTIONATION SPECIALIST for cases that fall outside these acceptable ranges.

Applying the Efficiency Prediction Method to Bubble Cap, Valve and V-Grid Trays - Although the efficiency prediction method was developed for sieve trays, its applicability to bubble cap, valve and Nutter V-grid trays has been demonstrated. For valve and V-grid trays a correction is applied to convert the tray hole area to that of an "equivalent" sieve tray that will have the same dry tray pressure drop as the valve or V-grid tray being designed. The hole size is arbitrarily set to 0.5 in. (13 mm). The following equations show how to correct the valve or V-grid tray hole area:

For valves fully open and V-grid:

$$A_{O(\text{sieve})} = A_{O(\text{valve})} \left[C_6 \frac{J_5}{K_2} \right]^{0.5} \quad \text{Eq. (2a1)}$$

where: C_6 = 0.9 (Customary); 246 (Metric)

J_5 = Pseudo orifice coefficient for sieve trays, dimensionless. See **Section III-B, Figure 13** and/or **Eq. (5a1)** of **Section III-B**. It is used to calculate sieve tray dry tray pressure drop.

K_2 = Dry tray pressure drop coefficient for valve trays with valves fully open. See **Section III-E, Table 3C** for customary and metric values.

For valves partially open:

$$A_{O(\text{sieve})} = A_{O(\text{valve})} \left[C_6 \frac{J_5}{K_1} \right]^{0.5} \left[\frac{1}{1 + \frac{1.35 t_m \rho_m}{K_1 V_{O(\text{valve})}^2 \rho_v}} \right]^{0.5} \quad \text{Eq. (2a2)}$$

where: C_6 = 0.9 (Customary); 246 (Metric)

K_1 = Dry tray pressure drop coefficient for valve trays with valves partially open. See **Section III-E, Table 3C** for customary and metric values.

t_m = Valve metal thickness, in. (mm)

ρ_m = Valve metal density, lb/ft³ (kg/m³)

For bubble cap trays, the efficiency calculations are based on the standard sieve tray correlations using 0.5 in. (13 mm) diameter holes and a total area of holes equivalent to 8% of the bubble cap tray's bubble area.

$$A_{O(\text{sieve})} = (0.08) A_b (\text{bubble cap tray}) \quad \text{Eq. (2a3)}$$

Once the corrected hole area is obtained, this value should be used for the entire efficiency calculation procedure, including the determination of clear liquid height, froth height, and froth density from **Section III-B**.

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BASIC DESIGN CONSIDERATIONS

SELECTING THE TYPE OF EFFICIENCY

The overall efficiency E_O is the one most often used for design purposes. However, two situations require the use of the tray efficiency E_{MV}^* :

- **Plate-to-plate computer programs** which require E_{MV}^* for each tray as input data, and
- **Graphical hand calculation methods**, such as the use of a McCabe-Thiele diagram.

These two approaches are used infrequently since they are tedious.

♦ KEY COMPONENTS

In general, E_{MV}^* varies from component to component and from tray to tray. However, the use of the principle of key components makes it possible to obtain a satisfactory design value of E_O , even though this variation in E_{MV}^* does occur. In a multicomponent system, it is usually sufficient to select two key components on a tray to determine efficiency and then to average these two values. This approach is valid, since there are usually more than enough trays to obtain the desired fractionation of components lighter than the light key and heavier than the heavy key. However, see comments about trace components below.

To determine key components:

1. Arrange the component order in PRO/II by boiling point.
2. For each section of the tower (see SECTIONING THE TOWER below) look at the liquid phase molar concentration profile.
 - a. Determine the least volatile component whose concentration increases up the entire section; this is the light key.
 - b. Determine the most volatile component whose concentration decreases up the entire section; this is the heavy key.

In most absorbers and strippers, it is necessary to determine efficiencies for one component only. The key component in a stripper can often be found as the component where 75% of the component goes overhead, this is, 75% has been stripped out. The key component in an absorber can often be found as the component where 75% of the component has been absorbed.

In some cases, severe specifications must be met on trace components such as H_2S and H_2O . The efficiencies for these components should always be checked in addition to those of the key components, to see if more trays are required to remove these trace components than are needed for the hydrocarbon separation.

SECTIONING THE TOWER

Since the equation for E_O is based on the assumption of constant E_{MV} and constant λ , it is helpful to divide the tower into two or more sections, such that the variations in these two parameters are not too great over a tower section. For any given section, it is usually sufficient to determine E_O at one intermediate tray location within the section.

Therefore, section the tower:

- ♦ **At each feed and drawoff location**, since changes in vapor and liquid loadings at these points affect the value of λ . The feed tray should be counted in the section below, while the drawoff tray should be counted in the section above.

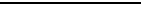
At points where m for the key components changes abruptly - Abrupt changes in the slope of the y^* vs. x equilibrium curve are likely to occur in multicomponent systems. It is therefore recommended that the designer prepare a plot of the y^* vs. x curves for the key components, as a means of spotting concentration reversals or other variations in the slope of the equilibrium curve. These locations should be used as section dividing points, but not as locations for determining efficiencies. Also, wherever a concentration reversal occurs, it is necessary to check for different key components. See the SAMPLE PROBLEM for more details on this point.

For design purposes, the number of actual trays for each section is based on the E_O for that section.

♦ DETERMINING THE SLOPE OF THE EQUILIBRIUM CURVE, m

Eqs. (10b1), (11d2) and (13b1) show the importance of λ , which is directly proportional to m . The value of m used in a tower design should be representative (roughly an average value) of the entire tower section in question. To insure that representative values of m are used, a plot of y^* vs. x for the entire section in question should be prepared and examined. Data for preparing this plot can be obtained from a PRO/II computer printout (PRINT TSLOPE).

In most cases, m can be calculated from the liquid and vapor compositions around the tray in question for the component for which the efficiency is being determined. (Note that in a multicomponent system, m is different for each component.) However, this approximation for m becomes inaccurate at points where the y^* vs. x curve has a large degree of curvature between trays or where there is a reversal in the concentration profile. Efficiency determination at such points should be avoided.

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BASIC DESIGN CONSIDERATIONS (Cont)

Two methods for obtaining m are given below:

Tray Composition Profile - Where the computer printout gives the equilibrium vapor and liquid compositions on each tray the slope m , can be determined by the equations below.

$$m = \frac{dy^*}{dx} \cong \frac{y_{n+1}^* - y_{n-1}^*}{x_{n+1} - x_{n-1}} \quad \text{or} \quad \text{Eq. (9a1)}$$

$$m \cong \frac{y_{n+1}^* - y_n^*}{x_{n+1} - x_n} \quad \text{or} \quad \text{Eq. (G)}$$

$$m \cong \frac{y_n^* - y_{n-1}^*}{x_n - x_{n-1}} \quad \text{Eq. (H)}$$

where: m = Slope of the y^* vs. x curve
 x = Mole fraction of component in the liquid from a tray
 y^* = Mole fraction of component in the vapor in equilibrium with the liquid from a tray
 n = Number of the tray at which m is being determined

This method should be used after the y^* vs. x plot has been examined to insure the tray selected will give a representative slope for the section.

Generally speaking, of these three equations, **Eq. (9a1)** gives the best approximation for m , although the differences in the results obtained from the three equations are usually slight, especially if m is changing very slowly in the particular section of the tower.

Henry's Law Constant (H_e) - At very low concentrations and moderate pressures, where Henry's law is valid.

$$y^* = Kx = \left[\frac{H_e}{P} \right] x \quad \text{Eq. (I)}$$

$$m \cong K = \left[\frac{H_e}{P} \right] \quad \text{Eq. (J)}$$

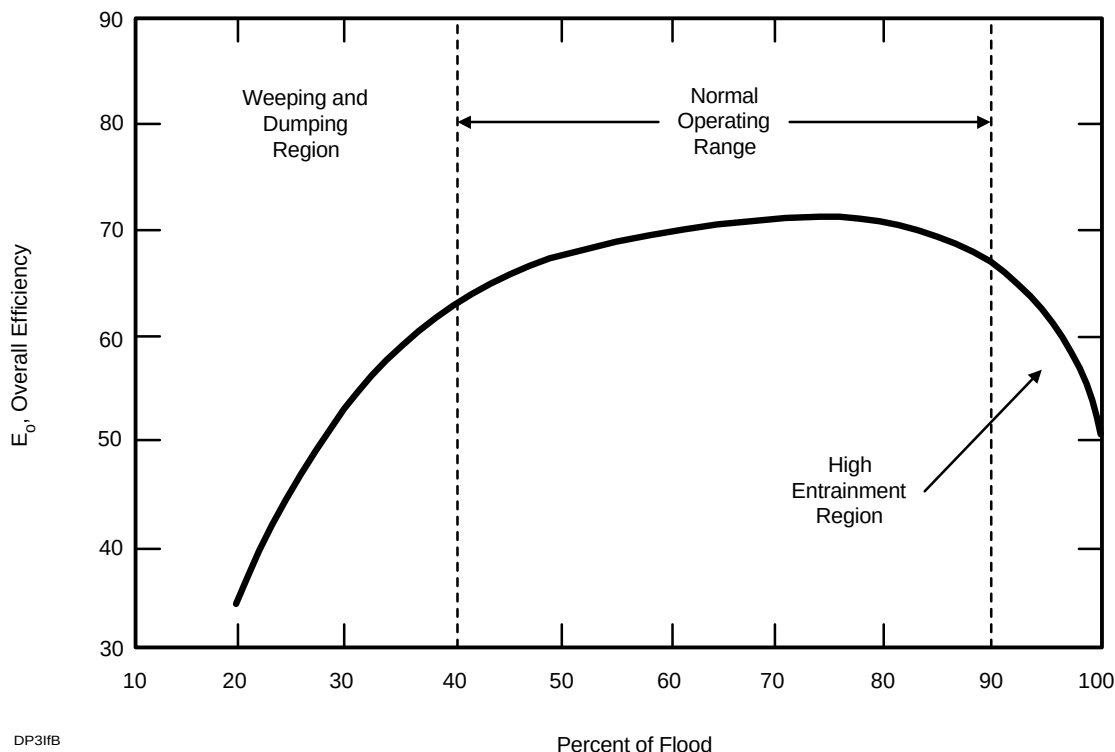
This method is usually valid for determining the equilibrium slopes of trace components.

♦ PERCENT OF FLOOD AND ENTRAINMENT

Trays should be designed to operate at or below 90% of flood. Above 90% of flood, efficiency decreases because of excessive entrainment. See **Figure B**. Since the efficiency prediction method given in this section does not correct for entrainment, the calculated entrainment rate should be kept below 20% to help ensure that design efficiency will be obtained. Since high entrainment rates can occasionally be obtained below 90% of flood, the fractional entrainment rate should always be checked to insure it is less than 20%. Note that this calculation can only be thoroughly done for sieve trays. Since bubble cap, valve and V-grid trays also entrain heavily above 90% of flood they should likewise be kept at or below this value to help ensure design efficiency is obtained. The entrainment (and spray to froth transition) of bubble caps, valve and V-grid trays should be checked in PEGASYS by running a sieve tray rating of the proposed design using 0.5 in. (13 mm) holes and the total hole open area calculated from **Eq. (2a1), (2a2) or (2a3)**. Consult your FRACTIONATION SPECIALIST if warning messages appear.

BASIC DESIGN CONSIDERATIONS (Cont)

FIGURE B
TOWER LOADING EFFECT ON TRAY EFFICIENCY



DESIGN GUIDELINES FOR HEAVY HYDROCARBON STRIPPERS AND TRAYS CONTAINING LARGE BLANKED AREAS

This section presents a number of design guidelines for maximizing tray efficiency in heavy hydrocarbon strippers. These techniques are primarily applicable to bottoms strippers on atmospheric and vacuum pipestills but can also be applied to any tower where loadings are significantly below normal capacity limits. The techniques are applicable to all crossflow type trays.

The efficiency maximization technique consists of maximizing both the number of mixing pools and the liquid residence time on a tray. This is done by any combination of: (a) increasing the flow path length, (b) decreasing the flow width, and/or (c) increasing the outlet weir height.

Under present design procedures, tray efficiencies are normally not calculated for heavy hydrocarbon strippers. Instead, a standard number of actual trays (usually 4 or 6) are specified. It is therefore difficult to quantify the amount of efficiency increase that can be expected. Nevertheless, these techniques are based on fundamental efficiency theory and will directionally improve the degree of fractionation. Thus, application to new designs will not reduce investment costs by lowering the number of trays, but should produce operating credits such as reduced pitch yield, lower penetrations when running asphalt, or a reduction in stripping steam.

WHEN CAN EFFICIENCY BE IMPROVED?

The efficiency maximization procedure outlined below is useful primarily for sections of towers in which the diameter is larger than that required for the given vapor and liquid loadings. This also applies to revamps of existing towers for downsized loadings. Diameters of heavy hydrocarbon strippers are usually oversized for a number of reasons. For example, in both sidestream strippers and bottoms strippers, a significant change in vapor load takes place between the bottom and the top tray. Thus, the tower diameter may be well suited for the top tray but oversized for the trays below. Because of the excess capacity available, downcomers are often made excessively large and/or portions of the tray bubble area are blanked to reduce weeping.

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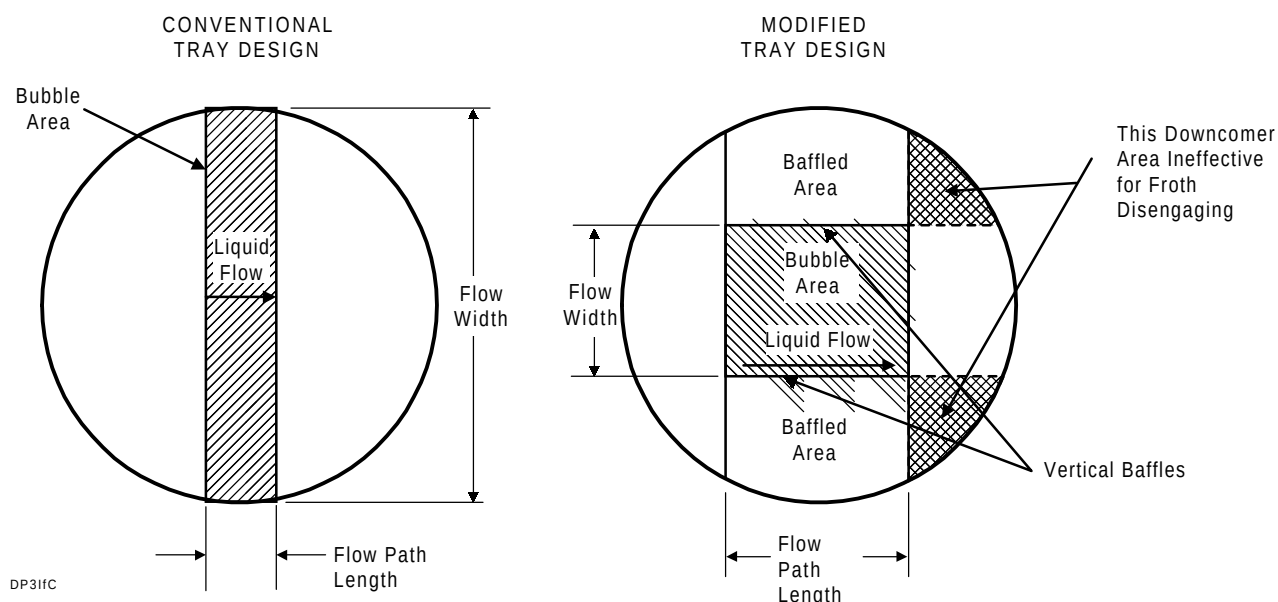
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DESIGN GUIDELINES FOR HEAVY HYDROCARBON STRIPPERS AND TRAYS CONTAINING LARGE BLANKED AREAS (Cont)

HOW CAN TRAYS BE IMPROVED?

The left side of **Figure** (below) shows a conventional tray design with an oversized downcomer whereas the right side shows a modified tray, which will operate with a higher efficiency. The modified tray contains a longer flow path length and a shorter flow width; however, the bubble area is the same in both cases. Trays with a large amount of blanking in the "conventional" design can also be modified in the same way by converting the blanked or "waste" area into baffled area.

**FIGURE C
IMPROVED STRIPPER TRAY DESIGN**



The improved efficiency is obtained in two ways: First the greater number of mixing pools across the tray due to the longer flow path length increases the ratio of tray efficiency to point efficiency [s **Eq. (11b1)** and **Figure 5**]. However, as **Figure** shows, no additional increase in efficiency is realized in most cases after 10 to 20 mixing pools. For this reason, no efficiency improvement due to a greater number of mixing pools should be expected when the flow path length is initially three ft (900 mm) or greater. Secondly, the shorter flow width increases the clear liquid height and the average froth height, thereby increasing the liquid and vapor residence time. In addition to these modifications, the outlet weir height may be increased to maximum of 4 in. (100 mm). This also serves to increase liquid and vapor residence time.

To estimate residence times, clear liquid height equations from **Section III-B** intended for double-pass **inboard trays** must be used. This applies even if the baffled tray is single-pass or double-pass outboard.

LIMITATIONS

As mentioned in the previous paragraph, less efficiency improvement is expected when the flow path length of the conventionally designed tray is 3 ft (900 mm) or greater. Thus, for revamps of existing trays with flow path lengths greater than 3 ft (900 mm), it is doubtful that increased efficiency due to greater residence time alone will justify the cost of the new trays. The designer may still consider the lower-cost alternative of increasing only the outlet weir height. In new tower designs, however, the cost differential between the conventional and modified tray designs is small, and the modified design is preferred without regard to flow path length.

As a limitation on flow width for a modified tray, the flow width should never be less than one-third of the tower diameter nor less than one-half the diameter if the tray in question is located directly under a conventionally designed tray. In addition, a lower limit on flow width and an upper limit on weir height may be imposed by downcomer limitations, as discussed in the next paragraph.

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DESIGN GUIDELINES FOR HEAVY HYDROCARBON STRIPPERS AND TRAYS CONTAINING LARGE BLANKED AREAS (Cont)

DOWNCOMER SIZING

In a modified tray, a significant portion of the original downcomer area is converted into baffled area. Therefore, downcomer hydraulic criteria (contained in other sections for the various types of trays) should be checked to insure proper operating conditions on the tray. The shorter weir length and decreased downcomer clearance area cause increases in both clear liquid height and head loss under the downcomer. Furthermore, the clear liquid height will increase still further if the weir height is increased. All of these effects cause an increase in downcomer filling, which may require an increase in tray spacing.

In addition, the decreased "effective" downcomer area will increase the downcomer velocity, which must also be checked. For these calculations, it should be assumed that the portion of the downcomer beyond the vertical baffles, which limit the flow width, is ineffective for downcomer disengaging. Thus, this area should not be included in the area used to calculate downcomer velocity. This area is shown (shaded) on the right hand side of **Figure C**.

TOWER FLEXIBILITY

As mentioned before, the clear liquid height is greater for the modified tray design case. However, since the baffles substantially reduce the bubble area, a smaller amount of hole area can be specified and still meet the minimum 3.5% hole area to bubble area ratio. Thus, the weeping rate should be evaluated to determine the modified tray's flexibility using the method given in **Section III-B**. This will permit optimizing the final tray design details.

DETERMINATION OF TRAY LOADINGS

Another factor to be considered in designing pipestill stripping trays is the change in loadings across the stripping section. The lowest vapor rate occurs at the bottom and increases up the tower as the liquid is stripped. Therefore, a designer should directionally specify lower hole areas and possibly lower widths for the bottom trays. The table below provides a guide for calculating tower loadings in standard four-tray and six-tray stripping sections in terms of the total moles of liquid stripped out. These values are applicable to sidestream strippers as well as bottom strippers.

FOUR-TRAY SECTIONS		SIX-TRAY SECTIONS	
TRAY NO.	PERCENT OF TOTAL STRIPOUT	TRAY NO.	PERCENT OF TOTAL STRIPOUT
1 (bottom)	30	1 (bottom)	20
2	15	2	15
3	15	3	10
4	40	4	10
% TOTAL	100	5	15
		6	30
		% TOTAL	100

For example, in a four-tray section, if the liquid load to the top of the stripper is 1000 moles/h and 100 moles/h of liquid are to be stripped out, 30 moles/h will be stripped on the bottom tray, 15 moles/h on the next tray, 15 moles/h on the next tray, and 40 moles/h on the top tray. The vapor load at the bottom tray will initially consist of only the stripping steam rate. This load increases on successive trays in increments equal to the amount of liquid stripped out.