to the minimum area to the minimum area required for vapour–liquid disengagement. The DCBUSF determines the approach of the downcomer froth height to the downcomer depth (= tray spacing + outlet weir height). Safety factors in the range of 1.5–2.0 are recommended.

Pressure Drop

The pressure drop across an operating tray should be specified if it affects the number of equilibrium stage requirements for the separation. This is often the case for vacuum applications. Stable operation can be obtained at a pressure drop of 1–3 in (2.5–7.6 cm) of liquid per tray for vacuum and 2–5 in (5.1–12.7 cm) for pressure operations.

Design Calculations

Tray Hydraulics

The hydraulic performance of a sieve tray for a given layout may be calculated using the methods presented in 'Distillation/Tray Columns: Performance'.

Tray Efficiency

Tray efficiency is a strong function of the physical properties of the vapour and liquid streams. It is also affected, to a lesser extent, by the flow rates and tray layout. In the latter case, only hole diameter, hole area and weir height have a small influence on the tray efficiency. The optimum design, which gives the maximum number of equilibrium stages in a column, is often obtained at minimum tray spacing and minimum number of flow paths that satisfy the hydraulic design criteria.

Conclusions

A well-designed tray should be economical while meeting all process design requirements. Economic considerations suggest that it is best to use the smallest column diameter and height that satisfy the process requirements within reasonable safety allowances. Process requirements include accommodation of the expected liquid and vapour flow ranges and the optimization of tray efficiency.

See also: **II/Distillation:** Packed Columns: Design and Performance; Theory of Distillation; Tray Columns: Performance.

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Tray Columns: Performance

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Introduction

As pointed out in the article entitled distillation tray columns: design, a sieve tray is designed with a number of objectives in mind. They include: (i) achieving high efficiency of contact between the liquid and the vapour so that the phases leaving a tray are as close to equilibrium conditions as possible; (ii) balancing the tray deck area provided for vapour/liquid contact with the downcomer area provided for disengagement of the two phases so that neither limits the capacity of the column to process large amounts of feed; and (iii) avoiding detrimental operating

conditions in the column such as *weeping*, *flooding* or high vapour *entrainment*.

Numerous geometrical factors have to be selected by the designer such as: (i) column diameter; (ii) tray spacing; (iii) top and bottom downcomer area; (iv) number of flow passes; (v) hole diameter and density; (vi) tray thickness; and (vii) weir design. This is a highly empirical process which depends on empirical design equations that describe the tray hydraulics and rule-of-thumb guidelines that have evolved over several decades of operating experience. Thus, the design of sieve tray columns has remained an art, although commercial process simulation software packages such as ASPEN, PRO II, HYSIM, etc., are trying to codify these procedures into their design packages. The conceptual steps in the design procedure together with the rule-of-thumb guidelines have been presented in the Tray Columns: Design article. Since frequent reference will be made to that article, we will henceforth refer to it simply as Part I.

In contrast, the performance analysis problem is relatively more scientific, in the sense that a series of well-defined steps leads to the estimation of the Murphree tray efficiency, the *column efficiency* and the *actual number of trays*. The overall column efficiency, E_o, is defined as:

$$E_o = \frac{N_{\text{equilibrium}}}{N_{\text{actual}}} \qquad [1]$$

where $N_{\text{equilibrium}}$ is obtained from stagewise equilibrium design calculations. Performance evaluation boils down to estimating E_o so that the actual number of trays, N_{actual}, can be determined. The overall column efficiency, E_o, is related to the Murphree tray efficiency, E_{MV}, through the Lewis relationship (assuming constant slopes of equilibrium and operating lines), given by:

$$E_o = \frac{\ln\left[1 + E_{MV}(\lambda - 1)\right]}{\ln\lambda}$$

where $\lambda = mG/L$ is the separation factor, m is the slope of the equilibrium line, and (G, L) are the vapour and liquid flow rates in kmol s^{-1}. Thus the Murphree tray efficiency, E_{MV}, must be estimated in order to determine the column efficiency. The urphree tray efficiency is defined to provide a measure of departure from the assumption of *ideal equilibrium tray* that is used to determine the number of *ideal stages* required to achieve a given separation. It is defined as:

$$E_{MV} = \frac{y_n - y_{n-1}}{y_n^* - y_{n-1}} \qquad [2]$$

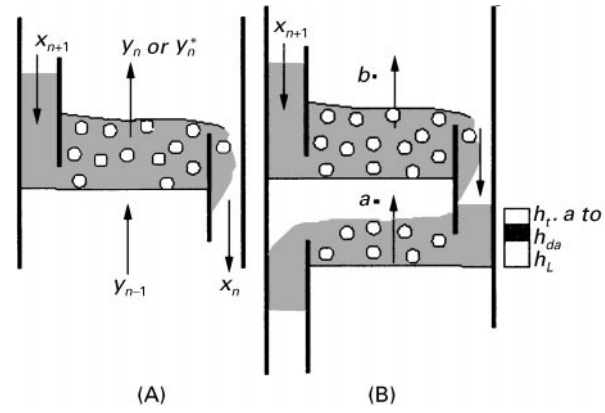

Figure 1 (A) Murphree tray efficiency. (B) Head in the downcomer.

Figure 1A illustrates various compositions. y_n is the actual composition of the vapour stream leaving tray n, while y_n^* is the composition that is in equilibrium with the exit liquid stream.

These two compositions would be the same, if and only if the condition of *ideal equilibrium tray* is satisfied. Since it is never satisfied in practice, it is important to be able to predict the tray efficiency. In fact, the compositions are not even uniform across the tray deck. Hence the above definition is applied at a local point on the tray and the point efficiency is integrated with the *variations in flow conditions* to predict a tray efficiency. The relationship between the inputs and the sequence of calculations is shown in **Figure 2**.

In Figure 2 the *point efficiency* is a function of local flow conditions such as local mass transfer coefficients in the liquid and vapour phases. The *dry Murphree* tray efficiency incorporates the effects of liquid and vapour distribution on the *point efficiency*, while the *wet Murphree* tray efficiency incorporates the additional effects of entrainment and weeping.

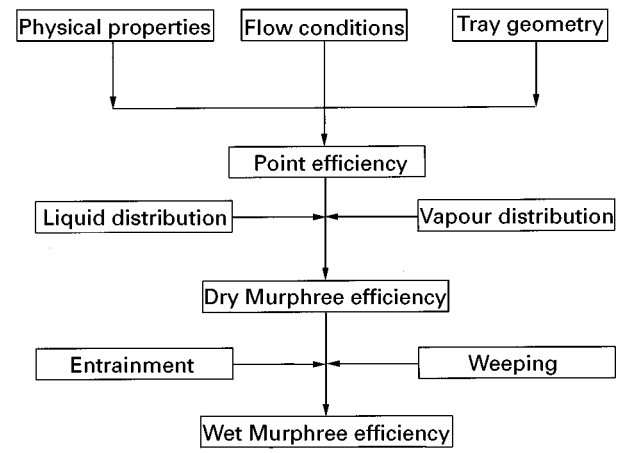

Figure 2 Steps in performance evaluation.

The tray efficiency, E_{MV}, clearly depends on: (i) the geometrical design parameters chosen as outlined in Part I; (ii) the physical properties of the system such as density, viscosity, surface tension, etc.; and (iii) the operating conditions like the vapour/liquid flow rates.

Having selected the design parameters identified in Part I, the objective of the performance analysis step is to predict: (i) the tray hydraulics (including the pressure drop, the flow regime, the froth density, the entrainment and weeping factors); (ii) the point efficiency; (iii) the Murphree tray efficiency; and (iv) the column efficiency. In the initial stages of designing a *new tray column*, there is feedback between the design and performance analysis steps to arrive at a set of optimal design parameters, as outlined in the flow chart (Figure 5 of part I). But the performance analysis steps to be outlined in this part, are also useful in analysing the performance of an *existing tray column*, although the opportunity to pick optimal design conditions is not present as one is forced to deal with an existing tray.

An excellent summary of the equations used to study the performance of a sieve tray column can be found in Zuiderweg (1982), Lockett (1986) and Kister (1992) (see Further Reading). The detailed steps involved in the performance analysis include: (i) the pressure drop prediction; (ii) froth height and density calculations; (iii) point efficiency prediction; and (iv) tray efficiency prediction. The inputs required are: (i) tray geometry; (ii) physical properties; and (iii) flow conditions.

Steps in Performance Analysis

Pressure Drop Calculation in Vapour Phase

The pressure drop in the vapour phase across a sieve tray is modelled as (Zuiderweg, 1982):

$$\Delta P = \Delta P_{dry} + \rho_l g h_l \qquad [3]$$

where the dry pressure drop is given by:

$$\Delta P_{dry} = \frac{1}{2}\rho_G\left(\frac{u_{g,h}}{C_D}\right)^2 \qquad [4]$$

Here, g is the acceleration due to gravity, h_l is the liquid height or hold-up in metres, $u_{g,h}$ is the vapour velocity in the hole in metres per second, C_D is the drag coefficient, and (ρ_G, ρ_l) are the densities of the vapour and liquid, respectively. The second term in eqn [3] represents the static head due to the liquid hold-up on the tray. Hence the liquid height, h_l must be predicted from correlations that depend on the weir geometry. One such equation that predicts the liquid height is given by:

$$h_l = 0.6H_w^{0.5}p^{0.25}(FP/b)^{0.25} \quad 25\text{ mm} < H_w < 100\text{ mm} \qquad [5]$$

Here, H_w is the weir height in metres, p is the pitch of the holes in the sieve plate in metres, $FP = (u_l/u_g)\sqrt{(\rho_l/\rho_g)}$ is the flow parameter, b is the weir length per unit bubbling area in metres^{-1}.

The discharge coefficient, C_D in eqn [4] is a function of the flow conditions near a hole. It is in fact dependent on the liquid present on the tray. It is correlated by:

$$C_D = 0.7\left[1 - 0.14\left(\frac{gh_l\rho_l}{u_{g,h}^2\rho_g}\right)^{2/3}\right] \qquad [6]$$

All of the quantities appearing on the right-hand side have been defined previously.

Pressure Drop Calculation in the Liquid Phase

The liquid is transported down through the *downcomer*. The capacity of the downcomer should be sufficient to handle the liquid load without becoming the rate limiting factor, i.e. without the liquid backing up the downcomer to a significant extent. Figure 1B shows the pressure differential components making up the total head differential on the liquid side as the liquid backs up the downcomer to a height of h_{dc}. The extent of liquid back-up can be estimated from:

$$h_{dc} = h_t + h_{da} + h_L \qquad [7]$$

where h_t is the pressure difference between points a and b in the vapour phase that is necessary to keep the vapour flowing upwards, h_L refers to the effective clear liquid height on the tray deck that must be overcome by the liquid in the downcomer, and h_{da} refers to the pressure loss due to the liquid flow under the downcomer apron. Note that h_t is necessary to keep the upward flow of vapour, but acts as a pressure differential that works against the natural liquid flow in the downcomer. If this pressure differential is large, the liquid will back up more in the downcomer. This points out the coupling between the pressure loss in the vapour phase through the tray deck area and the liquid flow in the downcomer. An optimal design must balance these two factors carefully. h_L and h_t can be estimated from the correlations provided in the previous section. h_{da} can be estimated from:

$$h_{da} = 165.2U_{da}^2$$

where h_{da} is in millimetres of liquid and U_{da} is the velocity under the downcomer apron in metres per second.

Froth Height and Density Calculation

The froth density (or the two-phase density) has been measured using gamma ray techniques. The average liquid volume fraction on a sieve tray, defined as $\bar{\varepsilon}_l = h_l/h_b$, is correlated by:

$$\frac{1}{\bar{\varepsilon}_l} - 1 = c_1\left[\frac{u_g}{(gh_l)^{0.5}}\left(\frac{\rho_G}{\rho_L}\right)^{0.5}\right]^n \qquad [8]$$

Here, h_b is the froth or bed height in metres and u_g is the vapour velocity on bubbling area in metres per second. The constants c_1 and n depend on the type of flow regimes. In the *spray regime*, they take on the values of $c_1 = 265$ and $n = 1.7$, while in the *mixed/emulsion regime*, they are 40 and 0.8. This requires one to estimate the flow regime to be expected under a given set of operating conditions. In Figure 3 of Part I, we identified the limits of operation to lie between the weeping and flooding conditions as the vapour rate is increased. Even within this permissible range of operation, the flow condition has been observed to change from *spray* to *froth* to *emulsion* to *bubble flow* regimes. The transition into the spray regime is given by the capacity factor defined as:

$$CF = u_g\left(\frac{\rho_g}{\rho_l}\right)^{0.5} = 0.85\,\frac{g^{0.5}h_l^{1.5}\cdot F}{d_h}$$

Here, CF is the capacity factor defined as $u_g\sqrt{(\rho_G/\rho_L)}$ in metres per second, u_g is the vapour velocity in the bubbling area in metres per second, F is the fractional hole area per unit bubbling area and d_h is the hole diameter in metres. The transition from the spray/froth to emulsion/bubble flow regime is controlled by the ratio of horizontal liquid momentum to vertical vapour momentum and is given by:

$$\frac{u_l}{u_g}\left(\frac{\rho_l}{\rho_G}\right)^{0.5} = \frac{FP}{b\cdot h_l} > 3.0$$

where u_l is the horizontal liquid velocity, u_g is the vapour velocity on bubbling area in metres per second, and FP is the flow parameter defined in eqn [5], b is the weir length per unit bubbling area in metres^{-1}, h_l is the liquid height or hold-up in metres.

Point Efficiency Calculation

There are many empirical correlations for predicting the mass transfer efficiencies on sieve trays. The most recent one is that proposed by Chen and Chuang (1993). It is based on data from industrial sized columns of Fractionation Research Inc. The point efficiency is related to the overall number of transfer units by:

$$E_{OG} = 1 - e^{-N_{OG}} \qquad [9]$$

Chen and Chuang present the following correlation for N_{OG} using data free of weeping and entrainment. But the data set spans both the froth and spray regimes:

$$N_{OG} = \frac{11\,\dfrac{1}{\mu^{0.1}\varphi^{0.14}}\left[\dfrac{\rho_L F_s^2}{\sigma^2}\right]^{1/3}(D_G t_G)^{0.5}}{\lambda\,\dfrac{11}{14}\left(\dfrac{D_G\rho_G}{D_L\rho_L}\right)^{0.5}\left(\dfrac{M_G L}{M_L G}\right)+1} \qquad [10]$$

Here $\lambda = mG/L$ is the separation factor, $F_s = u_s\sqrt{\rho_G}$ is the superficial F-factor in kg$^{0.5}$/m$^{0.5}$s, $t_G = h_f/u_s$ is the vapour-phase contact time in seconds, and h_f is the froth height in metres. Note that this correlation combines the *geometrical parameters* such as φ, the fractional perforated area, A_h the bubbling area, the *system properties* such as densities (ρ_L, ρ_G), diffusivities (D_L, D_G) viscosity (μ), the interfacial tension (σ) in newtons per metre, the molecular weights (M_L, M_G), and *operating conditions* such as (L, G), flow rates. This correlation appears to predict the point efficiencies to within 5% of experimental data over a wide range of pressures.

Murphree Tray Efficiency Calculation

The point efficiency model presented above is based on a detailed examination of mass transfer at the vapour/liquid interface. The *ideal equilibrium tray assumption* used in the McCabe–Thiele method asserts that the flow condition on a tray is homogeneous everywhere. If that were true, the point efficiency would be the same everywhere on the tray. But there is strong evidence that the flow is not homogeneous, the degree of inhomogeneity being larger in large diameter columns. Several researchers have tried to measure the velocity profiles across a sieve tray and increasingly computational fluid dynamics is being used as a tool to predict such flow fields. (See for example Solari and Bell (1986) and Mehta *et al.* (1998)). This information on flow profile must be integrated with the point efficiency calculations in order to predict a Murphree tray efficiency. One such method is given below as an illustration. This model considers only the effect of *longitudinal mixing*. A measure of the effective diffusivity, D_E is needed in this model. Models of other flow configuration are discussed in Lockett (1986):

$$\frac{E_{MV}}{E_{OG}} = \frac{1 - e^{-(\eta + Pe)}}{(\eta + Pe)\{1 + [(\eta + Pe)/\eta]\}} + \frac{e^{-\eta} - 1}{\eta\{1 + [\eta/(\eta + Pe)]\}} \qquad [11]$$

where:

$$\eta = \frac{Pe}{2}\left[\left(1 + \frac{4\lambda E_{OG}}{Pe}\right)^{1/2} - 1\right]$$

and Pe is the Peclet number, defined as $Pe = Z_l^2/D_E t_l$. Here Z_l is the length of liquid travel, or the distance between the two weirs and t_l is the liquid residence time. The effective diffusivity is given by:

$$\sqrt{D_E} = 0.0124 + 0.017u_G + 0.0025L + 0.0150W \quad [12]$$

where D_E is in square feet per second, u_G is superficial gas velocity, expressed as cubic feet per second divided by the active bubbling area in square feet. As the Peclet number becomes large, this model predicts efficiency enhancement much large than unity. In large diameter columns (large Z_l) the *Peclet* number can tend to take a large value which would suggest significant efficiency enhancements. But it should be remembered that the above model considers only the *longitudinal mixing* process. In large diameter columns, the liquid flow structure can be much more complicated as documented by Solari and Bell (1986). Hence, predicted values of E_{MV}/E_{OG} greater than 1.2 by the *longitudinal mixing* model should be viewed with caution, as they may not be realized in the field.

Effect of Entrainment on Murphree Tray Efficiency

The effect of entrainment on the Murphree tray efficiency is estimated from:

$$E_{MV,entrain} = E_{MV}\left[\frac{1}{1 + E_{MV}\psi/(1-\psi)}\right] \quad [13]$$

where:

$$\psi = \frac{e}{L+e} = \frac{absolute\ entrainment}{total\ liquid\ flow\ rate}$$

where e is the entrained liquid in moles per hour. Zuiderweg presents the following empirical equation to predict the liquid entrainment in the spray regime:

$$\psi = 1.0 \times 10^{-8}\left(\frac{h_b}{H_s}\right)^3\left(\frac{u_{g,h}}{u_l}\right) \quad for\ 0.3 < \frac{h_b}{H_s} < 0.9$$

Here H_s is the tray spacing in metres, h_b is the bed height as defined in eqn [8], $u_{g,h}$ is the vapour velocity in the hole in metres per second and u_l is the horizontal liquid velocity.

Weeping Point Determination

When the vapour velocity is too small, the liquid on a tray deck can flow down through the holes on the sieve plate, instead of the downcomer, which is the preferred path for the liquid. If weeping is significant, then it results in mixing of liquid streams between two neighbouring trays, thus degrading the performance of the column. The need to avoid weeping places a limit on the minimum vapour velocity. Zuiderweg presents the following correlations to predict the minimum operating limit.

Mixed/free bubbling regime

$$CF_w = F\sqrt{gh_l}\left[1 - 0.15\frac{FP}{bh_l}\right]$$

Emulsion flow regime

$$CF_w = 0.45F\sqrt{gh_l}$$

where $CF_w = u_g\sqrt{\rho_G/(\rho_L - \rho_G)}$ is the capacity factor at the weep point in metres per second, and F is the fractional hole area per unit bubbling area. Correlations to estimate the type of flow regime are given by Zuiderweg. Note that weeping will seldom occur in the spray regime as vapour velocities are sufficiently large under design conditions. The effect of weeping on the tray efficiency calculation has been studied by Kageyama (1969).

Extensions to Multicomponent Systems

The methods outlined above have been developed largely using experimental data for binary, two-phase systems. The question of whether they can be applied to multicomponent systems can be examined as follows. Tray hydraulics factors such as pressure drops, flow regimes, froth densities, etc., depend only on the fluid mechanics of the two-phase mixture on sieve trays; hence one can expect the correlations to be useful for multicomponent mixtures as long as mixture properties for densities, viscosities, interfacial tensions, etc., are used. On the other hand, the point efficiency (and hence the Murphree tray efficiency) depends on the mass transfer resistance of each component species in each phase. Since the diffusivities and the equilibrium ratios (or the slope of the equilibrium curve, m) could vary for each species, the point efficiency will be different for each species. The correlation given in eqn [10] is based on binary mass transfer data.

In the pseudo binary method of calculation (see Kister, 1992) two components are identified as the *light key* and *heavy key* components and the

Murphree tray efficiency is determined for such a binary pair. One then has the option of either using the efficiency so calculated for all of the remaining components or repeating the procedure for all possible binary pairs. Such detailed estimates of component efficiencies are then used as inputs to advanced process simulators such as ASPEN.

Issues Relating to Scale-up of Efficiency Data

Since the point efficiency data and correlations (like eqn [10] are (or should be) based on local conditions, they should, in principle, remain valid on all scales. They are then integrated with flow conditions to predict the overall tray efficiency. Correlations such as eqn [11], which provide this function of integrating the point efficiency to provide tray efficiency, do not remain valid at all scales. It has been well documented that the liquid flow patterns change quite dramatically depending on the diameter of the column and the location of the weirs near the downcomer. In future one can expect *computational fluid dynamics* to provide detailed flow information using models that remain scale invariant over a wide range of diameters.

Concluding Remarks

A series of correlations taken from the literature are presented. They permit the evaluation of the performance of a sieve tray, once a set of design parameters has been chosen as outlined in Part I. At the design stage of a new sieve tray column, one can embed this design and performance analysis steps into an optimization procedure, in such a way that the design parameters may be altered until a specified objective function is satisfied. The objective function could be a cost function that includes the capital cost of the equipment (which determines the column diameter, tray spacing, etc.) and operating costs (which determine the reflux and reboil rates and the number of ideal stages).

See also: **II/Distillation:** Historical Development; Instrumentation and Control Systems; Theory of Distillation; Tray Columns: Design; Packed Columns: Design and Performance; Vapour-Liquid Equilibrium: Correlation and Prediction; Vapour-Liquid Equilibrium: Theory.

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Vapour–Liquid Equilibrium: Correlation and Prediction

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Introduction

Distillation is a process used to separate liquid mixtures into two or more streams, each of which has a composition that is different from that of the original mixture. The process involves both the vaporization of the original liquid in order to generate the vapours and the subsequent condensation of the vapours to form the desired liquid products. It is evident that vapour–liquid equilibria (VLE) are essential to this separation process. Typical temperature–composition (T–x–y) diagrams, pressure–composition (P–x–y) diagrams, and vapour–liquid composition (x–y) diagrams for completely miscible binary systems are depicted in **Figure 1**.