1

Some general considerations

1.1 Introduction

Distillation has a long history. Reputedly it was the Chinese who discovered it during the middle of the Chou dynasty. Thereafter, the production of distilled liquors, the so-called liquids of the gods, followed the progress of civilisation. First India, then Arabia, the secret reached Britain before AD 500 as the production of mead. Surprisingly, it took a further millennium before whisky was first distilled in Scotland about AD 1500. The full history of distillation has been meticulously chronicled by Forbes (1948) and that specific to North America by Carr (1972).

Although alcoholic beverage production retains its importance for many, distillation plays a far greater role in human affairs today, for it is now the dominant separation process used in the petroleum and chemical industries. It has achieved this dominance, and seems likely to retain it, despite its apparently wasteful use of energy. Alternatives to distillation, such as solvent extraction, adsorption or membranes, can be more energy efficient, but they often have more than offsetting higher investment costs. As a result, distillation retains its advantage, particularly in large-scale applications. Because of its massive scale of operation, even small improvements in distillation can have significant impact, and Zuiderweg (1973) has estimated that two billion dollars in column investment costs alone were saved between 1950 and 1970 by research and development.

Turning now to distillation trays, the theme of this book, there is some evidence that a rudimentary form of sieve tray was employed by the Greeks in about the second century AD. However, it was in response to a competition sponsored by Napolean Bonaparte that continuous distillation using bubble cap trays was discovered by Cellier-Blumenthal in 1813. Sieve trays, as we now know them, were apparently first employed in the Coffey still in 1830. Fair (1983) has given a comprehensive review of the historical development of column internals.

The two dominant classes of column internals used today are trays and packing. Packing has a pressure drop which is about one-fifth that of trays. Consequently, it is often the preferred choice in cases where pressure drop is an overriding consideration such as vacuum distillation or where vapour recompression (heat pumping) is used. Sometimes a lower temperature heating medium can be employed in the reboiler using packing, but this is not always significant. An example of such a case is where quench water or steam is in surplus and at an appropriate temperature. Recent developments have made packed column scale-up less uncertain, but the more positive directed flow of each phase in a trayed column (at the expense of pressure drop) makes hydraulic and mass transfer behaviour more predictable for trays than for packing. Factors such as those shown in Table 1.1, and elsewhere (Thibodeaux & Murrill 1966, Fair 1970, Billet, Conrad & Grubb 1969), have to be considered for each application. It has been estimated that currently about 90% of installed distillation columns contain trays (Krummrich 1984).

1.2 Tray types

Factors which influence the selection of tray type include capacity, efficiency, turndown, pressure drop, fouling resistance, cost and, not least, tradition. Figs. 1.1 and 1.2 show simple representations of most of the tray types in common use classified by deck design and by flow path arrangement. A very large number of tray types are possible by combining different decks and flow path arrangements. Table 1.2 summarises points to

Table 1.1. *Trays or packing – some factors to consider*

	Trays	Random packing	Structured packing
Effect of scale-up on HETP[a]	Predictable	Difficult to predict	Predictable
Pressure drop	High	Low	Low
Established design techniques	Yes	Only for capacity, not for HETP	
Cost	Low	Low–medium	High
Suitability for fouling service	Yes	No	No
Feed point flexibility	Easy	Difficult	Difficult

[a] HETP – height of an equivalent theoretical plate

consider when selecting between the various options and indicates sources of more detailed information on each device. The tray patent literature illustrates the wide variety of tray types which have been proposed and it has been summarised by Jamal (1981). The great majority of trays currently being installed are either sieve or valve trays and this is reflected in the topics covered in subsequent chapters.

1.3 Classifying distillation systems

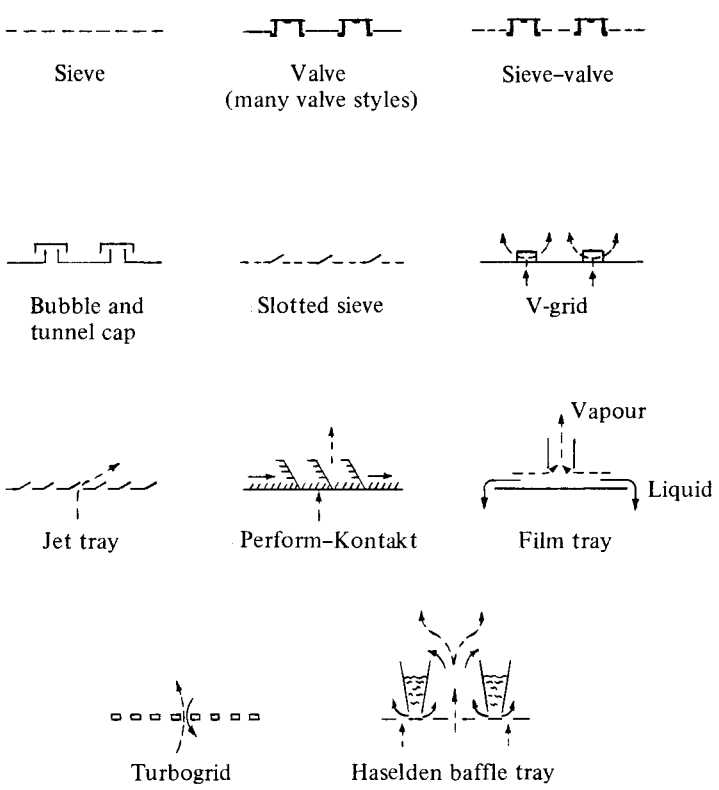
1.3.1 *Variation of physical properties with flow parameter*

The flow parameter (FP), defined by eqn. (1.1), is a useful dimensionless group which is frequently used in tray hydraulics correlations:

$$FP = \frac{M_L}{M_G} \cdot \left(\frac{\rho_G}{\rho_L}\right)^{0.5} \tag{1.1}$$

Except for easy separations of high relative volatility, the reflux ratio in distillation tends to be large such that M_G and M_L are not very different.

Fig. 1.1. Some styles of deck design in trayed columns.



Consequently, it is often a reasonable approximation to assume that

$$FP \approx \left(\frac{\rho_G}{\rho_L}\right)^{0.5} \tag{1.2}$$

Porter & Jenkins (1979) pointed out that, for typical distillation systems, changes in physical properties can often be correlated against each other. Thus, physical properties can be correlated against (ρ_G/ρ_L) or against FP using eqn. (1.2). Their suggested correlation is shown in Fig. 1.3. The correlation holds approximately for any combination of temperature and pressure providing they correspond to saturation conditions. A further useful approximation is achieved by noting that distillation is often carried out between 50 and 150°C. These limits allow condensation of overhead vapour against cooling water and avoid thermal degradation of bottom

Fig. 1.2. Some flow-path arrangements used in trayed columns.

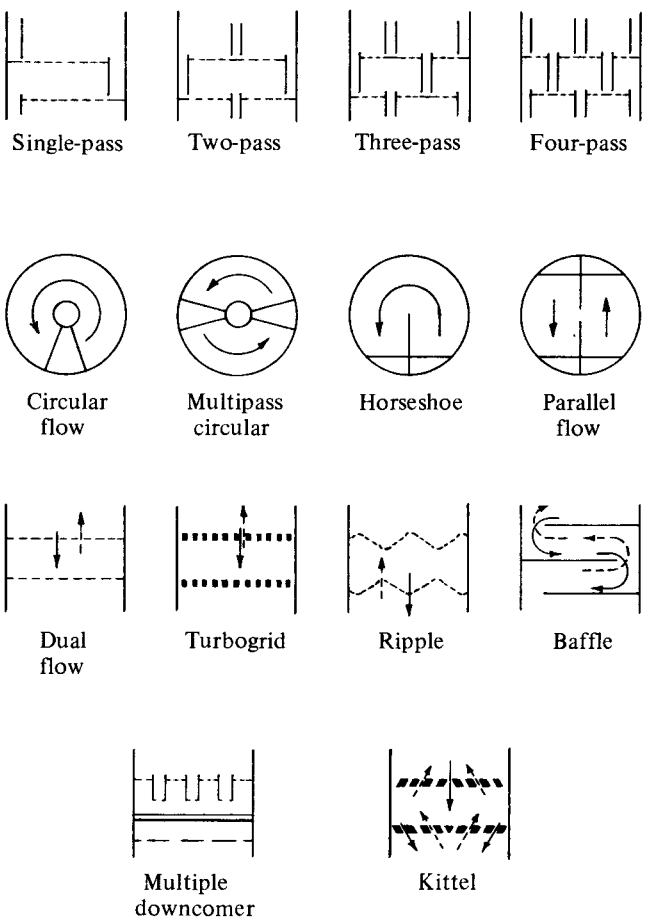
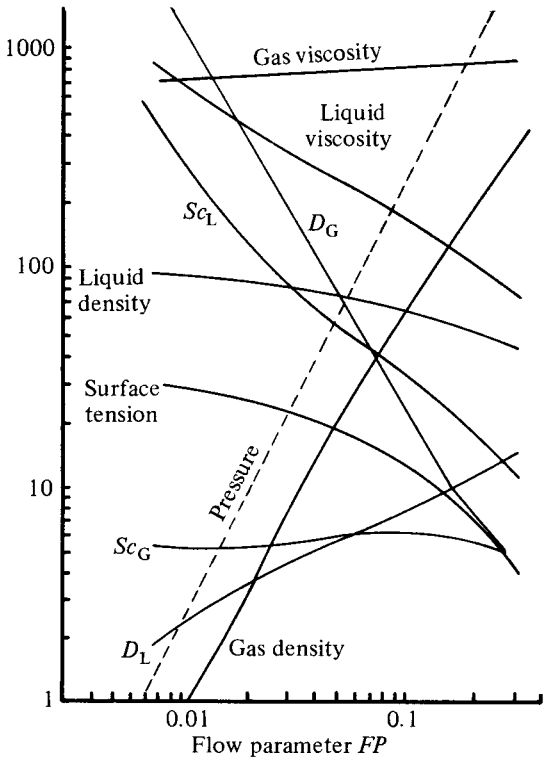


Table 1.2. *Some factors in choosing tray type*

Type	Comments	References
Sieve	Low cost, versatile, non-proprietary	
Valve	Lower turndown, 5–10% higher cost than sieve	Glitsch Inc. (1974) Koch Engineering Co. (1982) Nutter Engineering Co. (1976)
Sieve–valve	Lower turndown, higher efficiency than sieve	Billet *et al.* (1969)
Bubble cap	For extremely low turndown, high cost	Bolles (1956)
Slotted sieve	High efficiency, low pressure drop	Smith & Delnicki (1975)
V-grid	Less entrainment and weeping than sieve	Nutter (1971)
Jet	High liquid capacity – prone to liquid blow-off	Forgrieve (1960) Kirsten & Van Winkle (1970)
Perform–Kontakt	High vapour and liquid capacity	Raskop (1974)
Film	Low pressure drop	Leva (1972)
2–5 pass	For high liquid loads, prone to maldistribution, avoid odd number of passes	Bolles (1976*b*)
Multiple downcomer	For high liquid loads, high capacity	Union Carbide Corp. (1970) Delnicki & Wagner (1970)
Circular flow/horseshoe	Low liquid loads – Lewis's case 2	Bolles (1963)
Multipass circular	Higher liquid loads – Lewis's case 2	Ying *et al.* (1984)
Parallel flow	Maximum bubbling area – Lewis's case 2	Smith & Delnicki (1975)
Dual flow/turbogrid	Fouling services, poor turndown, low efficiency	Rylek & Standart (1964)
Ripple	Fouling services, handles solids	Hutchinson & Baddour (1956)
Baffle	Severe fouling or polymerisation service	Lemieux (1983)
Haselden's baffle tray	High capacity, high cost	Haselden & Witwit (1981)
Kittel	Low cost	Stanislas & Smith (1960)

product in the reboiler. Under this restriction, typical operating pressures can be included on Fig. 1.3 as shown by the broken line. Doig (1971) has also given a detailed discussion of the factors to consider in choosing the operating pressure for distillation. Clearly, inclusion of pressure on Fig. 1.3 is inappropriate when refrigeration is used, as for demethanisers or for air distillation, nor does it apply if overhead vapour recompression is used. Furthermore, it is too inaccurate to be used for design. Nevertheless it provides useful orientation. For example, it indicates that high-pressure distillation, corresponding to low-molecular-weight systems, is associated with low surface tension. This has important implications for column flooding and is discussed in Chapter 5.

Fig. 1.3. Typical variation of physical properties with flow parameter in distillation (Porter & Jenkins 1979). Ordinate multiplying factors ( ): for gas viscosity, N s m^{-2}, multiply ordinate by 10^{-8}; liquid viscosity, N s m^{-2} ($\times 10^{-6}$); gas self diffusion coefficient, D_G, m^2 s^{-1} ($\times 10^{-7}$); liquid self diffusion coefficient, D_L, m^2 s^{-1} ($\times 10^{-9}$); Schmidt numbers ($\times 10^{-1}$); pressure, bar ($\times 10^{-2}$); gas density, kg m^{-3} ($\times 10^{-1}$); liquid density, kg m^{-3} ($\times 10$); surface tension, N m^{-1} ($\times 10^{-3}$).



1.3.2 *Change of flow regime with flow parameter*

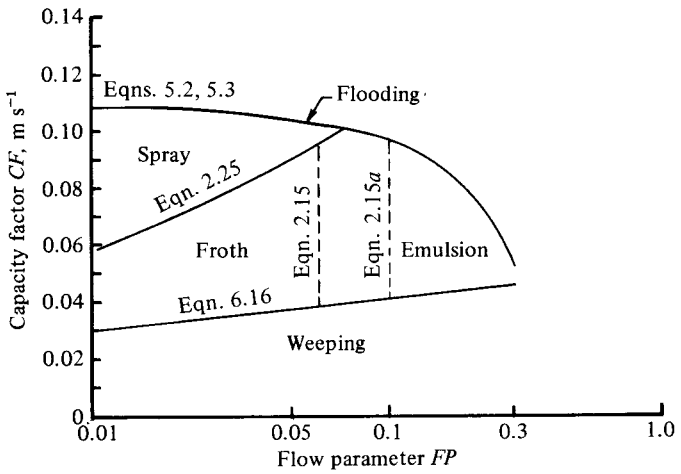
The flow regime diagram of Fig. 1.4 indicates how the flow regime on the tray typically changes with flow parameter. The precise location of the transition lines between regimes is open to dispute and can be established using the correlations outlined in subsequent chapters. The transitions also depend on the details of the tray design. However, Fig. 1.4 in combination with Fig. 1.3 does indicate that typically:

- vacuum distillation can result in spray regime operation;
- atmospheric distillation typically involves operation in the froth regime;
- high-pressure distillation is usually associated with the emulsion regime.

The characteristics of each flow regime are discussed in Chapter 2.

Again it must be emphasised that, useful though these generalised figures are, they only give a rough indication of trends and each case must be considered in detail. As an example, Figs. 1.3 and 1.4 imply that the separation of ethylbenzene-styrene, a typical vacuum distillation system, is associated with spray regime operation. In fact, this system is dominated by the need to minimise pressure drop so as to limit polymer formation in the reboiler. As a result, rather lower than normal superficial vapour velocities are used and typically the trays operate in the froth regime; see Fig. 1.5 (Lockett, Plaka & Ahmed 1984).

Fig. 1.4. Capacity factor vs. flow parameter showing flow regimes. Conditions as Table 2.1. *CF* based on bubbling area – see Chapter 5.



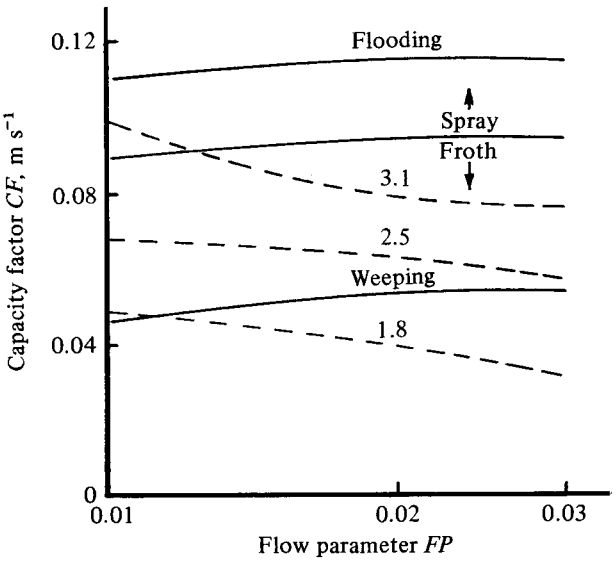
1.3.3 *Variation of column diameter and number of passes with loading*

Useful orientation can be obtained from the total flows chart developed by Porter & Jenkins (1979) shown in Fig. 1.6. It is based on unpublished charts prepared by the author. Fig. 1.6 was constructed using flooding data released by Fractionation Research, Inc. (Sakata & Yanagi 1979), and a constant downcomer liquid velocity of 0.19 m s^{-1}. Also included on Fig. 1.6 are lines corresponding to some typical pressures for distillation at total reflux plotted using Fig. 1.3. The chart quantifies common experience that high pressures are associated with multipass trays, whereas low-pressure distillation is usually carried out using single-pass trays. A more detailed discussion can be found in Section 5.1.3.

1.4 An outline design procedure

The basic steps involved in designing distillation trays have been well documented both for new columns and for retrays of existing columns (Billet 1979, Backhurst & Harker 1973, Bolles 1956, 1963, Chase 1967, Economopoulos 1978, Fair 1963, Frank 1977, Glitsch, Inc., 1974, Kister 1980, Koch & Kuzniar 1966, Koch Engineering Co. 1982, Neretnieks 1970, Nutter Engineering Co. 1976, Raper *et al.* 1977*b*, Sewell 1975, Stichlmair 1978). There is considerable creative skill and experience required to arrive

Fig. 1.5. Typical capacity factor vs. flow parameter variation for 7.0 m-diameter single-pass sieve tray in vacuum distillation (Lockett, Plaka & Ahmed 1984). ---- Tray pressure drop (mm Hg).

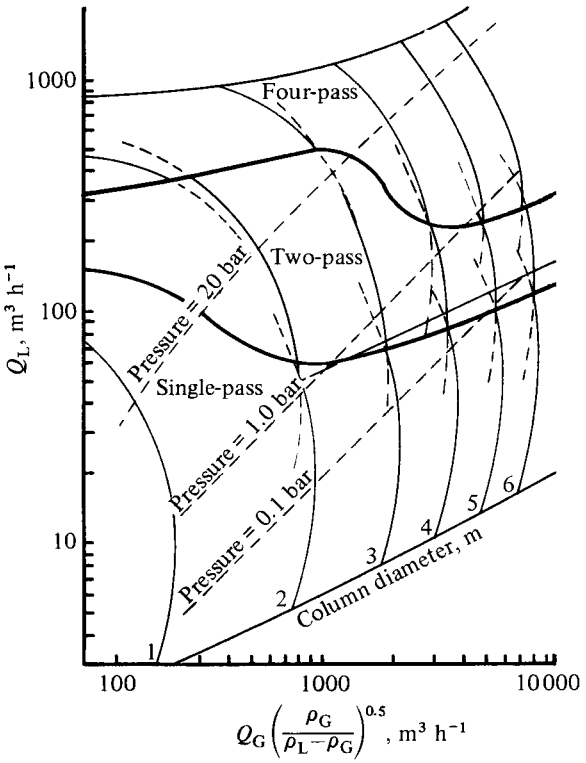


at a safe design but one which is also cost effective by incorporating a minimum of overdesign. The reason for this is that most of the variables involved interact in a complex way.

Set out below is an outline of a typical design procedure for sieve trays with some discussion of the tradeoffs involved. A similar procedure applies to valve trays. Reference is made to subsequent sections where more details are given.

(1) Fix the number of passes, tray spacing and hole diameter based on experience in similar applications. As a guide, use Fig. 1.6 to give the number of passes, and choose an initial tray spacing of 0.61 m. There are two schools of thought about hole diameter. In one, a hole diameter of 12.7 mm is nearly always used. A better approach is to use a hole diameter of 4.8–6.4 mm unless fouling or corrosion are likely to be excessive. Smaller holes give a higher vapour capacity and can allow increased turndown, although perforation costs are slightly higher.

Fig. 1.6. Total flows chart. $FF = 0.8$, $T_s = 0.6$ m. (Porter & Jenkins 1979.)



(2) Fix the turndown required (Chapter 6). A typical turndown for a sieve tray is 60–70% of full load flow rates. Sieve trays can be designed for far lower turndown, but at the expense of pressure drop and often of increased tray spacing. Since the provision of turndown is expensive, it is important to determine at the outset whether the turndown asked for is really required.

(3) Fix the exit weir height. A typical value is 50 mm, with 0–25 mm used in vacuum distillation and 100 mm in absorbers and strippers. Increasing the weir height increases the tray efficiency but at the expense of pressure drop.

(4) Determine the bubbling area and downcomer area from capacity correlations (Chapter 5).

(5) At turndown conditions, determine the fractional perforated tray area to ensure tray stability and minimise weeping at an acceptable level (Chapter 6 and Section 9.11). Since the weep point and weeping correlations involve the clear liquid height, which in turn depends on the fractional perforated area (Chapter 3), an iterative procedure is required.

(6) Calculate the following at full load conditions and take remedial action as appropriate:

   (*a*) Maximum liquid load over the weir. If the maximum weir load (Section 5.4) is exceeded, increase the number of passes.

   (*b*) Pressure drop (Chapter 4). If this is excessive, one remedy is to design at a lower percentage jet flood by increasing the bubbling area. Alternatively, the fractional perforated area can be increased at the expense of turndown. Also, the weir height can be reduced but with a reduction in efficiency.

   (*c*) Downcomer backup (Section 5.3). If this is excessive, remedies are to increase the tray spacing, reduce the dry tray pressure drop at the expense of turndown, or reduce the weir height.

   (*d*) Flow regime (Chapter 2). Some designers prefer to avoid the spray regime because of unpredictable performance, yet others deliberately choose it when possible to reduce column costs (Section 8.7). Consideration should be given to avoiding the emulsion flow regime because of downcomer capacity limitations and vapour entrainment (Section 5.3.2). This can be achieved by increasing the number of passes or using multidowncomer trays.

   (*e*) Entrainment (Section 5.2). Entrainment can be reduced by increasing the tray spacing. Other remedies are also available