The CONTACTOR™

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What Affects Efficiency in Trayed and Packed Columns

There are several definitions of efficiency in the context of a mass transfer or separation device, all having the ultimate goal of allowing the number of theoretical stages (NTS) or the number of theoretical plates (NTP) to be translated into the number of real trays (N), or the actual height of packing (H), needed to achieve a specified separation. The simplest is the overall efficiency (η for trays, HETP for packing) which allows for a direct translation between theoretical and actual:

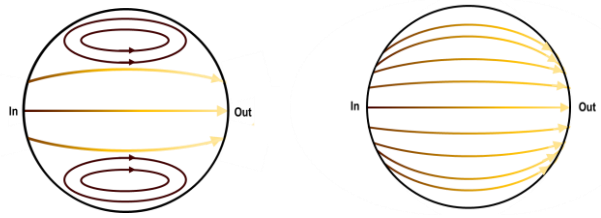
$$\eta = \frac{NTS}{N}; \quad HETP = \frac{H}{NTS}$$

For binary systems, NTS can be easily calculated graphically. For multicomponent systems the concept of key components is often used. These methods, ingenious at the time, were developed to allow distillation calculations to be done graphically, long before the advent of digital computers. Today, NTS is calculated digitally with great speed† while avoiding simplifying assumptions such as equimolar overflow. The central problem with using theoretical stages, however, is determining the efficiency or HETP, especially for a new system, or for a familiar system under unfamiliar operating conditions. Efficiencies are hard to calculate reliably (see Duss and Taylor‡,§ for very readable expositions). Perhaps the difficulty partly stems from our inability to quantify accurately the extent of back-mixing of liquid as it crosses a tray and back-mixing of liquid as it descends through a bed of packing.

Tray and Packing Hydraulics

Vapor flow is mostly vertical and, without significant vapor maldistribution gross back-mixing is limited. At first glance, the liquid flows in trayed and packed columns appear quite different; however, the difference is somewhat superficial. Liquid flows *horizontally* across a tray where it contacts the vapor before descending through downcomers. There is no *axial* dispersion during the flow from tray to tray (unless there is massive entrainment or weeping) but a variety of patterns is possible as the liquid flows across the tray. As shown in Figure 1, these range from plug flow to completely mixed, with varying recirculation between the extremes. Liquid mixing can greatly affect tray efficiency and a liquid in nearly plug flow can produce overall efficiencies well in excess of 100%. This is not possible if the liquid is completely mixed or shows recirculation as in Figure 1(a).

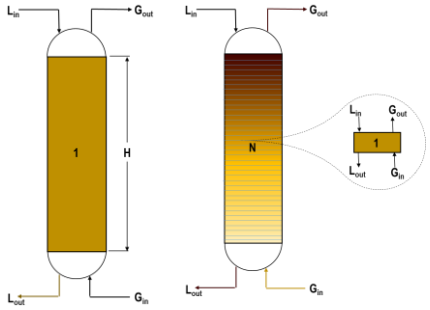
Liquid moves through a packed column in films flowing over the surface of the packing. The packing surfaces have a whole range of orientations from vertical to horizontal, and each orientation will correspond to a film of different thickness flowing with a different net vertical velocity component. Thus, packets of liquid will move with different vertical velocities and this necessarily results in back-mixing. If there is maldistribution of liquid (and therefore also of vapor), axial dispersion is exacerbated. One should certainly expect axial dispersion to depend on packing size—in fact, it probably scales directly with the packing size within any one family of packings.



(a) Recirculation with Short Weir (b) Flow Corrected with Push Valves

Figure 1 **Typical Flow Patterns on Crossflow Trays**

For both trays and packing, axial dispersion in the vapor is usually insignificant compared with the liquid. In trayed columns, liquid flows are not subject directly to *axial* dispersion. Instead they experience cross-flow dispersion with liquid then moving intact from tray to tray. However cross-flow dispersion parallels axial dispersion because both are in the direction of flow. In packed columns of both random and structured packing, the liquid flow experiences direct axial dispersion. But in an overall sense, dispersion in these very different setups has similar effect.



(a) Fully Back-mixed Packed Bed is a CSTR *(b) Overly Segmented Bed is in Plug Flow — No Back-mixing*

Figure 2 **Axial Dispersion in a Packed Bed**

Although useful in the interpretation of measured performance information as expressed by tray efficiencies and HETP values, the issue of phase dispersion is highly relevant in rate-based modelling. In such models, the actual separation is computed directly without considering separate efficiency calculations. On each tray, liquid is taken to be completely mixed.

† The rapidity with which the number of theoretical stages can be computed is no longer an overriding factor because mass transfer rate-based methods can now be done extremely rapidly too, and they completely circumvent the need to calculate or estimate efficiencies.

‡ Duss, M.; Taylor, R.; 2018, *A New Tray Efficiency Model: How Simple May It Be?*, Chemical Engineering Transactions, 69, 691-696 DOI: 10.3303/CET1869116

§ Duss, M.; Taylor, R.; *Predict Distillation Tray Efficiency*, www.aiche.org/cep, July, 2018

Packed columns are simulated by discretizing the total bed depth into a number of segments with each segment corresponding to completely back-mixed liquid. At the other extreme, an infinite number of segments in a tower would correspond to perfect plug flow of both phases. Figure 2 compares visually how these extremes might affect a separation. The truth is somewhere between these limits. ProTreat® segments packed towers according to general rules of thumb and internally chosen generalised heuristics to achieve best agreement with a library of performance data. There is rough equivalence between a packed segment and a real tray but *these devices have very different mass transfer characteristics* so they perform quite differently.

Tray and Packing Mass Transfer

Analysis of distillation and gas absorption can use either theoretical stages and efficiencies (or HETPs), or a mass transfer rate model using concentration differences and responding directly to mass transfer coefficients and interfacial areas. But ultimately, both approaches must have recourse to the *mass transfer characteristics* of the internals. The mass transfer characteristics of different *types* of internals are very different although random and structured packings are closer kin to each other than to trays. The mass transfer characteristics of any tower internal can be expressed in terms of (a) effective interfacial area, a, per unit volume, (b) gas-side mass transfer coefficient, k_G, and (c) liquid-side mass transfer coefficient, k_L. If there is chemical reaction involved, then the enhancement factor is a fourth parameter. Further details are unimportant here.

Trays

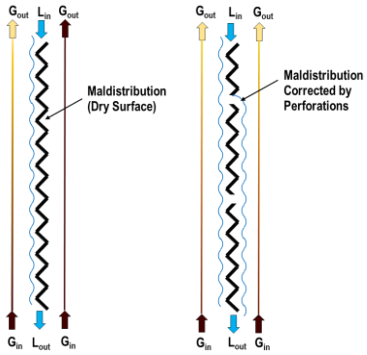
The effectiveness of mass transfer depends on interfacial area and how vigorously the phases contact each other. Vapor is injected forcefully through perforations in the tray deck and if the perforations are covered by valves, the vapor has an energetic horizontal velocity component. Typical vapor velocity through tray perforations is 3 m/s, generating interfacial area of typically 100 m^2/m^3 based on volume between the trays. Interfacial area on trays is greatly affected by vapor rate which at higher vapor rates generates a finer gas-liquid dispersion; area is hardly affected at all by the liquid velocity across the tray. Vapor density has a similar effect but again, liquid density is almost immaterial. Lower liquid viscosity tends to increase the interfacial area a little by allowing for a finer gas-liquid dispersion, perhaps by using more of the gas kinetic energy to shatter the liquid rather than just moving it around. However, liquid viscosity has a significant effect on the liquid-side mass transfer coefficient because low viscosity gives thinner mass-transfer films, lowering the liquid phase resistance to diffusion. Low viscosity fluids also have higher diffusion coefficients.

The liquid-side coefficient is unimportant in most distillation applications because distillation tends to be controlled by the gas-side resistance to mass transfer. However, it can be significant in the absorption of sparingly soluble gases because such processes can be controlled by liquid-phase diffusion. This has a rather counterintuitive consequence in gas treating, particularly CO_2 removal. Increasing the concentration of reactive component in the solvent (usually an amine) would be expected to increase the reaction rate of dissolved gas with the solvent component; however, it also increases the viscosity of the solvent which lowers the diffusion coefficient and thickens the diffusion film for mass transfer. The net result can be *reduced* absorption rate when intuition suggests the opposite.

One of the more useful geometric embellishments on trays is the use of push valves. These are valves that cover the tray perforations with a hood but with the side of the hood facing the oncoming liquid closed off and the downstream side open between hood and tray deck. Among other attributes, push valves boost the horizontal velocity of the liquid, preventing the retrograde liquid motion shown in Figure 1(a), i.e., they tend to reduce back-mixing. Another enhancement is to use directional vanes in the inlet downcomer's escape area, forcing liquid near the tower wall to flow more uniformly. These additional devices can be important in keeping the tray efficiency as high as possible; indeed, trays with plug flow liquid can have efficiencies in excess of 100% because liquid that has already been processed on the entry region of the tray is not having the already achieved separation destroyed by mixing it with as yet unprocessed liquid. Of course, a conservative design will assume well-mixed liquid on a tray.

Packing

Mass transfer in packed columns shows almost the opposite dependence on physical properties. Now a discontinuous liquid flows as a film over solid surfaces through a continuous gas, and mass transfer rates are greatly affected by packing size, packing geometry, and in the case of structured packing, by the surface treatment of the (usually metallic) packing. Packing size almost directly correlates with the effective interfacial area. Packing geometry is unique to each packing brand although the dry surface area is still the controlling factor. With structured packings, surface treatments such as embossing of the sheet metal improves the ability of the liquid to spread, but perforations are even more important. Communication is poor between adjacent sheets of packing if the metal sheets are imperforate. Figure 3 illustrates how perforations can open up communication and allow the evening out of liquid flows. Thus, flow over perforate sheets is much more uniform than over imperforate sheets, and lack of perforations only encourages the continuance of liquid (and vapor) maldistribution and uneven flows. Although the liquid is still agitated as it flows over both random and structured packings, it is much less so than the liquid on a tray because its movement is constrained by the thin nature of the film flow itself.



(a) Maldistribution on Packing sheet *(b) Maldistribution Corrected by Perforations*

Figure 3 **Perforations Even Flow**

Summary

Both structured and random packings exhibit strong dependence of effective interfacial area on liquid flow rate, but trays show only a weak dependence. And unlike packing which shows strong correlation of HETP with the design (type) and size of packing, crossflow trays show only a weak relationship between efficiency and tray geometry.

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