



How Hydraulics Affects Efficiency in Towers — Part 1: Trays

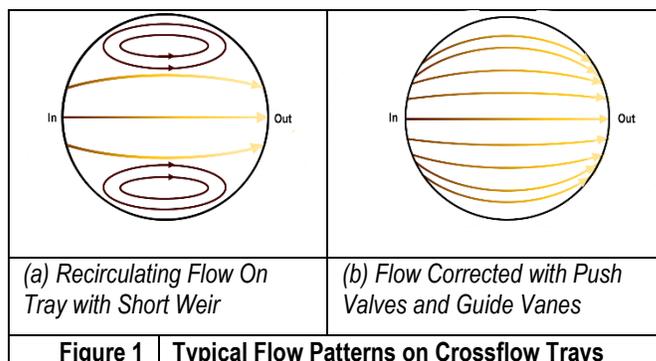
There are many different definitions of mass transfer efficiency but they are all intended to allow the number of theoretical stages (NTS) or the number of theoretical plates (NTP) to be translated into the number of real trays (N), or the actual height of packing (H), needed to achieve a certain separation. 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 calculated graphically either using a McCabe-Thiele x-y diagram or, for systems with large heat effects, using the Ponchon-Savarit (H-x-y) graphical method. For multicomponent systems the concept of key components is most 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 in part the difficulty 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, something that is related directly to hydraulics.

Tray and Packing Hydraulics

Vapour flow through trays and packed beds is predominantly in the vertical direction and, unless there is significant vapour maldistribution, there is limited opportunity for vapour back-mixing. The same cannot be said for liquid flows. At first glance, the liquid flows, in trayed versus packed columns appear quite different; however, the difference is somewhat superficial. Liquid flows horizontally across a tray where it contacts the vapour before descending through downcomers. There is no *axial* dispersion during the flow from tray to tray (unless there is entrainment or weeping) but there is a variety of patterns possible as the liquid flows across the tray. As shown in Figure 1, these range from plug flow to completely mixed, with varying recirculation patterns between the extremes. The state of liquid mixing affects the tray's efficiency. For example, with liquid traversing the tray in nearly plug flow it is quite possible to realize an overall tray efficiency 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 vapour), 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 in any one family of packings.

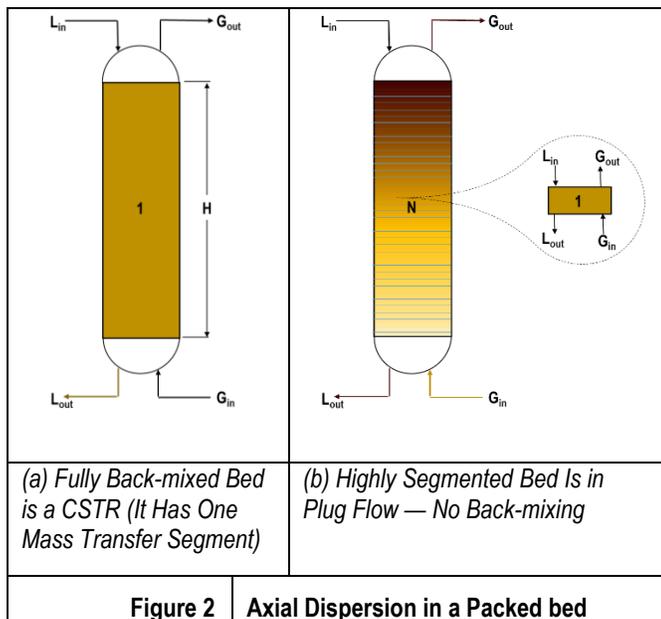
For both trays and packing, axial dispersion in the vapor phase is usually less significant than in the liquid. In trayed columns, liquid flows are not subject directly to *axial* dispersion. Instead they experience cross-flow dispersion, and with liquid then moving intact from tray to tray. However cross-flow dispersion parallels axial dispersion because both are in the direction of flow. In packed columns, the liquid flow experiences direct axial dispersion. But in an overall sense, the dispersion in these very different setups is similar in effect. The flow on trays are often approximated by plug flow vapour and back-mixed liquid.

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 mass transfer 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. 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.

[†] 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

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* because of different modes of phase contact, so they perform quite differently.



Tray Mass Transfer

Any component separation in distillation and gas absorption can be considered on the basis of theoretical stages and efficiencies (or HETPs), or as a mass transfer rate process driven by concentration differences and responding directly to mass transfer coefficients and interfacial areas. Both approaches ultimately must have recourse to the *mass transfer characteristics* of trays and packing, either to assess efficiencies or to perform central mass transfer rate calculations. The mass transfer characteristics of trays, random packings, and structured packings 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, E , is a fourth parameter — this parameter involves chemical reaction kinetic and reaction equilibrium constants and the liquid-side mass transfer coefficient. Further details are unimportant here.

The effectiveness of mass transfer depends on how vigorously the vapour and liquid phases are brought into contact with each other. In the case of trays, vapour is injected quite forcefully through perforations in the tray deck and if there are valves (covers) over the perforations, the vapour is given an energetic horizontal component to its velocity. A typical vapor velocity through tray perforations, thence into the liquid, is 3 m/s and this generates an interfacial area of typically 100 m² per cubic metre of total volume as seen between the trays and above the active tray deck area. Interfacial area on trays is greatly affected by vapour rate which imparts energy into the liquid and at higher vapour rates generates a finer gas-liquid dispersion, but area is hardly affected at all by the liquid velocity across the

tray. Vapour 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 results in thinner mass-transfer films and so lowers the liquid phase resistance to diffusion. Furthermore, low viscosity fluids have significantly 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 into reactive amines because such processes tend to be controlled by the resistance to mass transfer (diffusion) in the liquid-phase. This has a rather counterintuitive consequence in gas treating, particularly CO₂ 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 and the installation of guide vanes which are intended to keep the liquid moving in the right direction across the tray and discourage retrograde motion. 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 impart a boost to the horizontal velocity of the liquid and tend to prevent the retrograde liquid motion shown in Figure 1(a), i.e., they tend to reduce back-mixing caused by the establishment of recirculating liquid flow on the tray. Another enhancement is to use directional vanes in the escape area under the downcomer leading onto the tray to force 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 back mixing it with still to be processed liquid. The vapor leaving the tray also shows a composition profile in the crossflow direction so in reality the vapor and liquid have compositions that are changing across the tray. Of course, a conservative design will assume well-mixed liquid on a tray.

We will continue the discussion by looking at packing in more detail in the August issue of The Contactor.

To learn more about this and other aspects of gas treating, plan to attend one of our training seminars. For details visit www.oqtr.com/seminars.

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