

What Affects Efficiency in Trayed and Packed Columns¹

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There are many different definitions of efficiency as it pertains to a mass transfer or mass separation device but they all have 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 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. (It should be said that 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.) 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^{2,3} 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.

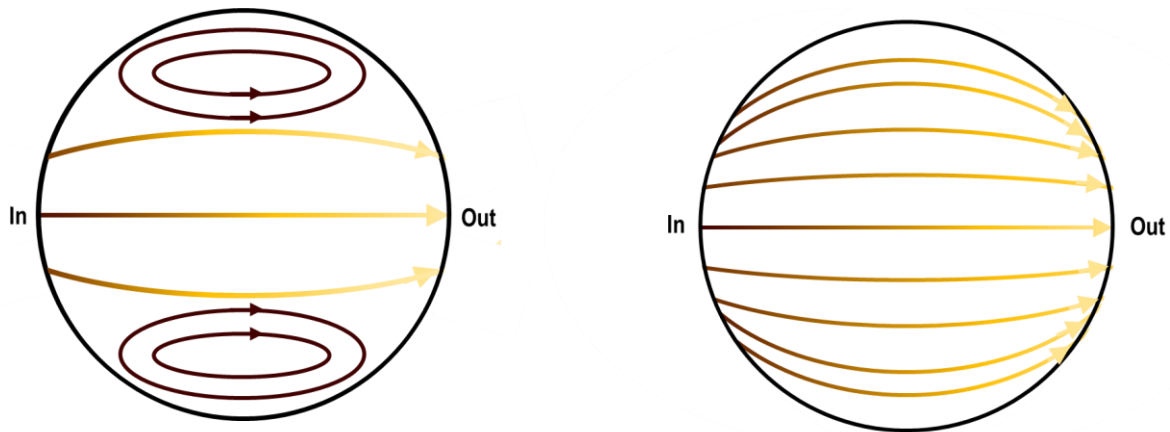
Tray and Packing Hydraulics

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 massive 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 can greatly affect 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).

¹ Published in Hydrocarbon Engineering, February, 2020

² 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

³ Duss, M.; Taylor, R.; *Predict Distillation Tray Efficiency*, www.iche.org/cep, July, 2018



(a) *Recirculating Flow On Tray with Short Weir*

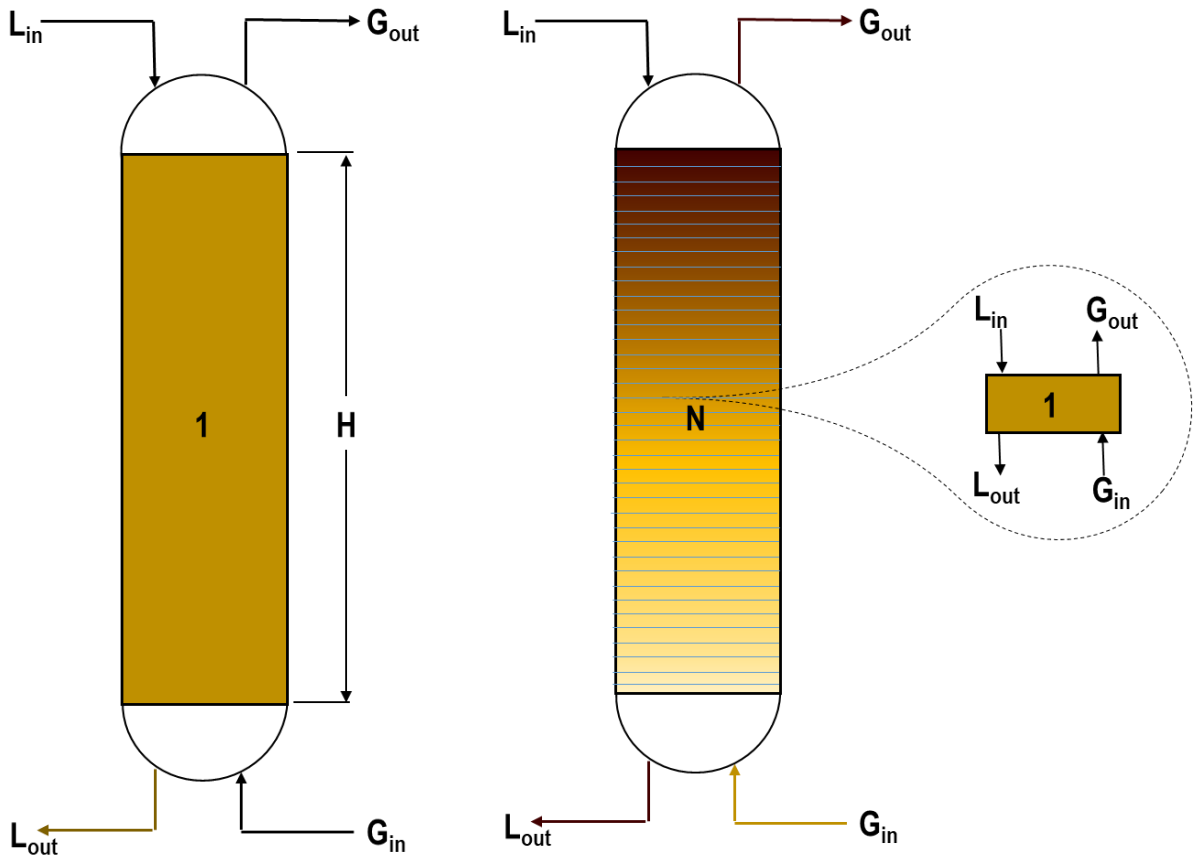
(b) *Flow Corrected with Push Valves and Guide Vanes*

Figure 1 Typical Flow Patterns on Crossflow Trays

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 vapour phase is usually insignificant compared with that in 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, with both random and structured packing, the liquid flow experiences direct axial dispersion. But in an overall sense, the dispersion in these very different setups is very similar in effect. One approach is plug flow vapour with completely 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. 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.



(a) A Fully Back-mixed Packed Bed (One Mass Transfer Segment) is a CSTR
 (b) Overly Discretized (Segmented) Bed Is in Plug Flow — No Back-mixing

Figure 2 Axial Dispersion in a Packed bed

Tray and Packing 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.

Trays

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

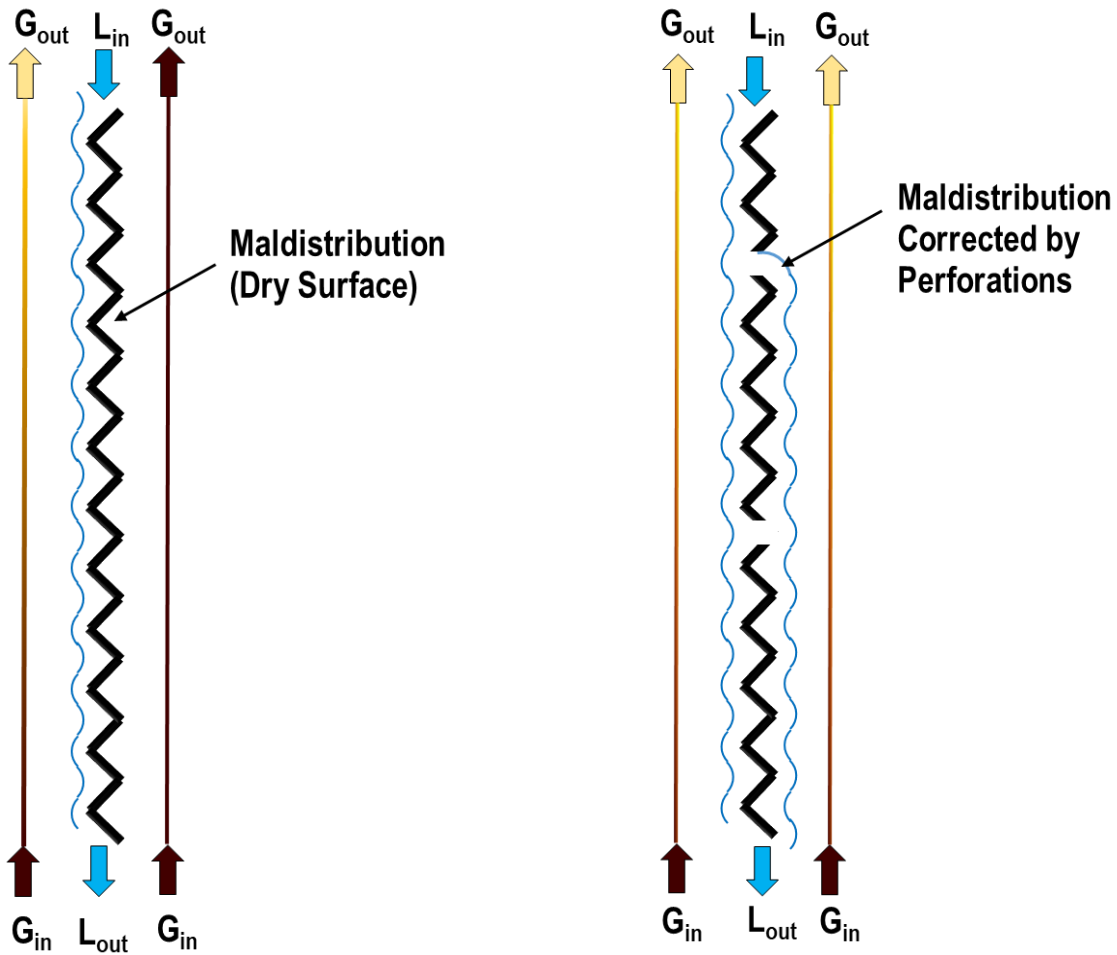
deck and if there are valves (covers) over the perforations, the vapour is given an energetic horizontal component to its velocity. A typical vapor velocity through tray perforations, thence into the liquid, is 3 m/s and this generates an interfacial area of typically 100 m² per cubic metre of total volume as seen between the trays and above the tray active 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 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. 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 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 vs. 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 vapour) 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 a sheet of Structured Packing (b) Packing Sheet Perforations Can Correct Maldistribution

Figure 3 Effect of Sheet Perforations on Liquid Flow over Structured Packing

Structured packings made from gauze rather than embossed sheet metal (both are perforated) excel at spreading liquids over much of the packing surface area, particularly when the liquid is flowing at low rates. However, they are certainly more costly than sheet metal and their use is hard to justify at high liquid flows. Gauze packings tend to be small crimp and find their greatest use in low production-rate distillation of fine chemicals (small diameter equipment) which again limits their use to the low flow rates needed to avoid column flooding from small crimp size.

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) of packing, trays show only weak relationship of efficiency to tray geometry. Exceptions are trays with push valves and guide vanes. Another exception is so-called high (hydraulic) performance trays such as ULTRAFRAC® and Shell ConSep* where there is no real crossflow, and point efficiencies are the best that can be achieved because phases are centrifugally separated with very short pointwise (no crossflow) contact. These trays, however, have the advantage of enormous processing capacity albeit at the cost of relatively lower efficiency.

Crossflow trays have higher efficiency but with lower gas and liquid handling capacity because the gas-liquid separation is gravity driven rather than through induced centrifugal force.

Efficiency can be improved by increasing the gas-liquid interfacial area either by using finer packing or by using smaller valves such as mini-valves and higher gas velocities through the tray openings. Low viscosity also favours finer gas dispersions in the liquid and thinner more-turbulent film flows on packing surfaces. The effect is more pronounced with packing than trays. Low viscosity also favours increased diffusion coefficients in the liquid and, for liquid-side controlled separation processes, this can provide substantial benefits.

There are basically three parameters that describe the mass transfer characteristics of a contacting device: interfacial area, k_L , and k_G . These characteristics are dependent on phase flow rates, physical properties such as density, and transport properties such as viscosity and diffusion coefficients. Mass transfer rate-based column models like ProTreat use fluid- and device-specific correlations for all the relevant properties that affect mass transfer rates and do their separation calculations without reference to efficiencies and HETP values. The calculations might take a little longer but significant benefits come from the engineer not being asked for hypothetical data. On the other hand, theoretical stage calculations execute very quickly but leave the engineer with the task of providing information that, at best, he can estimate only with limited reliability, or perhaps can only guess.