Axial Dispersion and Backmixing in Packed Columns in Gas-Liquid Operations

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In recent times there has been some interest in the effect of axial dispersion in the fluid streams upon the mass transfer performance of packed towers in gas-liquid operations. Conventionally, and as a matter of mathematical convenience, the dispersion is described using the dispersed plug-flow model. Its intensity is reported in terms of effective diffusivities (dispersion coefficients). It is recommended, though perhaps in guarded language, that these diffusivities can then be used to predict consequent reductions in mass transfer efficiency (*1* to *3*).

Experimental evidence is scarce. Furzer and Ho (*1*) examined some distillation column data reported by Yoshida et al. (*4*). They concluded that the large end effects found there had been caused by axial dispersion. They were able to make an accurate calculation of the effects using dispersion coefficients drawn from the results of unsteady state tracer tests made by De Maria and White (*5*) and by Sater and Levenspiel (*2*). But the earlier work of Brittan and Woodburn (*6*), undertaken specifically to assess the influence of axial dispersion, produced essentially negative results. Their experiments were carried out under steady state conditions and the resulting dispersion coefficients were much less than those used by Furzer and Ho.

The work of Furzer and Ho was later criticized by Mellish (*7*). He pointed out that the dispersion coefficients used in their calculations were almost certainly too large, being derived from experiments in apparatus that would exhibit excessive wall effects.

But there is another more basic objection to the use of such data in predicting mass transfer performance. They are largely irrelevant. Before the measured dispersion coefficients can be used it is necessary to establish that the model is appropriate—that the measured dispersion is caused by backmixing.

Dispersion effects change neither the interfacial area nor the local true coefficients for mass transfer between the streams. They affect the local (and hence the overall) rates of mass transfer only insofar as they modify the concentration profiles.

The adverse result of backmixing has been very clearly described by Miyauchi and Vermuelen (*8*). As each phase enters the column it mixes with material returned from downstream. If there is negligible diffusion across the boundary this mixing produces a step change in concentration at the phase entry. Similarly, the return of some material from the exit region back into the body of the column causes the concentration gradient to fall to zero at the exit. The resulting concentration profiles are illustrated roughly in Figure 1. The boundary phenomena have their effect right through the column. Local concentration driving forces are everywhere reduced below the piston flow values. Such profiles have been observed in certain kinds of extraction equipment.

Backmixing, however, is not the same as dispersion. As Klinkenberg has recently shown in another connection (*9*), the two should be carefully distinguished. Backmixing is only one component of the dispersion measured in unsteady state tracer experiments.

In packed beds a number of mechanisms act in parallel to produce dispersion of fluid particles entering simultaneously (*10*):

Variety of Flow Paths. The available paths have a range of lengths and cross-sectional areas. Thus an important random variation in transit times is introduced. Wall effects may be particularly significant in this regard.

Velocity Gradients. Within each pore there is a velocity gradient between the center and the wall. Once more there is a spread of residence times for different fluid particles.

Dead-End Pores. The existence of stagnant pockets exchanging material to and from the flowing streams adds a further element of variability.

Molecular Diffusion.

Eddy Diffusion. The ordinary mixing action of the two modes of diffusion acts in packed beds as in pipes. The presence of packing, however, restricts the scale of turbulent motion.

Additionally, when there is countercurrent flow: Entrainment of parts of each stream in the second stream moving in the opposite direction.

But only the latter three mechanisms can cause backmixing, a movement of some fluid particles backward relative to the fixed packing as well as the average motion. The others produce dispersion in the downstream direction only. The action might be called forward mixing; it has no influence on the profiles of steady state mean concentration of the flowing streams. (An exception should be made for the case of wall flow which impairs efficiency by producing regions where the L/G ratio differs from the design value.)

It seems that for the practical ranges of flow used in packed towers, and particularly for the liquid phase, forward mixing modes are likely to predominate.

It may be objected that some of the fluid in the packing is nearly or actually stationary; that such fluid would be nearly at equilibrium with the second phase; and that consequently a concentration jump at the entry, as is illustrated in Figure 1, might be expected to occur quite independently of backmixing. And this is partly true. This phenomenon reveals another ambiguity in the description of packed tower operation. The single term concentration is used for the mean concentration at any level, $\int^{dz} c \, dV/V$, and also for the concentration of the net flow, $\int^{dz} uc \, dV/$

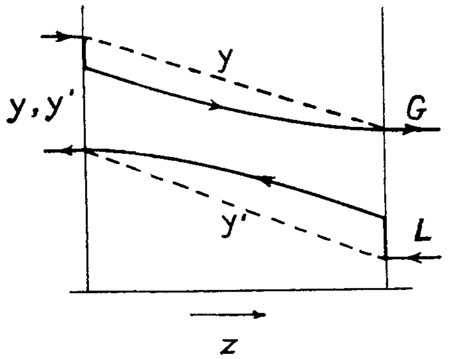

Fig. 1. Concentration profiles in a gas absorption case. With backmixing ———; without backmixing - - - - -.

$\int^{az} u\, dV$. If concentration is correlated with velocity, the two are different: the latter is the concentration in the material balance equations, the former is nearer the correct value to be used in mass transfer rate calculations. This stagnancy effect has been held to account for the significant difference between mass transfer rates in absorption-desorption and vaporization experiments. It is usually dealt with by assuming a reduced effective interfacial area. It appears to produce no effects which depend upon column length.

Forward and backmixing have not been separately measured in tracer tests. But it has been shown that the dispersion can be explained, and with better success, by models other than dispersed plug flow (*11, 12*).

Brittan (*13*) concluded that under ordinary conditions backmixing has only a slight influence on mass transfer efficiency. That conclusion is endorsed. It would appear that the end effects found in the work of Yoshida et al. should be ascribed, instead, to the rather crude liquid distribution system that was employed.

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NOTATION

c = concentration, lb./cu.ft.
G = gas rate, lb./(hr.)(sq.ft.)
L = liquid rate, lb./(hr.)(sq.ft.)
u = velocity, ft./sec.
V = volume of either phase, cu.ft.
z = height of column, ft.
y = mole fraction of solute in gas phase
y' = gas-phase mole fraction in equilibrium with liquid phase

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Compressible Isothermal Slip Flow in the Entrance Region

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Flow development in the hydrodynamic entrance region of a tube and a parallel plate channel in slip-flow regime has been given attention in the recent years. This problem is, in general, treated as an extension of continuum flow with slip at the wall. Sparrow et al. (*1*) linearized the momentum equation with the methods of Langharr and Targ and solved the channel flow, while Hanks (*2*) solved the tube flow with Targ's linearization. Quarmby (*3*) applied an improved linearization by equating the pressure drop based on momentum to that from mechanical energy considerations and solved for both tube and channel. A finite-difference analysis of the boundary-layer equation has also been carried out by Quarmby (*4*). Chen (*5*) used a flat center core and a parabolic boundary-layer velocity profile to integrate the continuity, momentum, and mechanical energy equations and then solved the resulting differential equations numerically.

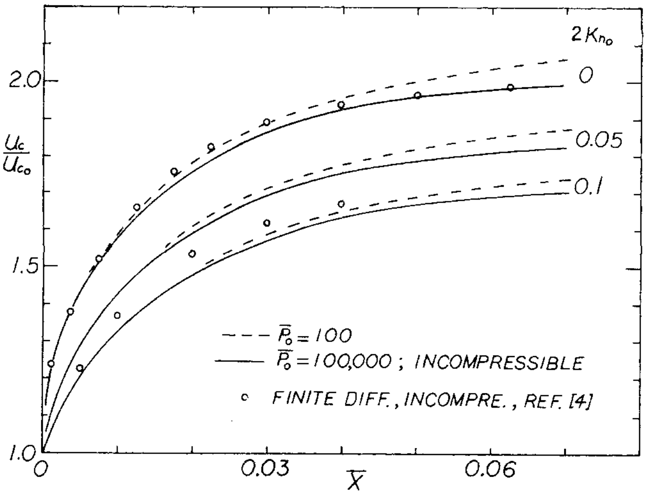

Fig. 1. Centerline velocity development in a tube.

The various analyses so far have been done for incompressible slip flow and are applicable to flow in which $\Delta p/p << 1$. In practical application, however, the pressure drop in the entrance region may be so large that the change in gas density cannot be neglected. In this work the integral method (*5*) is extended to include the effect of compressibility, and a solution for isothermal slip flow in the entrance region of a tube and a parallel plate channel is presented.

THEORETICAL ANALYSIS

Entrance Region of a Tube

For steady flow through a circular tube the continuity and boundary layer momentum equations are