Dispersed Phase Holdup in Baffled Mixing Vessels

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When a two-phase liquid mixture flows through an agitated mixing vessel, one of the liquids becomes dispersed in the other. Unless the agitator power exceeds some moderately large value, the holdup, or volume fraction, of dispersed liquid in the vessel will be different from the volume fraction of that liquid in the feed mixture. Figure 1 shows data typical of those obtained when flow is upward and the lighter liquid is dispersed.

Recently it was shown (Weinstein and Treybal, 1973) that an expression of the following sort successfully correlated such data for covered unbaffled vessels operated full, without an air-liquid interface, and therefore free of vortexing:

$$N_\phi = C\, N_{Pv}{}^a\, N_F{}^b\, N_\rho{}^c N_G{}^d\, N_\mu{}^e \tag{1}$$

The expression may now be shown to apply to similar vessels fitted with baffles, at least for a limited set of conditions thus far studied.

The work considered here (Seewald, 1960) was done with a covered vessel, cylindrical, diameter and height = 0.305 m, fitted with a 6-bladed flat-blade turbine located axially at the center of the vessel, diameter = 0.1525 m, and four equidistant, radial wall baffles, 0.0475 m wide. The vessel was operated with upward flow, full, with no air-liquid interface, and no vortexing. The entire arrangement was similar to that described earlier (Weinstein and Treybal, 1973) except for the baffles. The conditions studied are listed in Table 1. Average holdup of dispersed phase in the vessel was measured by suddenly stopping the flow of entering liquid, allowing the vessel contents to settle, and noting the location of the interface.

The constants for Equation (1) were determined from the data by a least-squares fit. Impeller power was computed (Laity and Treybal, 1957). For impeller power/vessel volume greater than 106 N · m/s · m³ and N_ϕ less than 0.95, 229 runs are correlated by

$$N_\phi = 10^{-0.117}\, N_{Pv}{}^{0.300}\, N_F{}^{-0.178}\, N_\rho{}^{-0.0741}\, N_G{}^{-0.276}\, N_\mu{}^{0.136} \tag{2}$$

with an average deviation of 14.3%.

Since x_F was as large as 0.65 for some runs, Equation (2) is applicable up to $\overline{\phi} = 0.62$. It is possible that phase inversion may have occurred unnoticed at such $\overline{\phi}$ values. Consequently, for the same restrictions on P/v and N_ϕ, but for $\overline{\phi}$ less than 0.5, 214 runs are correlated by

$$N_\phi = 10^{0.1362}\, N_{Pv}{}^{0.310}\, N_F{}^{-0.222}\, N_\rho{}^{-0.0621}\, N_G{}^{-0.295}\, N_\mu{}^{0.204} \tag{3}$$

with an average deviation of 13.9%.

In other work done in this laboratory, a great many measurements were made in a 0.1525-m diameter vessel, but the data cannot be correlated with those from the 0.305-m vessel. This may be related to recent observations of Nienow and Miles (1971), who show, for turbines of the style used here, great variation in the power numbers as the vessel diameter increases from 0.1525 to 0.305 m. Since power numbers are independent of size for diameters 0.305 to 0.458 m (Laity and Treybal, 1957), and for unbaffled vessels holdup correlations include diameters from 0.245 to 0.372 m (Weinstein and Treybal, 1973), it may be that roughly 0.2 m is the lower limit of diameter for general correlations.

The only other detailed study for a baffled vessel is that of Bouyatiotis and Thornton (1963, 1967), the diameter of whose vessel was 0.178 m. Their correlation seems to be confined to agitator speeds which led to uniform dispersions throughout the vessel, with $\overline{\phi}$ not equal to x_F, and this corresponds to impeller power/vessel volume of at least approximately 340 N · m/s · m³. At these power levels or higher, their correlation describes the present data reasonably well, but it does poorly at the lower power levels.

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NOTATION

g = acceleration of gravity, m/s²
g_c = conversion factor, 1 kg · m/N · s²
N_F = $Q_D \rho_C{}^2 \sigma g_c/\mu_C{}^3$, dimensionless
N_G = $\mu_C{}^4 g/\rho_C \sigma^3 g_c{}^3$, dimensionless
N_{Pv} = $P Q_D \mu_C{}^2/v\sigma^3 g_c{}^2$, dimensionless
N_μ = μ_D/μ_C, dimensionless
N_ρ = $\Delta\rho/\rho_C$, dimensionless
N_ϕ = $\overline{\phi}/x_F$, dimensionless
P = power input to impeller, N · m/s
Q = flow rate, m³/s
v = vessel volume, m³
x_F = volume fraction dispersed phase in feed mixture
$\Delta\rho$ = density difference, $\rho_C - \rho_D$, kg/m³
μ = viscosity, N · s/m² = kg/m · s
ρ = density, kg/m³
σ = interfacial tension, N/m

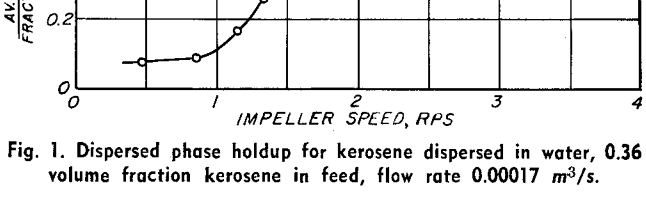

Fig. 1. Dispersed phase holdup for kerosene dispersed in water, 0.36 volume fraction kerosene in feed, flow rate 0.00017 m³/s.

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Table 1. Conditions Studied

Liquid-liquid systems, 25°C: Continuous liquid	Dispersed liquid	Density, kg/m^3 (g/cm^3) (1000) Cont.	Disp.	Viscosity (N · s/m^2) × 1000 = centip. Cont.	Disp.	Interfacial tension, N/m = (dynes/cm)/1000
Water	Kerosene	1000	809.3	0.905	1.364	0.0396
Aqueous $CaCl_2$	Kerosene	1343	811.3	4.775	1.326	0.0383
Aqueous $CaCl_2$	Kerosene	1175	810.5	1.734	1.391	0.0368
Water	Kerosene + isobutanol	990	811.1	1.140	1.327	0.01365

Flow rates: 0.000118 to 0.000255 m^3/s (1.87 to 4.05 gal/min).
Volume fraction dispersed liquid in feed: 0.250 to 0.650.
Impeller speed: 0 to 3.83 rev/s.
Holdup of dispersed liquid in vessel: 0.017 to 0.650.

$\bar{\phi}$ = vessel-average dispersed phase holdup, m^3 disp. phase/m^3 dispersion

Subscripts

C = continuous
D = dispersed

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On Fixed Bed Sorption Behavior of Gases with Nonlinear Equilibria

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In a recent paper Zwiebel, Graiepy, and Schnitzer (1972) have considered the problem of predicting the performance of a fixed bed adsorption column from basic kinetic and equilibrium data. It was shown that for systems exhibiting a favorable Langmuir type of equilibrium relationship the saturation breakthrough curve is much sharper than the regeneration curve obtained under comparable conditions. This difference has been observed in practical adsorption columns (for example, Chi and Lee, 1969).

The above analysis was based on the assumption that the external film offers the rate controlling resistance to mass transfer. In industrial adsorption processes, the adsorbents are generally porous solids and external film resistance is usually small in comparison with internal diffusional resistance, particularly for vapor phase processes. The theoretical curves of Zwiebel et al. are of limited quantitative applicability for such processes.

We have recently considered a similar problem except that the mass transfer rate was assumed to be controlled by internal diffusional resistance (Garg, 1972; Garg and Ruthven, 1973). The analysis was developed to account for the performance of molecular sieve adsorption columns, but a similar treatment is applicable to other systems.

We considered an isothermal constant pressure plug flow system involving sorption of a single component from a dilute mixture. For such a system the differential fluid phase mass balance is

$$v \frac{\partial c}{\partial z} + \frac{\partial c}{\partial t} + \frac{1}{m} \frac{\partial \bar{q}}{\partial t} = 0 \qquad (1)$$

By substituting the variables $X_1 = z/mv$ and $\theta = t - z/v$, Equation (1) may be simplified to obtain

$$\frac{\partial c}{\partial X_1} + \frac{\partial \bar{q}}{\partial \theta} = 0 \qquad (2)$$

The appropriate initial and boundary conditions for a step change in sorbate concentration at the column inlet, at time zero are

$$\bar{q}(X_1, 0) = 0, \quad c(0, \theta) = c_0 \qquad \text{(saturation)} \qquad (3)$$

$$\bar{q}(X_1, 0) = q_0, \quad c(0, \theta) = 0 \qquad \text{(regeneration)} \qquad (4)$$

The equilibrium isotherm is assumed to obey the Langmuir equation

$$\frac{q^*}{q_s} = \frac{bc}{1 + bc} \qquad (5)$$

and the kinetics of sorption are assumed to be controlled by micropore diffusion within the zeolite crystal with instantaneous equilibration at the crystal surface. The zeolitic diffusivity is taken to vary with concentration according to the expression (Ruthven and Loughlin, 1971):