Mass Transfer to Suspended Particles in Gas-Liquid Agitated Systems

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Introduction

Mass transfer to suspended particles in gas-liquid systems is important in a wide variety of processes such as aerobic fermentations, waste treatment, and three-phase slurry reactions. In ungassed, agitated systems mass transfer to particulates is dependent upon a variety of geometric, physical, and hydrodynamic properties. A number of successful correlations for mass transfer to particles in ungassed, agitated systems have been developed based upon a modified particle Reynolds number that is derived from the power dissipated in the vessel (Levins and Glastonbury, 1972; Brian et al., 1969; Nienow, 1967; Middleman, 1965; Armenante, 1983).

$$Re_p \equiv d_p^{4/3}\epsilon^{1/3}/\nu \qquad (1)$$

The term ϵ is the average power input to the fluid per unit mass which can be obtained from numerous correlations for agitator power in liquid systems. (Bates et al., 1963; Nagata, 1975).

The correlations for liquid-solid transfer in agitated vessels are typically of the form

$$Sh = 2 + A\, Re_p^x Sc^y \qquad (2)$$

where the power on the Sc number is most commonly ⅓.

Despite the wealth of literature on unaerated liquid-solid mass transfer systems, very little information is available for the three-phase systems. Sano et al. (1974) evaluated mass transfer coefficients between a liquid and solid particle in which agitation was provided either by an impeller or by aeration in a bubble column. They found that the same mass transfer correlation could apply when ϵ was calculated either from agitator power or from the isothermal expansion of the gas in the bubble column. Sanger and Deckwer (1981) measured solid-liquid transfer in a bubble column where ϵ was calculated from the gas flow and pressure drop through the column. Their results were in reasonable agreement with the correlation of Sano et al. and with ungassed liquid correlations such as that of Levins and Glastonbury.

In that there appear to be no investigations of mass transfer to particles in agitated liquids with simultaneous gas flow, we chose to investigate this condition. The flow of gas through a sparger and into an agitated liquid can be expected to have a number of effects on mass transfer. First, the impeller power is expected to be reduced (Nagata, 1975; Michel and Miller, 1961). Second, the gas flow dissipates power to the liquid in two ways (Vogel, 1983). The gas undergoes a nearly isentropic expansion through the sparger, resulting in an increase of kinetic energy which is then dissipated in the liquid. The gas leaving the sparger also undergoes an approximately isothermal expansion to the pressure in the head space of the vessel. Finally, the introduction of the gas affects the impeller power necessary to maintain the particles fully suspended. The total power dissipated per unit mass of liquid is

$$\epsilon = (P_I + P_G)/M_L \qquad (3)$$

The impeller power is a function of gas flow as well as geometry, physical properties, and operating parameters. The power input from the gas is the sum of the contributions of the isentropic and isothermal expansion terms.

$$P_G = PV_o\left\{\frac{k}{k-1}\left[(P_0/P_1)^{k/(k-1)} - 1\right] + \ln\frac{P_1}{P_2}\right\} \qquad (4)$$

where the pressure at the bottom of the vessel is calculated from the height of ungassed liquid in the tank ($\rho_L gh$).

The purpose of this investigation was simply to examine whether particle fluid mass transfer coefficients could be correlated with total power in the usual fashion if the power was calculated as the sum of the impeller and gas contributions. In other words, did the mode of introducing turbulence to the liquid

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phase affect the coefficient if the same power dissipation occurred?

Experimental

Details of the experiments may be found in the thesis by Marrone (1985). The basis of the measurement of the solid-liquid mass transfer coefficient was to observe, with a Na+ ion electrode, the uptake of NaOH by ion exchange resins in a gassed, agitated tank. Previous investigators have utilized this technique and established conditions under which it measures a mass transfer coefficient rather than a kinetic process within or at the surface of the resin (Armenante, 1983; Harriott, 1962; Sanger and Deckwer, 1981). Particles of Dowex 50-W8^R cationic resin of Sauter mean diameters 48.7 and 88.9 μm were used along with Aminex^R 17.5 μm cationic resin.

The experiments were carried out with a Chemineer Model ELB agitator kit with a variable speed drive and dynamometer for torque measurement. The test vessel was a fully baffled flat-bottom tank with diameter and ungassed liquid height equal to 24 cm. The impeller was a 6.4 cm, six-blade disk turbine located 6.4 cm from the tank bottom. Nitrogen gas was introduced (at a measured temperature, pressure, and flow rate) into the tank bottom through a perforated plate containing 45 holes of 1 mm dia. drilled in three concentric circles. A brass screen was stretched over the plate to distribute the gas evenly and prevent loss of particles through the bottom. A minimum flow rate of 1.5 L gas/L solution per min was necessary to provide a uniform gas distribution across the plate. This uniform gas distribution maintained a good dispersion of particles, as opposed to the situation where all the gas was introduced directly under the impeller. A tank lid provided support for the electrodes and thermometer. Nitrogen gas was also sparged into the head space to prevent CO_2 dissolution in the liquid.

For both gassed and ungassed systems the impeller power was determined by torque and speed measurements. For ungassed systems the measured power agreed with the correlation of Bates et al. (1963) within about 25%.

Results and Discussion

The measured mass transfer coefficients are correlated according to Eq. 2 in Figure 1. The power input term, ϵ, in the Reynolds number is the sum of that from the impeller and from gas flow. The four different symbols identify different ratios of gas to total power ranging from 0 to 1. There does not appear to be any significant trend of the data points with the mode of power input to the fluid.

Under the assumption that the Sherwood number varies as the Schmidt number to the one-third power, a least-squares correlation of the data points in Figure 1 is $Sh = 2 + 0.36\ Re_p^{0.75}\ Sc^{1/3}$. The exponent on the Reynolds number, in fact, is primarily determined by the three different particle diameters employed, as Re_p depends on the power input to only the one-third power. In Table 1 the results of this work are compared to those of Levins and Glastonbury for unaerated liquids, to Sanger and Deckwer for bubble columns, and to Sano et al. for bubble columns and agitated vessels. The results are all in reasonable agreement.

It does appear that the mass transfer coefficient in gas-liquid-solid systems may be predicted with reasonable accuracy by simply using the total power input to the fluid. Additional study would be required to extend this conclusion to different modes of gas sparging or impeller design.

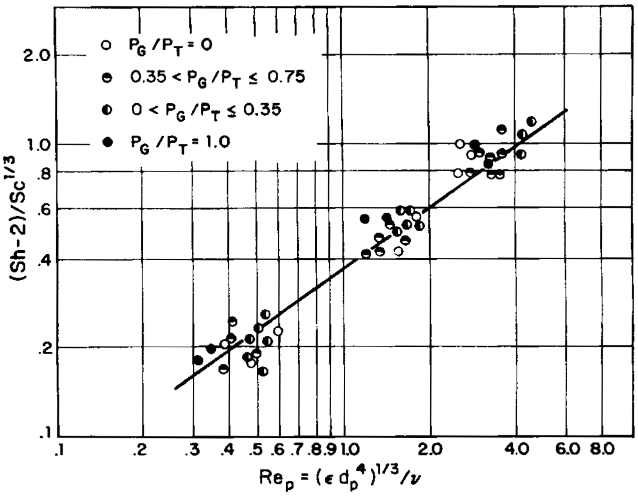

Figure 1. Correlation of solid-liquid mass transfer in gassed, agitated vessels.

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Notation

dp = particle diameter, m
g = acceleration due to gravity, m · s^{-1}
h = height of ungassed liquid in vessel, m
k = ratio of specific heats of nitrogen
M_L = mass of liquid in tank, kg
P_o = pressure before gas distributor plate, Pa
P_1 = pressure after gas distributor plate, Pa
P_2 = head space pressure, Pa
P_g = power input to fluid from gas, W
P_I = power input to fluid from impeller, W
V_o = gas volumetric flow rate upstream of distributor plate, m^3 · s^{-1}

Greek letters

ϵ = power dissipated per unit mass of liquid, m^2 · s^{-3}
ρ_L = liquid density, kg · m^{-3}
ν = kinematic viscosity of liquid, m^2 · s^{-1}

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Table 1. Correlations of Solid-Liquid Mass Transfer

Correlating Equation	Reference
$Sh = 2 + 0.36\ Re_p^{0.75}\ Sc^{1/3}$	This work
$Sh = 2 + 0.50\ Re_p^{0.62}\ Sc^{1/3}$	Levins and Glastonbury (1972)
$Sh = 2 + 0.55\ Re_p^{0.79}\ Sc^{1/3}$	Sanger and Deckwer (1981)
$Sh = 2 + 0.45\ Re_p^{0.76}\ Sc^{1/3}$	Sano et al. (1974)

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