Use of an Oldshue-Rushton Extractor as a Gas-Slurry Contactor

D. J. Gray
Department of Chemical Engineering
University of Rhode Island
Kingston, RI 02881

There are a substantial number of chemical engineering processes requiring the intimate contact of a gas with a liquid in which a finely divided solid is suspended. Present techniques often require extensive equipment and maintenance in order to obtain the desired product. In purely diffusive or mass transfer operations, such as those involving absorption of a gas into a slurry of an absorbent and subsequent adsorption of the dissolved gas, the equivalent of a countercurrent multistage operation will be required if an appreciable percent removal of solute gas from the feed mixture is expected. The multistage agitated column shown in Figure 1, originally designed by Oldshue and Rushton (1952) for liquid-liquid extraction, was investigated for use as such a countercurrent gas-slurry contactor.

Experimental Equipment and Procedure

The column is a 0.1524 m dia., 13-stage, fully baffled, mechanically agitated vessel employing countercurrent flow of gas with a liquid-solid slurry. The stages are 0.0834 m in height and are formed by stainless steel annular baffles having a central opening of 0.0826 m dia. The top stage is 0.4752 m in height in order to allow space for liquid and solid feeding. Each stage was agitated with a 0.0508 m dia., six-blade, flat disc type turbine with blades 0.0127 m wide and 0.0095 m high.

Work was done using water slurries of limestone, sulfur, coal, and PVC pellets of varying particle sizes. The solid properties of interest are listed in Table 1. The water slurries were premixed, and were fed and removed from the top and bottom of the column using constant-volume displacement Moyno pumps.

Air, at rates measured by a rotameter, entered the column through a side port in the bottom stage. As the air was being dispersed, the liquid level in the column was maintained by an overflow port in the top stage. No collection of air was necessary at the top of the column.

Volume samples of the slurry feed and discharge streams were measured, filtered, dried, and weighed to determine the solid volume percent of the two streams. Approximately five turnovers of the column were required before these samples became equal, indicating steady state conditions, at which point the Moyno pumps and air inlet valves were shut down simultaneously. After clearing all the air from the column, the liquid level was recorded to determine the gas holdup by the bed expansion method. The holdup slurry was then removed, agitated to uniformity in a smaller tank from which samples were taken, and analyzed as above to determine the volume percent solid holdup based on the liquid and solid volume only. The uniformity of the mixture in the sampling tank was periodically checked by weighing the entire solid holdup removed from the column after oven drying for a number of runs for each solid used. The range of conditions studied within the 13-stage column are listed in Table 2.

The mechanical power input, being a better indicator of the degree of mixing than the impeller speed, was also investigated using a geometrically similar three-stage column in order to eliminate the steady bearing friction effects in the larger column. The power consumption was measured using a digital readout torque meter with the torque transducer attached between a variable speed motor shaft and the impeller shaft. Both the aerated and nonaerated power numbers were determined at various impeller speeds and aeration rates.

Results

Power consumption of impellers

Gas rates and holdup were measured by the same methods used in the 13-stage study. Power input measurements were made only for air-water semibatch systems (continuous phase not being fed or removed from the column).

The data for the power consumed by impellers agitating only water were correlated using the conventional power number for single-phase mixing. The average power number ($N_p = P/N^3D^5\rho_L$) for a single impeller mixing within the turbulent regime was found to be 3.92. Uhl and Gray (1967) report power numbers closer to 5 for six-blade disc impellers operating within the turbulent regime. The reduction in the power number in the column study is attributed to the low clearance between the impeller and the annular plates, which promotes a single figure-

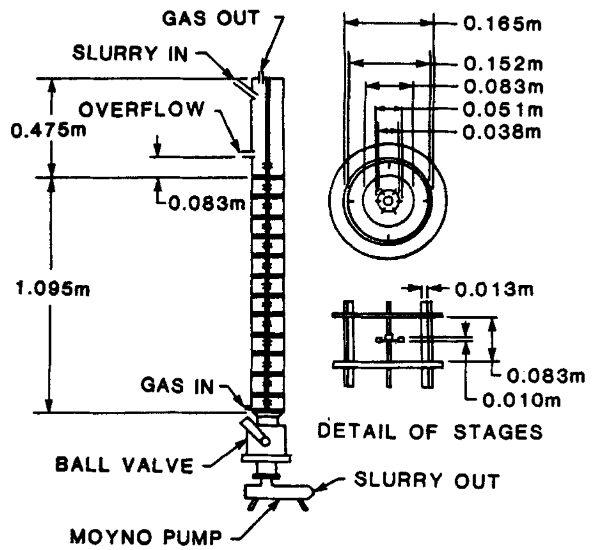

Figure 1. Agitated multistage countercurrent gas-slurry contactor.

eightlike flow pattern as opposed to the conventional double-eight flow pattern found for impeller clearances of at least one impeller diameter in baffled tanks (Nienow, 1968). Similar reductions in N_p have been found by Conti et al. (1981) and Bates et al. (1963) for low off-bottom impeller clearances in baffled tanks. Conti et al. report a change in the flow pattern for an eight-blade disc impeller in a baffled tank at a C/T value of 0.22 independent of the impeller diameter. Within the column the single-eight flow pattern was observed for the C/T value of 0.26 used in this study.

The power consumed in the two-phase gas-liquid system was found to vary with the aeration number, similar to the findings in baffled tanks (Calderbank, 1958). However, the gassed to ungassed power ratio was observed to also vary with the impeller speed, as shown in Figure 2, similar to the more recent findings of Van't Riet et al. (1977), Warmoeskerken and Smith (1982), and Kuboi and Nienow (1982) in baffled tanks. This is attributed to the greater gas recirculation to the impeller as well as the changing cavity formation characteristics behind the impeller with increasing impeller speeds. With the exception of the data at 40 rps at higher aeration numbers, the trend is for the power ratio to decrease with increasing impeller speeds at constant aeration number. The curve drawn for 40 rps in Figure 2 is the expected curve, since the data at higher aeration numbers were questionable when compared to the overall trend. The gassed to ungassed power ratio can be estimated for aeration numbers greater than 0.04 by the equation

$$P_G/P_O = 2.3\,N^{-0.55} - 0.5\,(Q/ND^3) \tag{1}$$

Table 1. Characteristics of Solids

	Density kg/m³	Avg. Size μm	Avg. Settling Veloc. m/s
Sulfur	1,940	75	slow
Limestone	2,880	125	0.028
Coal	1,590	1,500	0.075
PVC pellets	1,341	3,250	0.121

Table 2. Range of Conditions Studied

Solids rate	$0.033–0.33\ m^3/s \times 10^{-4}$
Liquid rate	$0.08–0.108\ m^3/s \times 10^{-3}$
Gas rate	$0–0.2034\ kg/s \cdot m^2$
Agitator speed	10.0–33.33 rps

where Q/ND^3 is the aeration number and Q is the volumetric gas rate.

Gas holdup

The gas holdup is reported as the volume fraction of the active column volume. Previously, Sullivan and Treybal (1970) studied the dependence of gas holdup on the superficial liquid and gas rates and impeller speeds for air and water in this same column. The gas holdup ϕ_G for the two-phase air-water system can be obtained from the equation

$$\phi_G = 0.2\,V_G^{0.2}P_G^{0.4} + 2.1\,V_{SL} \tag{2}$$

The total gas holdup was the sum of the gas dispersed throughout the liquid and the gas trapped under the annular plates. Both of these holdups increased with increasing gas rate, while increasing the impeller speed increased the overall holdup and decreased the holdup trapped beneath the plates.

At high gas rates and low impeller speeds the impellers became flooded with air and the rising gas approached slug flow as in highly aerated bubble columns. The dependence of this transition point to the gas rate and impeller speed was not determined, however, for the maximum superficial gas rate of 0.16 m/s; the air was dispersed throughout the liquid at impeller speeds greater than 23 rps. The transition to bubble slug flow in bubble columns generally occurs below 0.08 m/s superficial gas velocity (Taitel et al., 1980) at a maximum holdup between 20 and 25 vol. % gas (Eissa and Schugerl, 1975; Kelkar et al., 1984; Kato et al., 1973) for air-water systems. This transition point

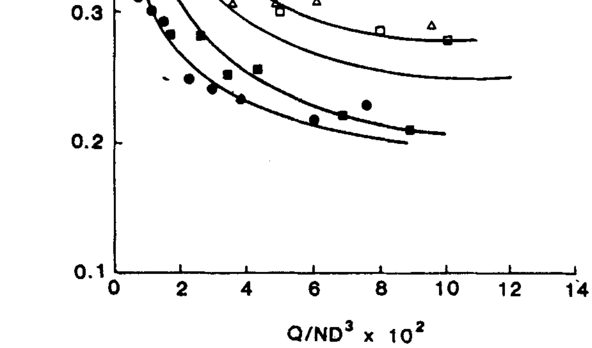

Figure 2. Gassed to ungassed power ratios vs. aeration number.

represents an optimum operating condition, since above this gas rate bubble size is increased and holdup decreased due to the faster rising slug bubbles. In addition to doubling this optimum gas rate, the agitated column produced gas holdups in excess of two times that observed in the bubble column studies at the optimum gas rate.

The dependence of the gas holdup on the solids rate is shown in Figure 3. The broken line in Figure 3 is the projected curve to zero solid concentration as calculated from Eq. 2. As can be seen in the figure, a substantial increase in the gas holdup is expected with only a small amount of limestone, coal, or sulfur present. It is believed that the solids coat the air bubbles as in flotation operations, thus reducing the rate of rise of the bubbles through the column. This coating effect was readily observed in the stable foam produced by limestone and by coal when aerated in a batch system, whereas PVC pellet slurries produced no foaming.

The coating of the air bubbles should be dependent upon the size and wettability of the particles employed. The coal, although initially large particles, does break up slightly during premixing and feeding through the Moyno pump. These smaller particles were observed when filtering samples to determine the feed and solid holdup concentrations. The exact size and concentration range was not determined, however the smaller particles would be classified as coal dust rather than coal particles. As can be seen in Figure 3, only a small concentration of small particles would be needed to effectively increase the gas holdup. The wetting angle of the solids used in this study was not measured; however, all the solids can be classified as nonwettable. Additional work in this area is needed before a correlation of the data can be made.

The overall effect of the solids concentration on the gas holdup was to decrease the holdup with increasing concentration. The effect is twofold. First, the settling of solids through the column produces a positive upward displacement of the gas. Second, the presence of solids within the liquid phase essentially reduces the active volume in which the gas phase may be dispersed. Kato et al. (1973) and Kelkar et al. (1984) both have reported a decrease in the gas holdup with increasing solids concentration for air-water contacting in bubble columns.

An examination of the slopes of the lines in Figure 3 also indicates the existence of two distinct types of gas-slurry contacting. For the larger particles the gas holdup is less sensitive to the solids concentration. Larger particles were able to penetrate the gas bubbles coming in contact with the gas and breaking the bubbles into smaller, slower rising bubbles. The smaller particles did not contain the necessary inertia to overcome the surface tension forces and break up the bubbles. For the smaller limestone and sulfur particles, the gas holdup decreases more rapidly with solids concentration as compared to the larger coal and PVC particles.

Solids holdup

Figure 4 shows the effect of increasing solids rate on the solids holdup. The dependence was almost linear, with a slight curvature due to hindered settling at higher solids feed rates. The rate of increase of solids holdup to the solids feed rate was dependent upon the settling velocity of the solids, and to a lesser degree on the slurry rate through the column.

Figure 5 is a cross plot at constant feed concentration from curves similar to Figure 4 to show the effect of the impeller speed on the solids holdup. For impeller speed ranges from 10 to 33 rps, little change in the solids holdup was observed. Sulfur-water slurries approached a solids holdup concentration of approximately 90% of the feed concentration. Limestone, coal, and PVC slurries reached holdups of approximately 45, 20, and 15% of the feed concentration, respectively. The smaller, less dense solids such as sulfur were more uniformly mixed at the impeller speeds studied. The heavier solids could not be suspended off the annular plates, and it was determined that a power input of fifteen times that for suspension in flat-bottom baffled tanks was required to suspend solids off the annular plates.

Increasing the gas rate effectively increased the slurry rate through the column and reduced the solids holdup.

Conclusions

The results indicate that each stage within the multistage column behaves in a similar manner to a single baffled vessel. The

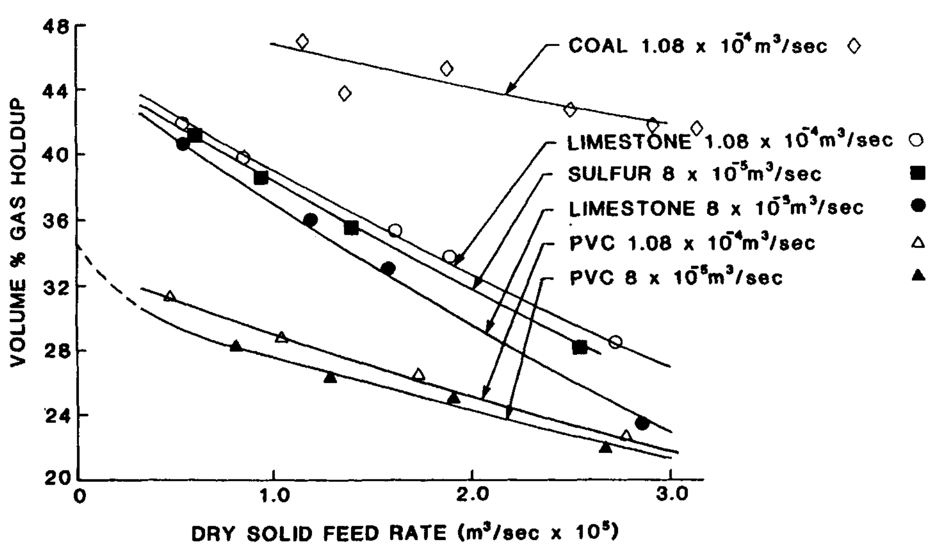

Figure 3. Gas holdup variation with dry solids feed rate at constant slurry rate.
Gas rate, 4.45×10^{-4} kg/s · m²; impeller speed, 33.3 rps

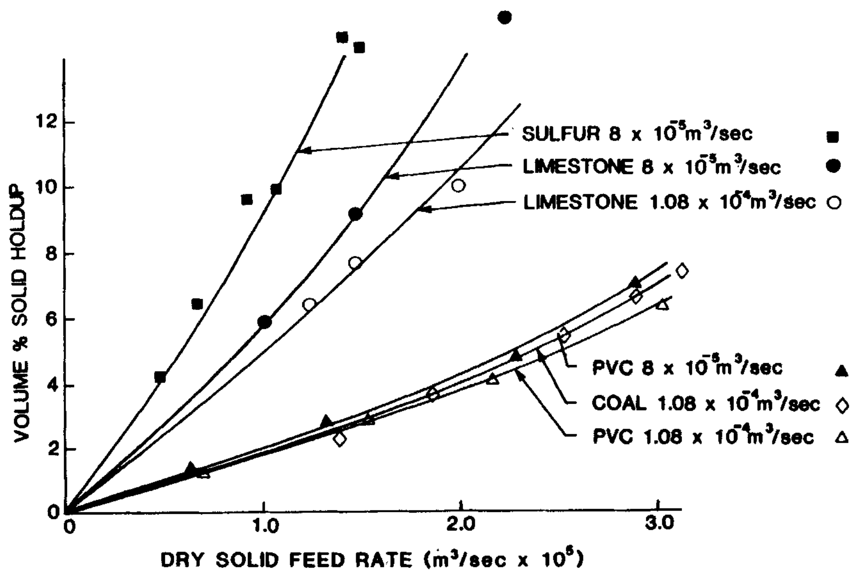

Figure 4. Solids holdup variation with dry solids feed rate at constant slurry rates.
Impeller speed, 33.3 rps; no gas

power number for nonaerated disc impellers within the column is shown to be of the same magnitude as within baffled tanks. For aerated systems, the gassed to ungassed power consumption ratios vs. aeration number curves are comparable to baffled tanks, and the gas holdup dependence on the power input is identical to single-stage vessels.

Compared to bubble columns, the multistage column can handle twice the gas rate at the optimum operating conditions for the impeller speeds studied. Under these conditions more than twice the gas holdup can be attained.

The column shows limitations to the size and density of the particles that can be handled. As compared to baffled tanks, fifteen times the power input is required to completely suspend the solids.

For slurries of lighter particles, the agitated column can provide much better controls and mass transfer capabilities than bubble columns, at the expense of power costs. Lower column heights than bubble columns and/or less floor space than a cascade of single vessels can be expected to provide a desired product.

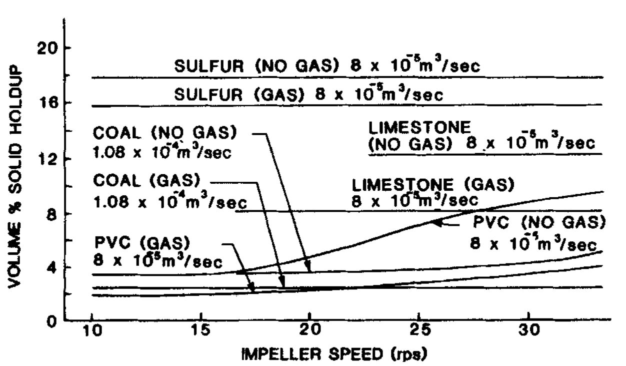

Figure 5. Solids holdup variation with impeller speed at constant slurry rates.
Gas rate, 4.45 × 10^{-4} kg/s · m^2

Notation

C = impeller clearance, m
D = impeller diameter, m
N = impeller rational speed, s^{-1}
N_p = power number, $P_o g_c / N^3 D^5 \rho_L$
P_o = nonaerated impeller power input, W/s (aerated impeller power input, W/s)
P_G = aerated impeller power input, W/s
T = tank diameter, m
Q = gas volumetric flow rate, m^3/s
V_G = superficial gas velocity, m/s
V_{SL} = superficial gas-liquid slip velocity, m/s
ϕ_G = fractional volumetric gas holdup
ρ_L = liquid density, kg/m^3

Literature cited

Bates, R. L., P. L. Fondy, and R. R. Corpstein, "An Examination of Some Geometric Parameters of Impeller Power," *Ind. Eng. Chem. Process Des. Dev.*, **2**, 310 (1963).
Calderbank, P. H., "Physical Rate Processes in Industrial Fermentation. I: The Interfacial Area in Gas-Liquid Contacting with Mechanical Agitation," *Trans. Inst. Chem. Eng.* (London), **36**, 443 (1958).
Conti, R., S. Sicardi, and V. Specchia, "Effect of the Stirrer Clearance on Particle Suspension in Agitated Vessels," *Chem. Eng. J.*, **22**, 247 (1981).
Eissa, S. H., and K. Schugerl, "Holdup and Backmixing Investigations in Cocurrent and Countercurrent Bubble Columns," *Chem. Eng. Sci.*, **30**, 1251 (1975).
Kato, Y., A. Nishiwaki, T. Kago, T. Fukada, and S. Tanaka, "Gas Holdup and Overall Volumetric Adsorption Coefficient in Bubble Columns with Suspended Solid Particles," *Int. Chem. Eng.*, **13**, 562 (1973).
Kelkar, B. G., Y. T. Shah, and N. L. Carr, "Hydrodynamics and Axial Mixing in a Three-Phase Bubble Column. Effects of Slurry Properties," *Ind. Eng. Chem. Process Des. Dev.*, **23**, 308 (1984).
Kuboi, R., and A. W. Nienow, "The Power Drawn by Dual Impeller Systems under Gassed and Ungassed Conditions," *4th Eur. Conf. on Mixing*, Paper G2 (1982).
Nienow, A. W., "Suspension of Solid Particles in Turbine-Agitated Baffled Tanks," *Chem. Eng. Sci.*, **23**, 1453 (1968).
Oldshue, J. Y., and J. H. Rushton, "Continuous Extraction in a Multistage Mixer Column," *Chem. Eng. Prog.*, **48**, 297 (1952).
Sullivan, G. A., and R. E. Treybal, "Axial Mixing and Gas Adsorption

in a Mechanically Agitated Absorption Tower," *Chem. Eng. J.*, **1**, 302 (1970).
Taitel, Y., D. Bornea, and A. E. Dukler, "Modeling Flow Pattern Transitions for Steady Upward Gas-Liquid Flow in Vertical Tubes, *AIChE J.*, **26**, 345 (1980).
Uhl, V. W., and Gray, J. B., *Mixing Theory and Practice*, Academic Press, New York, **1**, 133 (1966).
Van't Riet, K., J. M. Boom, and J. M. Smith, "Power Consumption, Impeller Coalescence and Recirculation in Aerated Vessels," *Trans. Inst. Chem. Eng.*, **54**, 124 (1976).
Warmoeskerken, M. M., and J. M. Smith, "Description of the Power Curves of Turbine-Stirred Gas-Liquid Dispersions," *4th Eur. Conf. on Mixing*, Paper G1 (1982).

Manuscript received May 14, 1986, and revision received Sept. 29, 1986.