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SIMPLE PERFORMANCE CORRELATIONS FOR AGITATED VESSELS

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ABSTRACT

This paper considers simplified correlating equations for preliminary design of process equipment. The uncertainties associated with process change are much greater than inaccuracies resulting from the streamlining of the original equations.

Predictions are presented for gas hold-up and mass transfer rates in aqueous aerated mixtures which are independent of process scale. Experimental data shows that these simple dimensionless correlations can accommodate a wide range of process scales and many impeller and tank geometries.

INTRODUCTION.

The spectrum of gas liquid processes sets daunting requirements on those concerned with design specification and performance prediction. During recent years our understanding of these systems has increased very greatly, almost to the extent that the complexities threaten to mask any possible underlying simplicity. The infinite variety of process conditions has tended to make designers and operators suspicious of general rules - the every case is different syndrome - to the extent that work carried out say with air and water - is dismissed as generally irrelevant to process needs. At the same time those concerned with computational fluid mechanics are quite pleased to assume that their programs will solve flow fields - for air-water or any other species. It must be allowed that some processes - notably those associated with coalescence phenomena and hence a whole group of transfer performance indices - are not yet at the point of rigorous solution. Nevertheless, outside this group of phenomena there is real progress to be made.

The search for reliable predictive and scaling rules for process equipment has led to the development of a multitude of correlating relationships. Sometimes these have been dimensionally inconsistent whilst in other cases an unrealised commonality in the correlated variables has misleadingly enhanced apparent value of the method presented. (Michel and Miller (1962), Nienow (1977)) The present paper demonstrates that dimensionless correlating equations can be used to establish the hydrodynamic conditions and on this basis to predict useful performance data. Knowledge of the relative importance of any differences between conditions used in establishing the correlation and those in the operating plant allows sensible assessment of both the value and limitations of the proposed equations.

Much of the literature relating to gas-liquid operations has been established in academic studies carried out in "standard" geometry equipment at small scale. This limitation is perceived by industry as being significant and severe. Unfortunately it is both expensive and time consuming to carry out scientific investigations in process scale equipment: there is a need to establish guidelines for the use of small scale data to predict performance in large scale applications, including some assessment of validity and reliability. Ideally the extrapolation procedures should allow extension to somewhat modified conditions, e.g. to reactors using multiple impellers on a single shaft.

Six blade Rushton impellers have been the usual choice for gas dispersion and mass transfer, while pitched blade impellers have featured for solids suspension. In recent years many publications have dealt with a succession of profiled impellers - hydrofoils and their derivatives - which have been particularly successful in applications requiring efficient pumping of large volumes of liquid, e.g. heat transfer, large scale liquid blending and solids suspension. The widely distributed and low power dissipation of these impellers makes them rather suitable for liquid- liquid dispersion; one would expect a rather more uniform drop size distribution than can be achieved with a disc turbine for example.

Although this paper is presented as relevant to Rushton turbines, the principles are more general.

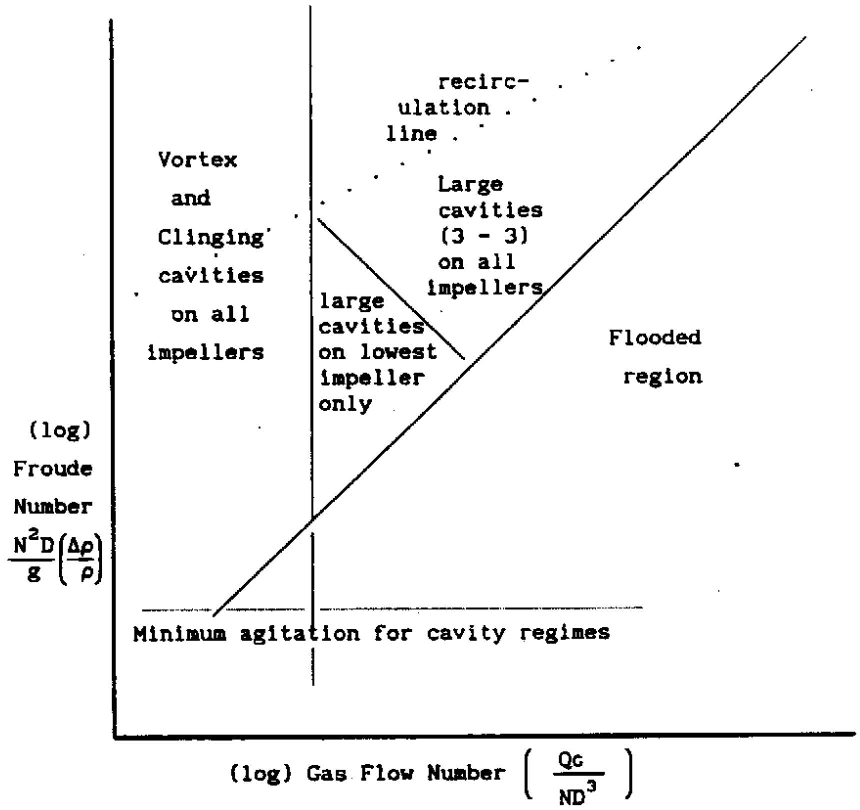


Fig. 1. Generalised Flow Map for Rushton Turbines.

Note. As will be evident from the equations presented by Smith et al. the precise positions of these lines depends on the geometry of the vessel. The minimum agitation line is at a Froude number of about 0.03, the cavity formation line at a gas flow number of about 0.03, the flooding line has a slope of exactly 1.0, corresponding approximately to Fl = 1.2*Fr. The complete dispersion line has a slope of exactly 0.50 and crosses the cavity development line at a Froude number of about 0.5.

THE ROLE OF FLOW MAPS

It is recognised that impellers operating in gas liquid systems develop local flow regimes which depend on the conditions in ways which can be quantified in terms of key dimensionless groups, Smith and Warmoeskerken (1985). Conditions in the immediate vicinity of the impeller control the energy transfer from the shaft to the fluid, both in terms of the (directional) pumping and its (pseudo) random turbulent motion. The near field processes include the regions of most rapid micromixing, and local phase dispersion. The bulk convective flows generated by the impeller determine the conditions further away, notably possible recirculation of dispersed phase to the impeller, macroscopic mixing time, solid suspension or drawdown, and heat transfer to containing surfaces or internals.

Second phases do not always affect the flows significantly: moderate amounts of solid or a second dispersed liquid phase are of minor influence whereas the effects of dispersed gases cannot be ignored.

Flow Maps.

Gas liquid flow maps provide a very useful tool for establishing the effects. With a Rushton turbine, the impeller flow regime may involve vortex-clinging cavities, large cavities or be flooded. The vast majority of industrial scale operations take place in the large cavity regime, in which alternate larger and smaller cavities are held captive after successive impeller blades. The flow map, Fig 1, which allows prediction of hydrodynamic conditions in vessels agitated with either single or multiple Rushton impeller, delineates the main areas of operation, (Smith et al., 1987). The transitions are defined by dimensionless criteria.
The straight lines on this graph correspond to:
vertical - the constant flow number at which large cavites develop,
horizontal - the minimum agitation at which cavities adhere to the impeller blades,
slope 1.0 - the flooding to loading transition,

slope 0.50 - the recirculation line above which dispersion is complete and virtually uniform throughout the vessel.

Comparisons between correlating equations used for single and multiple impeller systems give information about differing conditions at various locations within the vessel. The flow map shows that conditions in the upper part of the vessel differ significantly from those near the base.

with great accuracy. The sporadic overflows occasioned by waves in the free liquid surface were easily accommodated by this system. The measurements differ slightly from those conventionally obtained from changes in the free liquid surface level in that the volume of liquid in the vessel is reduced slightly under the highest conditions of gas retention: this will only be significant at very large void fractions, say above 10%.

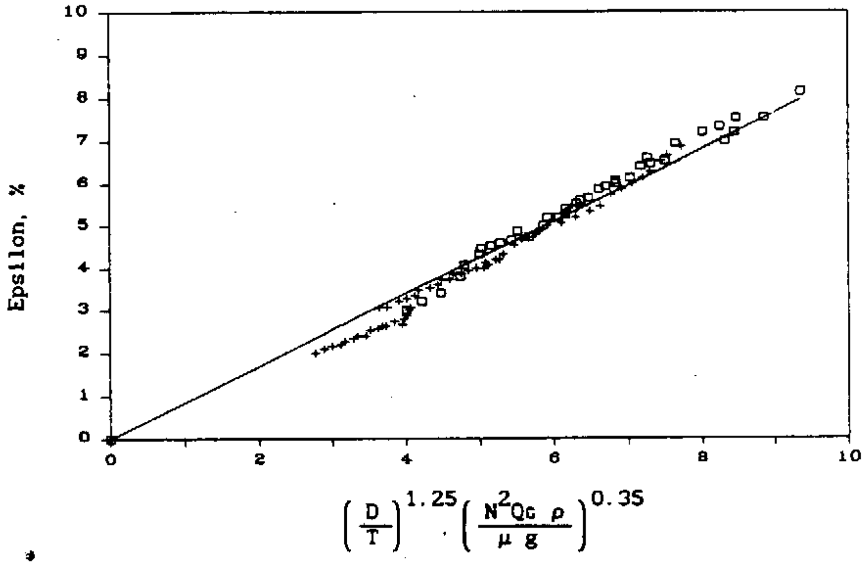


Fig. 2. Void fraction in standard vessels, Delft data.
Tank diameter + T = 0.44 m □ T = 1.20 m tank.

CORRELATION OF GAS FRACTION RESULTS.

Experimental determination of the gas retention in agitated vessels is complicated by the inherent instability of the surface. The simplest method is to measure the change in liquid level during operation though this may be difficult because of the constantly fluctuating free liquid surface. Subjectiveness can be reduced by detecting the position of the free liquid surface electronically. Various methods have been used for this, including ultrasonic, capacitance and conductivity probes, with a variety of traversing and remote sensing arrangements. Unfortunately these methods are often affected by the presence of even small quantities of foam on the liquid surface and there is a real and persistent difficulty in deciding the most representative position(s) on the liquid surface for the determination of the change in level. The measurements that are reported here utilise a slightly different method; the liquid in the vessel was allowed to overflow through a funnel to a measuring cylinder from which it was pumped back at a steady rate to the vessel. The change in the level in the measuring vessel could be very accurately determined. The steady flow conditions give an unvarying holdup in the return pump system and so that the volume of gas retained in the vessel could be measured

Void Fraction: Single Impellers in Standard Tanks.

In good agreement with Hassan and Robinson (1977), it was found that the gas fraction correlates linearly with the volumetric gas rate and the square of the impeller speed. This group occurs in the product (Flc•Fr•Re), which, like the Flow Map, is independent of scale. Results for two scales of equipment (T = 0.44 m and T = 0.61 m) are shown in Fig. 2.

Almost all the data points represent conditions in which the single Rushton impeller is operating in the large cavity regime; the condition of interest for most industrial operations. The lack of scatter in these data is remarkable. The correlating equation has the form

$$\varepsilon_G = \text{const} \cdot (\text{Re} \cdot \text{Fr} \cdot \text{Flc})^{0.35} \qquad (1)$$

The independence of scale is confirmed in data which has kindly been made available by BHRA and ICI. The ICI data, (measured by Middleton), and BHRG (Fluid Mixing Processes consortium, measured by Muskett) are shown in Fig. 3. against the same correlating line used in Fig. 2.

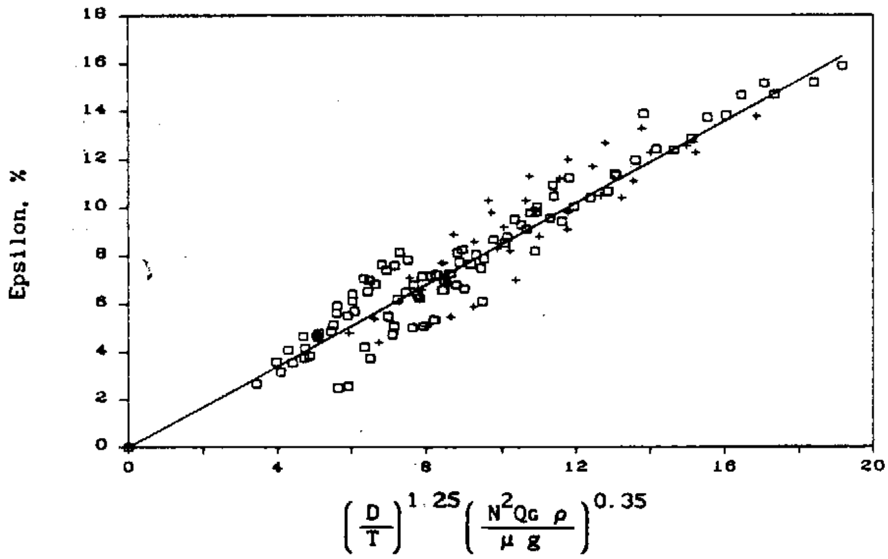


Fig. 3. Industrial measurements of Gas Fraction
+ ICI data T = 1.8 & 2.1 m □ BHRG data T = 0.6, 1.6 & 2.7 m

The data represented in these graphs has all been obtained in air-water systems. The largest equipment was 4.2 m diameter. D/T ratios were in the range 0.3 to 0.5, and most experiments were carried out with H/T = 1 in standard tanks. In the multiple impeller results referred to below, the vessels up to 1m in diameter, with H/T ≈ 3 and impeller separations ≥ D. Bottom clearances, C/T, were usually 0.25 or 0.33.

TANK TO IMPELLER DIAMETER RATIO.

The industrial data has been used to establish the (D/T) dependence to the power 1.25. Although although the data show rather greater scatter than that from the Delft laboratory study, the good fit is extended to much higher gas fractions. It is likely that in a coalescing system a transition at a gas fraction above perhaps 25% there will be a transition in flow regime to something corresponding to the churn turbulent regime in a vertical pipeline.

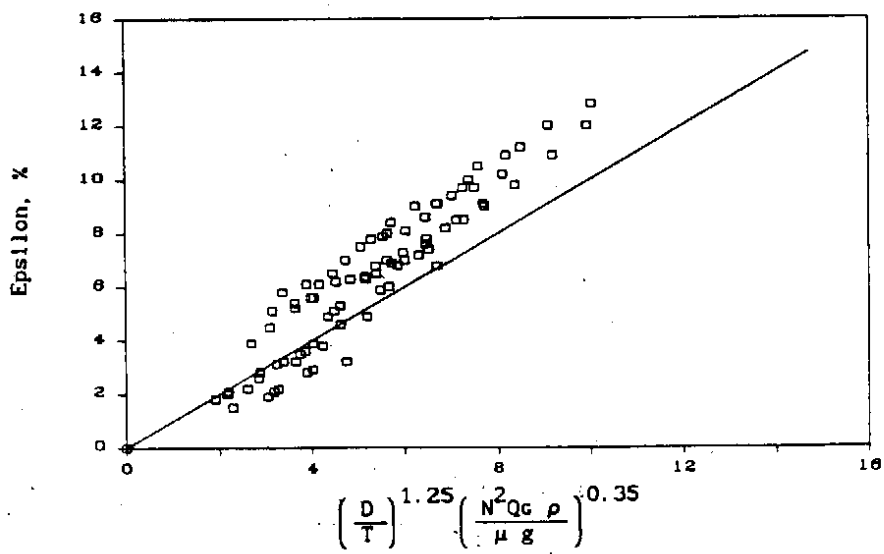


Fig. 4. Data from Greaves and Barigou
T = 1 m; D = 0.25, 0.33 & 0.5 m

Although the sensitivity to this exponent is low, and the trends are consistent, the data of Greaves and Barigou (1987), obtained with various impellers in a 1m diameter tank, do not fit well with the scale independent generalised result. There is no reason to doubt the accuracy of their extensive experimental results which were obtained with rapidly coalescing distilled water.

All their data were obtained in a 1 m diameter flat bottomed tank with a base to impeller clearance of 0.25T. For all three impeller diameter ratios used, their results show good initial agreement with the proposed correlation which is shown in Fig. 4. It is only as gas fraction increases that the differences become large, amounting to an additional 3% or so voidage. Although the largest impeller (D/T ratio = 0.50) seems to show a somewhat greater divergence than the other two sizes, the performance is reasonably consistent. Statistical analysis of their results, and imposing the requirement of an independence of scale, gives an exponent of almost exactly 2 on (D/T). However this worsens the correlation using (N^2Q) which has been justified by the large mass of other data, leaving 1.25 as the best compromise.

The resulting correlation then brings together

$$\varepsilon_G = \text{Const} \cdot \left(\frac{N^2 Q_G}{g\mu}\right)^{0.35} (D/T)^{1.25} \qquad (2)$$

In dimensionless terms this can be written as

$$\varepsilon_G = 0.85\,(Re \cdot Fr \cdot Fl_G)^{0.35}\,(D/T)^{1.25} \qquad (3)$$

MULTIPLE IMPELLER GEOMETRIES.

The recycled overflow technique was has been used to measure the holdup in a vessel with triple Rushton turbines (Zeef (1985) and Hogervorst (1987)). Fig. 5. shows their results plotted against the same dimensionless grouping.

It is evident that it is not possible to fit a straight line through the origin to these multiple impeller data, with lower measured gas fractions at low gas rates than a linear correlation would lead us to expect. A line fitted through the origin to the higher gas fraction data has a significantly greater slope (1.25) than the standard tank results (0.85).

The flow regimes in multiple impeller vessels are significantly different from those in standard vessels. The flow map reflects the reduced gas loading on the upper impellers. The data which deviate at low gas fractions correspond to regimes in which the upper impellers are in the vortex cavity regime.

Again additional industrial results are available to confirm and extend the suggested correlation. Fig. 6 includes data from BHRg and ICI. One remarkable point about these data is that the configurations used were not all the same: some equipment consisted of triple disc turbines while in other experiments combinations of a dispersing disc turbine with profiled impellers above.

The full set of data available is shown in Fig. 7.

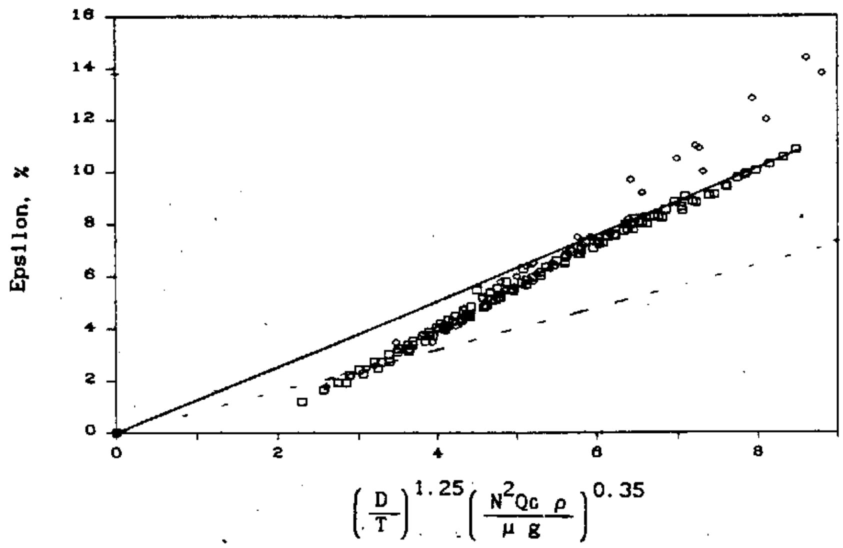
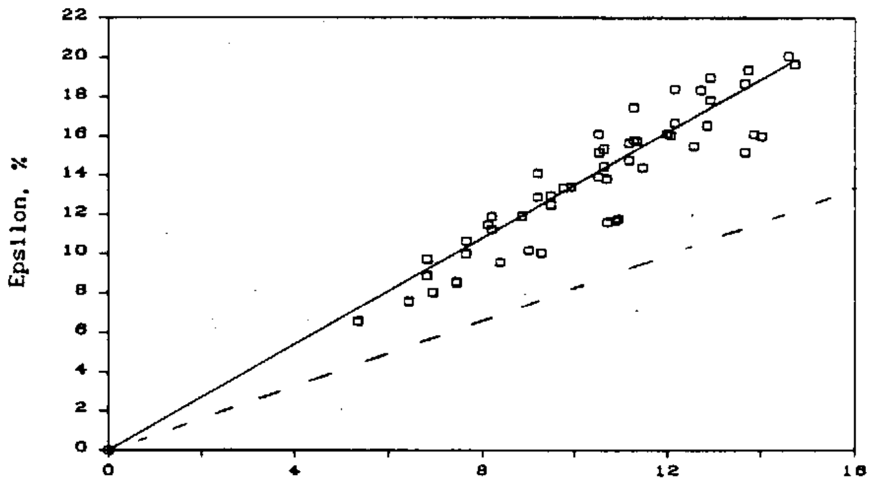


Fig. 5. Multiple Impeller Gas Void Fraction, Delft Data.
Triple Rushtons, T = 0.64 m D = 0.25 m.
◇ Zeef □ Hogervorst



$\left(\frac{D}{T}\right)^{1.25}\left(\frac{N^2 Q_G\ \rho}{\mu\ g}\right)^{0.35}$

Fig. 6. Industrial Void Fraction Data,
Various Combinations of Triple Impellers, T = 0.64 m and 0.95 m.

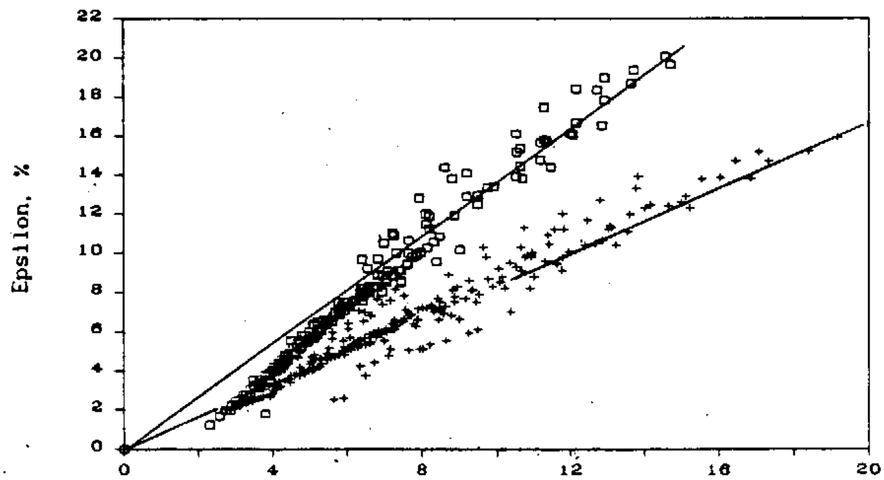
Implications of higher voidage in Multiple Impeller vessels.

The ratio between the constants of the correlating equations (0.85 versus 1.25) implies that there is a very considerable deference in the local holdup above and below the impeller plane, at the very least in a single impeller vessel and possibly also above and below upper impellers, though this is less likely in view of the very different way in which gas reaches the upper impellers.

If we assume the simplest possible configuration of uniform gas distribution above the plane of the lowest impeller with three pairs of compartments of equal volume, centered on the impellers, a simple volume balance implies that the gas fraction below the lowest impeller is only about 15% of that in the upper reaches of the tank.

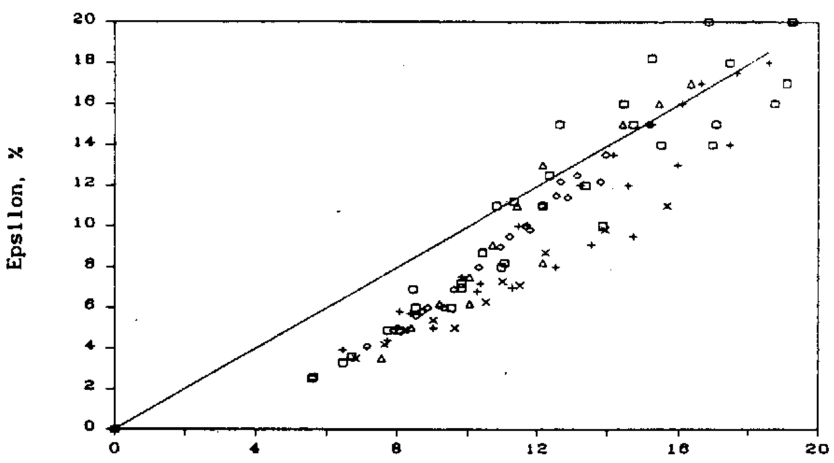
PHYSICAL PROPERTIES.

Although the analysis presented in favour of (Re•Fr•Flc), may appear convincing, important questions remain. The dependence on $N^2Q\bullet\left(\frac{\rho_L}{g\ \bullet\ \mu_L}\right)$ implies a mutual relationship



$\left(\frac{D}{T}\right)^{1.25}\left(\frac{N^2 Q_G\ \rho}{\mu\ g}\right)^{0.35}$

Fig. 7. All data results for void fraction.
+ Single impellers, standard tanks. □ Triple impellers



$$\left(\frac{D}{T}\right)^{1.25}\left(\frac{N^2Qc}{\sigma}\right)^{0.35}\left(\frac{\rho^2\mu}{g^2}\right)^{\frac{1}{3}}$$

Fig. 8. Data of Hassan and Robinson in water, glycol and glycerol
□ Water, T = 0.15 m + Water, T = 0.29 m ◆ 8% Glycerol, T = 0.29 m
Δ 20% Glycerol, T = 0.29 m x 40 % Glycerol, T = 0.29 m

between density and viscosity which has not been justified by experiment. Useful guidance is provided by the data reported by Hassan and Robinson, (1977). Their paper studied coalescence affecting properties, notably the surface tension of the liquid phase. They found that the (dimensional) group (N^2Q/σ) was effective in correlating their results. If the surface tension is to be included in the correlation then dimensional considerations lead us to look for a correlation involving the terms :

$$\varepsilon_G \propto \left(\frac{N^2Q}{\sigma^x}\right)\left(\rho^{(1-x/3)}\mu^{(4x/3-1)}g^{(x/3-1)}\right) \quad (4)$$

With a value on x of 1.0 this reduces to:

$$\varepsilon_G \propto \left(\frac{N^2Q}{\sigma}\right)\cdot\left(\frac{\rho^2\mu}{g^2}\right)^{\frac{1}{3}} \quad (5)$$

which implies quite different interdependence of the variables in the last group to that which follws from Equ. (3). Unfortunately the data of Hassan and Robinson, while following the same general trends, diverge significantly from the other data presented in this paper. Many of their results were obtained in a very small vessel and involved extreme void fractions - in excess of 20 %. However using the other physical properties in the examination of their data does improve the consistency of their results, Fig. 8.

Even though the individual experimental data do not fit the proposed correlation quantitatively, the trends are unambiguous and the differences amount to perhaps 30% in the appropriate constants.

Henzler and Kauling (1985) included surface tension and other fluid properties in their consideration of mass transfer performance in viscous liquids. The real role of surface tension is as a coalescence rate parameter, and in that light the dynamic surface tension and surface elasticity are certainly as important as the static values used by Hassan and Robinson.

Given the sensitivity of the coalescence mechanism to small degrees of contamination it is worthwhile considering the present approach as being as likely as any other to offer the key to the useful description of the physical process.

MASS TRANSFER.

The correlation of mass transfer performance is much more complex than that of gas fraction. Apart from anything else, k_La is not dimensionless and k_La/N has an unfortunate dependence on the particular operating conditions. It is suggested that the group $k_La(D/g)^{0.5}$, derived by multiplying k_La/N by the square root of the Froude Number, would be useful, since this will be characteristic of the equipment itself rather than its mode of operation.
Using various mass transfer results from Delft, BHRg and ICI in vessels in the range 0.44 m to 2.4 m diameter, and basing the mass transfer calculation on well mixed liquid with a plug flow for the gas phase, a usable correlation emerges, Fig. 9. The line, which is within an rms deviation of 8% of the data points, has the equation:

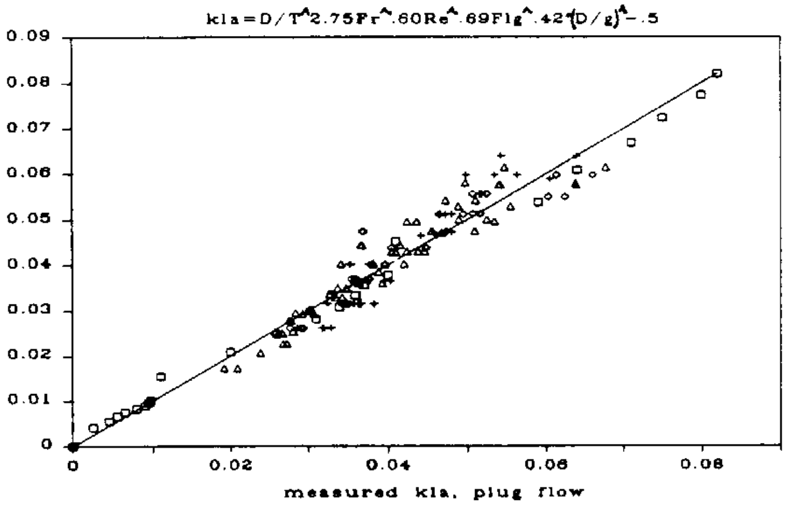


Fig. 9. Mass transfer performance in dimensionless terms.
□ Delft, (hollow blade) T = 0.44 m △ BHRA (Rushton) T = 0.6
\+ BHRG (imp type 1) T = 2.6 ◇ BHRA (imp. type 2) T = 2.6

$$k_La = 1.25 \cdot 10^{-4} (\frac{D}{T})^{2.8} Fr^{0.6} Re^{0.7} Flc^{0.45} (\frac{D}{g})^{-0.5} \quad (6)$$

Although the exponents on the correlating equation are not very sensitive, it has not been possible to reduce this grouping to a diameter independent set of variables as described the void fraction. This data was all derived from physical oxygen mass transfer experiments in clean air-water systems. It is appropriate to repeat the warning associated with the void fraction results above. Almost certainly surface tension (or some physical property related to it), ought to be included. If this is done then the interdependence of density and viscosity will, on the physical grounds of dimensionality, be different from that implied in equation.

CONCLUSIONS.

This paper has considered the possibilities of basing agitated vessel performance correlations on dimensionless groups. It has pointed out that in common with many multi variable optimizations the sensitivity to the parameter exponents is rather low, and that simplified rules are often worthwhile. For the correlation of gas holdup in agitated vessels operating with aqueous aerated mixtures the equation

$$\varepsilon_G = 0.85\ (Re \cdot Fr \cdot Flc)^{0.35}\ (D/T)^{1.25}$$

is sufficiently accurate for many purposes. The constant for multiple (triple) impeller equipment is 1.25. This leads us to the conclusion that the relative gas retention below the lowest impeller is probably not more than 20% of that in the upper regions.

Acknowledgements.

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Notation

D	Impeller diameter	m
g	Gravitational acceleration	$m\ s^{-2}$
$k_L a$	Mass transfer factor	s^{-1}
N	Impeller Speed	s^{-1}
Q_G	Volumetric gas flow rate	$m^3\ s^{-1}$
T	Tank diameter	m
ε_G	Gas void fraction, percent	-
μ	Liquid viscosity	Pa s
ρ	Liquid density	$kg\ m^{-3}$
σ	Surface tension	$N\ m^{-1}$

Dimensionless Groups

Re $\left(\frac{N D^2 \rho}{\mu}\right)$ Fr $\left(\frac{N^2 D}{g}\right)$

Flc $\left(\frac{Q_G}{N D^3}\right)$ We $\left(\frac{\sigma}{\rho N^2 D^3}\right)$