x = axial coordinate in adsorber in the direction of liquid flow, m

L = length of carbon bed, m

Greek Letters

ϵ_B = void fraction in the bed

ϵ_L = dynamic liquid holdup, (volume of liquid)/(volume of empty tube)

ϵ_p = particle porosity

μ = liquid viscosity, kg/(m)(s)

ρ = liquid density, kg/m^3

τ = retention time in the bed, $\epsilon_L l/u_L$, s

τ_t = total average retention time from injection point to detector, s

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Mixing of Single and Two Phase Systems: Power Consumption of Impellers

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The mechanical power consumed in agitating aerated and nonaerated aqueous phases was measured and correlated using the data of previous investigators. The correlation is for a flat bladed turbine operating within the turbulent regime in fully baffled vessels. The accuracy of the predictions and the range of geometrical variations and equipment sizes support the use of the correlation for scaleup operations.

SCOPE

The power consumption of impellers used for agitating gas-liquid systems is a widely used parameter for determining gas holdup, interfacial area and mass transfer rates in baffled vessels. The use of this parameter leads to complications since to date, there appears to be no general working equations for the prediction of power consumption. Calderbank's (1958) equations

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for predicting gas holdup and gas-liquid interfacial area require prior knowledge of the agitator power consumption per unit volume agitated. Graphical predictions of the ratio of gassed to ungassed power consumption as a function of the sparging rate may not be applicable when geometrical variations from Calderbank's systems or scaleup are involved.

Rushton's (1968) equations for predicting gas holdup in baffled vessels also include the power per unit volume consumed by the impeller. He recommends the use of a pseudo-density chart from the data of Foust et al. (1944) in which power for a gassed system can be predicted from single phase power consumption information. This suggestion appears invalid, however, since a curved blade impeller was used in Foust's work and a flat blade impeller in Rushton's work. Handbooks and other literature often suggest Michel and Millers' (1962) prediction of power consumption. However, the variation of the constant of proportionality with small geometry changes, as well as the low gas rates studied, indicates as the authors themselves state, that care should be employed when applying their equation to scaleup operations.

The empirical correlation obtained in Hassan and Robinson's (1977) work, although covering a wider range of gas rates than Michel and Miller, are also limited to the laboratory scale geometries investigated. Because of all of these limitations, there is still a need for a general working equation for the prediction of power consumption by impellers used in gas-liquid systems.

The lack of uniformity in test conditions and parameters used, as well as the discrepancies found between investigators, indicates that it may be appropriate to modify conventional dimensionless groupings derived for single phase mixing with an additional parameter to include the gas velocity term. Here we develop a general working equation for disc-type impellers with six flat blades agitating a single phase in a baffled tank. The equation is extended to gas-liquid systems.

CONCLUSIONS AND SIGNIFICANCE

The effect of geometric variations on the power number for disc-type impellers with six flat blades used in agitating liquids in the fully turbulent regime was investigated. Specifically the ratios of tank to impeller diameter (T/D) and impeller height to impeller diameter (C/D) were studied for fully baffled ($W/T = 0.1$) vessels. The result was a constant power number of 5.17 representing the data for $C/D > 1.1$. For $C/D < 1.1$, N_p varied with $(C/D)^{0.29}$.

The single phase correlation was extended to aerated systems using a liquid and gas Froude number which incorporates pressure and gas velocity effects, respectively, on a two phase system. A general working equation was obtained from the data of Bimbinet (1959) and Hassan and Robinson (1977) with additional data obtained to support the scaleup applications of the results. The general simplified equation obtained was

$$\left(\frac{P_T g_c}{N^3 D^5 \rho_L}\right) = 0.75\left(\frac{CTg^2}{(v_G/\phi_G)^2 N^2 D^2}\right)^{0.25}$$

The accuracy of the prediction (±20%) over a wide range of geometry, speeds and gas rates indicate that a prediction of power consumption for scaleup can be made from this equation. Application to other gas-liquid systems appears reasonable if the effect of surface tension is taken into consideration.

INTRODUCTION

Previous studies, due to the complexity of mixing gas-liquid systems, generally use dimensional analysis for developing correlations of power consumption data. The major parameters used in the correlations are the aeration number (Q/ND^3), gassed system power to ungassed system power (P_G/P_O) ratio, apparent or pseudo-density (ρ_D), gas holdup (ϕ_g), impeller Weber number ($N^2D^3\rho_L/\sigma$) and several dimensionless geometry factors. Conventional dimensionless numbers derived for single phase mixing will be used here with an additional parameter included in the correlation to extend the single phase correlation to aerated systems.

EQUIPMENT

The work was begun with a 0.762 m diameter tank equipped with four 0.0762 m vertical baffles spaced at 90° intervals. A 2237 W motor was used to drive a 0.1524 m diameter, disc-turbine impeller with six flat blades attached to a 0.0159 m diameter shaft equipped with a digital-readout torque meter (accuracy ± 0.35Nm). Speed changes were accomplished with changes of pulley sizes on motor and shaft.

Power was measured for agitating water alone with the liquid level held constant at 0.5 m (0.232 m^3 volume) and the impeller located at 0.0762 or at 0.178 m from the tank bottom ($C/D = 0.5$ and 1.167, resp.).

Following this, water alone was agitated in a 0.287 m diameter tank with a 0.0906 m diameter impeller, $W/T = 0.1$, $C/D = 1.5$.

Power measurements for gas-liquid mixtures were studied only in the 0.762 m tank under the same conditions as the water study. Gas rates were measured using calibrated rotameters and pressure gages.

DISCUSSION

Single Phase System

The Buckingham *pi* theorem gives the following general dimensionless equation for the relationship of the variables:

$$f(D^2N\rho/\mu, DN^2/g, P_Gg_c/\rho N^3D^5, D/T, D/C, D/z, D/W, D/r, D/w, D/L, n_1/n_2) = 0 \quad (1)$$

The last eight dimensionless groups representing the system geometry are often neglected by investigators leading many readers to the mistaken belief that the power number-Reynolds number relationships which were developed for specific standard geometries, and are found in handbooks or other literature, are universally applicable to all geometries. The number of curves needed to represent all geometries leads to the need for the inclusion of these geometrical groups into a general equation.

For disc impellers, of the six flat blade type, the last four groups are constants. Generally tanks are filled to a liquid depth equivalent to the tank diameter and the dimensionless group D/Z, in this case, is represented by the group D/T and can be dropped from our equation. Standard size baffling of $1/12 \leq W/T \leq 1/10$ is usually encountered in mixing and baffling effects should be small over this range and may be neglected. This leaves as a working equation;

$$P_Gg_c/\rho N^3D^5 = \text{const.}\left(\frac{D^2N\rho}{\mu}\right)^a\left(\frac{DN^2}{g}\right)^b (D/T)^c(D/c)^d \quad (2)$$

The group on the left side of the equation is a drag coefficient grouping of terms known as the power number (N_p). The first group on the right side of the equation is the impeller Reynolds number representing the ratio of inertial to viscous forces. The second group is the impeller Froude number representing the ratio of inertial to gravitational forces. All three groups can be derived from the Navier-Stokes equation.

For single phase, fully baffled systems, operating in the turbulent regime, the inertial forces dominate and the Reynolds and Froude numbers may be neglected. In this case, the equation now states that for geometrically similar systems, operating in the turbulent regime, the power number becomes a constant. To date, however, the geometry effects are not totally clear in the literature and will be discussed below.

Effect of Impeller Position and Tank Size on N_p

The impeller position from the free surface in a baffled vessel should have no effect on N_p as long as vortexing or air entrainment to the impeller depth are not promoted by too high a position. Clark and Vermeulen (1964) report an equation for predicting when air entrainment to the impeller depth becomes important for power consumption. Generally, this condition can be avoided. The clearance beneath the impeller, however, has been found to be of significant importance. Bates et al (1963) studied C/D effects on N_p in the turbulent regime using six blade flat disc-type impellers and correlated his data by log N_p vs. log C/D. He indicates that there is a reduction in power with a decrease in clearance over a range of C/D from approximately 0.17 to 1.1. Above $C/D = 1.1$, there is no effect and N_p remains constant. The slope of a plot of his data for the region $0.5 < C/D < 1.0$ is about 0.18. Figure 1 represents power correlations for agitating water in two separate fully baffled batch systems. The data are represented by a plot of $t2\pi\, g_c/D^5\rho_L$ vs. N^2. From the torque to power relationship $[2\pi Nt = P_o = N_p N^3 D^5 \rho_L/g_c$, the average N_p can be found from the slopes in Figure 1. The data for both the 0.287 and 0.762 meter diameter tanks for $C/D \geq 1.17$ and D/T equal to 0.33 and 0.2 respectively can be plotted by one line of slope $= N_p = 5.17$. Bates found a small variation in N_p for $W/T = 0.1$ over a range of $0.25 \leq D/T \leq 0.5$ for open style six blade flat blade turbines. The variation became more pronounced as the extent of baffling increased. The discrepancy here cannot presently be explained, however, the effect of D/T is small for the baffling employed.

The effect of C/D was investigated in the 0.762 meter tank only. Assuming a power relationship can be applied, which Bate's graphs appear to indicate, $N_{p1}/N_{p2} = [(C/D)_1/(C/D)_2]^x$. At $C/D = 0.5$, $N_p = 4.04$ and at $C/D = 1.17$, $N_p = 5.17$ from which we obtain $x = 0.29$. The effect of C/D on N_p is more pronounced in our system as compared to Bate's findings noted above. Our resulting equations for single phase mixing are:

$$N_p = 5.17\,(C/D)^{0.29} \qquad \text{for } C/D < 1.1 \tag{3}$$

$$N_p = 5.17 \qquad \text{for } C/D > 1.1 \tag{4}$$

Two Phase Systems

There has been limited success in correlating power consumption in a gassed system (P_G) as a function of P_o and the dimensionless group Q/ND^3. Although Q/ND^3 can be derived from dimensional analysis, to date no one has placed any physical significance to this dimensionless grouping and it may be more appropriate to use the conventional Froude number.

Importance of the Froude Number in Mixing

For single phase mixing in unbaffled vessels, it is known that the power number is dependent upon the Froude number, the magnitude of importance being dependent upon the Reynolds number. The greater the Reynolds number, the greater the vortexing resulting in a greater dependence of the flow pattern on the gravitational forces acting on the fluid. For the fully turbulent regime, in a baffled system, the inertial forces of the fluid acting vertically in our system tend to dominate over the gravitational forces. The flow pattern for a centrally placed impeller in a baffled vessel with limited vortexing should become symmetrical as these inertial forces begin to dominate.

As stated earlier the Froude number generally can be neglected in our correlation. It could be, however, that for a gassed system, the Froude number may not be insignificant since the gravitational forces tend to promote a vertical separation of liquid and gas. Gravity acting on the liquid leaving an impeller traveling radially to the tank wall in a gassed system will tend to promote the downward flow of the liquid producing an unsymmetrical flow

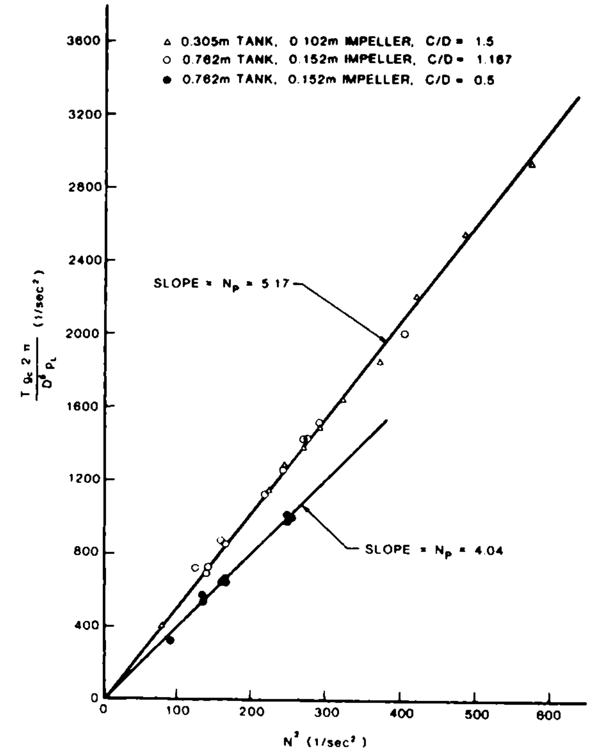

Fig. 1. Variation of power grouping with impeller speed for agitated water systems with various impeller heights.

pattern as opposed to that of a single phase system. The degree of assymmetry of the flow pattern should be dependent upon the inertia of the fluid leaving the impeller, thereby giving rise to the need for a liquid Froude number in our correlation. In addition, the liquid Froude number, as will be discussed below, will also represent the pressure effects in our mixing system.

An increase in the gravitational force should also promote the upward flow of the gas phase leading to a reduction in the gas holdup. An increase in the gas feed rate (vertical inertial force) should increase the gas holdup within the system. It appears, therefore, that a gas Froude number should also be incorporated into our correlation.

The two Froude numbers discussed above are u_L^2/Dg and u_G^2/Dg respectively. In order for these Froude numbers to be appropriate, however, the velocity terms used should be the actual liquid and gas velocities at the impeller. Physically these velocities are not known and reference velocities of ND and V_G/ϕ_G are used in place of the actual liquid and gas velocities, respectively, in our correlation.

Power Correlation

A correlation of total power input was carried out for the two phase system. The work done by the expansion of rising gas bubbles is included in the power term. A least square regression analysis of the power number $P_T g_c/N^3D^5\,\rho_L$ as a function of the two Froude numbers and the geometry parameters C/D and T/D was done on the air-water data of Bimbinet (1959) and Hassan and Robinson (1977). The data covered tank sizes of 0.151, 0.29 and 0.454 meters for D/T ranging from 0.25 to 0.45 and C/D ranging from 0.75 to 1.5. The result was

$$P_T g_c/N^3D^5\rho_L \propto \left(\frac{g}{N^2D}\right)^{0.230} \times \left(\frac{Dg}{(v_Q/\phi_G)^2}\right)^{0.235} (C/D)^{0.270} (T/D)^{0.246} \tag{5}$$

The closeness of the exponents above allows a simplification of the general equation into the one dimensionless group $(TCg^2/(V_G/\phi_G)^2\,N^2D^2)$. The new general equation obtained was

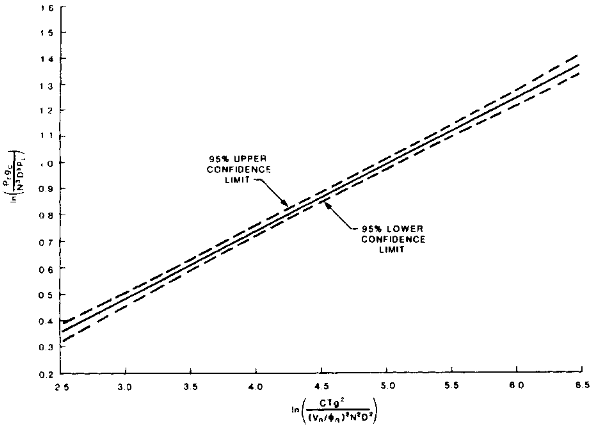

Fig. 2. Variation of aerated power number with geometry-Froude number grouping

$$P_T g_c / N^3 D^5 \rho_L = 0.75 \left(\frac{g^2 TC}{(V_G/\phi_G)^2 N^2 D^2}\right)^{0.25} \qquad (6)$$

In the development of Eq. 6, 247 data points were used from the two sources mentioned above. The average absolute percentage error was 8% with only 8% of the data deviating by more than 20% error. The largest errors occurred in systems with low gas rates and/or impeller speeds. Small errors in the measurement of the gas holdup would result in a large percentage error in these regions since the gas holdup is small under these conditions. Figure 2 is a plot of the 95% confidence level for Eq. 6.

Our data was added to the correlation using Calderbank's equation (1958) for predicting gas holdup. Approximately 90% of the data fell within ±20% error.

The conclusion that P_G/P_O is independent of N has been found by some investigators, most notably by Hassan and Robinson (1977) in their Figure 1. Close inspection of Eq. 6 does not totally disagree with Hassan and Robinson's finding. The gas holdup for their air-water system varied with impeller speed raised to the 1.14. Substitution of this value for the gas holdup into equation 6 results in P_G/P_O as a function of N raised to the 0.07. Thus P_G/P_O is relatively independent of N if one substitutes for the gas holdup in Eq. 6.

The effect of C/D was investigated in order to check the above findings. The 0.762 meter tank agitated with a 0.151 meter impeller for C/D equal to 0.5 and 1.17 and for superficial gas rates from 8.28×10^{-4} to 1.84×10^{-2} m/s resulted in $P_G \propto (C/D)^{0.29}$, the same findings as that of the single phase study. P_G/P_O, then, is independent of C/D.

Pressure Effects in Mixing

Visual observations of gas distribution at low gas rates could be made in mixing vessels. Gas would accumulate at the impeller tip just above the radial line through the impeller position and could be held there for some time after sparging was terminated. An explanation for this phenomenon could be that the pressure at the impeller tip due to the high liquid velocity can be less than the pressure of the fluid above the impeller; the gas will seek the lower pressure area and remain there indefinitely. Calderbank (1958) found that gas holdup in the vertical direction is greatest just above the impeller. From the observation above and Calderbank's findings, it appears that the gas distribution was not uniform in either the radial or vertical directions and pressure was probably the important parameter which determines gas distribution.

In practice, the pressure distribution in a mixing vessel is not known; however, from Bernoulli's equation, we can relate pressure to fluid velocity as $\Delta P/\rho = u_L^2/2g_c$. Using the reference velocity of ND, our pressure relation becomes

$$\Delta P \propto \left(\frac{\rho N^2 D^2}{2 g_c}\right)$$

and assuming a linear pressure drop in the radial direction $P/T \propto ND^2/2Tg_c$.

The liquid Froude number, as mentioned earlier, already incorporates the liquid velocity needed to relate pressure drop to power consumption. The fact that the D/T geometry factor has the same exponent relation to power consumption as the liquid Froude number, then, is not surprising when viewing from the effect of pressure distribution in the mixing tank.

Application to Other Systems

Clark and Vermeulen (1963) studied power consumption of four blade flat paddles and turbines of various sizes and impeller to tank diameter ratios used for mixing gas-liquid dispersions. They found that $P_G/P_O = f(\phi_g, N_{we}^{1/4}, G_V^{1/2})$ for water and organic liquids of different density, surface tension and viscosity. Hassan and Robinson (1977) found $P_G/P_O \propto (\rho_L/\rho_D) N_A^{-0.38} N_{we}^{-0.25}$ for six blade flat blade impellers and paddles for a variety of aqueous and organic liquids of variable density, surface tension and viscosity. The exponent on the N_{we} increased to -0.22 for four blade paddles. Application of our general equation to other liquid systems therefore appears reasonable with the inclusion of σ_L/σ_W in our dimensionless group which results in P_T as a function of σ to the 0.25. Additional work is needed, however, to confirm this assumption.

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NOTATION

C = impeller height off the tank bottom, m
d = impeller diameter, m
g = local gravitational acceleration, m/s^2
g_c = gravitational constant, kg-m/kg · s^2
G_v = geometrical group, $D^2 w/T^2 Z$, dimensionless
L = impeller blade length, m
n = number of impeller blades
N = impeller rotational speed, s^{-1}
N_a = sparging rate, Q/ND^3, dimensionless
N_p = power number, $P_o g_c/\rho N^3 D^5$, dimensionless
N_{we} = impeller Weber number, $N^2 D^3 \rho_L/\sigma$, dimensionless
p = blade pitch
P_G = mechanical agitation power in gas-liquid dispersion, W
P_o = mechanical agitation power in ungassed liquid, W
P_T = total power input in gas-liquid dispersion, W
ΔP = pressure difference, N/m^2
Q = volumetric gas rate, m^3/s
t = mechanical agitation torque, W-s
T = tank diameter, m
u = actual velocity, m/s
v = superficial velocity, m/s
w = impeller blade width, m
W = baffle width, m
z = liquid height above the tank bottom, m

Greek Letters

σ = air-liquid surface tension, N/m
ϕ = volume fraction holdup, dimensionless
ρ = mass density, kg/m^3
μ = liquid phase viscosity, N/s/m^2

Subscripts

D = properties of gas-liquid dispersion
G = property of the gas

L = property of the liquid
w = property of water
1 = condition of system 1
2 = condition of system 2

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Kinetics of Coal Drying-Decomposition

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A series of seven coals of different ranks and from various locations were heated in an inert atmosphere under relatively mild conditions to measure the rates of gas evaporation. Samples crushed to various fractions in the particle size range −6 +50 U.S. mesh were studied at temperatures in the range of 150 to 300°C. The results show that: (1) most water is released at 100°C, evidently an evaporation process, (2) CO_2 is evolved at 150°C and above, and (3) CO is evolved at 250°C and above. An interpretive model was developed to fit the CO_2 production rate and kinetic constants were obtained. Gas evolution rates are independent of particle size for the most porous coals, but vary among coals, depending on both chemical composition and physical structure. A 15% loss in heating value was incurred during drying and subsequent oxidation when the pretreatment temperature was increased from 150 to 225°C.

SCOPE

The rates at which gases are evolved during coal drying under an inert atmosphere were measured in a fixed bed reactor. Seven coals from different sources were crushed to the particle size range −6+16 U. S. mesh and dried under isothermal conditions at 150°C. Four of the coals were further tested for particle size and temperature effects. The effects of drying severity on subsequent oxidation rates were also measured. The dried coals were analyzed for changes in chemical composition and heating value. A model was developed and kinetic parameters were estimated for the carbonic gas evolution rate.

CONCLUSIONS AND SIGNIFICANCE

As a result of experimental drying tests on seven coals from different sources, it has been found that (1) water is released mainly as the result of an evaporative process at 100°C, (2) CO_2 is produced under an inert atmosphere at temperatures as low as 150°C, and (3) CO is produced at 250°C and higher temperatures. The carbonic gases are evidently produced as the result of thermal decomposition of oxy-functional groups on the coal surfaces, since CO_2 production rate increases with the original oxygen content of the coal. A drying model that assumes zero and first order decompositions of oxy-functional groups fits the data, providing activation energies as low as 13 to 22 MJ/kmol for CO_2 production. Gas evolution rates are independent of particle size for a given coal, indicating that the relatively low activation energies are not the result of diffusional limitations. The oxidation rate of the lignite coal tested was found to be quite sensitive to prior drying temperature: the rate at 150°C doubled when the coal was dried at 300°C instead of 150°C.

The oxidative pretreatment of coal to reduce or eliminate its swelling and caking propensity consists of two steps: (1) coal drying while the coal is heated to pretreatment temperatures and (2) coal oxidation. These steps, usually performed simultaneously, include both rapid heating and addition of gaseous oxygen. The first step is often taken for granted by assuming that the thermal drying involves only the removal of water; thus the two steps are studied as one. Since for design and optimization of an oxidative pretreatment process knowledge of low temperature drying kinetics is useful, this study was undertaken to isolate the drying step as separate from the oxidative pretreatment and to examine possible chemical as well as physical changes which may occur.

Coals may contain large quantities of water; lignites often contain water in excess of 20 weight percent. Prior to oxidizing coals, water is typically removed in the laboratory by either (1) heating

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