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SOME FEATURES OF A NOVEL GAS DISPERSION IMPELLER IN A DUAL-IMPELLER CONFIGURATION

D. PINELLI[1], A. BAKKER[2,*], K. J. MYERS[3], M. F. REEDER[2], J. FASANO[2] and F. MAGELLI[1]

[1]*DICMA, Department of Chemical, Mining and Environmental Engineering, University of Bologna, Bologna, Italy*
[2]*Chemineer Inc., Dayton, USA*
[3]*Department of Chemical and Materials Engineering, University of Dayton, Dayton, USA*

The performance of the concave blade BT-6 impeller was evaluated in a dual-impeller agitated tank ($T = 0.48$ m, $H/T = 2$). Power draw was measured and almost constant P_g/P_u ratio with aeration was found in agreement with what had been reported previously. Gas hold-up exhibited the same dependence on power per unit volume and superficial velocity as shown by other types of impellers. The same holds true for the volumetric mass transfer coefficients. Mixing time was measured at several vertical positions in the tank, after a pulse of an electrolytic tracer or a dye was introduced. No significant compartmentalization was apparent. Mixing time at a given specific power input under ungassed and gassed conditions is in between that of standard radial turbines and hydrofoil axial impellers, very close to the latter. Mixing time dependence on power consumption per unit volume exhibits $-1/3$ law, i.e. the same as reported in the literature for other impeller types. The experimental curves from which mixing time was determined were also analysed in terms of the axial dispersion model, that proved satisfactory to interpret the behaviour of dual BT-6 impellers under both ungassed and gassed conditions.

Keywords: stirred tanks; gas–liquid systems; concave blade impellers; dual impellers; mixing time; fluid dynamic model; gas hold-up; mass transfer coefficients.

INTRODUCTION

Rotating impellers are widely used for dispersing the gas in chemical reactors and fermenters (Tatterson, 1991). Although the Rushton turbine has been used for decades for this purpose, alternative designs meant to enhance and optimize gas–liquid contact in tanks have been developed in recent years. They include radial disk turbines with modified blades (Van't Riet *et al.*, 1976; Warmoeskerken and Smith, 1989; Bakker *et al.*, 1994; Orvalho *et al.*, 2000) and high solidity ratio hydrofoil impellers pumping down (Oldshue *et al.*, 1988; Pandit *et al.*, 1989; McFarlane and Nienow, 1996; Myers *et al.*, 1997a) or up (Mishra *et al.*, 1998). A comparative analysis of the fluid dynamic performance of these systems was provided by Nienow (1990, 1996). A new type of disk turbine for gas dispersion was introduced most recently that is characterized by concave, vertically asymmetric blades (Bakker, 1998; Myers *et al.*, 1999). In large-scale vessels, multiple impellers are often used with either impellers of the same type or a combination of novel and traditional ones (Smith *et al.*, 1987; Myers *et al.*, 1994; Pinelli *et al.*, 1994; Baudou *et al.*, 1994; 1997; Manikowski *et al.*, 1994; Bouaifi *et al.*, 1997; John *et al.*, 1997; Myers *et al.*, 1997b; Whitton *et al.*, 1997).

Each system is usually characterized in terms of overall parameters, namely power draw, flooding point, gas hold-up, mixing time and mass transfer coefficient. The possibility of modelling the fluid dynamics would be most useful for design purposes as well as for process evaluation. However, this is a really formidable task due to the complex nature of the flow pattern and the turbulence field produced by a rotating impeller, the influence of impeller details on the overall and local fluid dynamics, and the additional complications imposed by the gas phase in gas–liquid mixing. Although preliminary attempts to apply computational fluid dynamics to the description of mixing for gas-liquid systems have been made (Whitton *et al.*, 1997; Noorman *et al.*, 1993; Djebbar *et al.*, 1996; Jenne and Reuss, 1997; Lane *et al.*, 2000; Ranade *et al.*, 2001), fully predictive reliable means is still unavailable. Therefore, more traditional macromixing modelling seems to be a suitable alternative for describing the fluid dynamic behaviour, at least for a preliminary analysis. A few examples of this approach can be found in the literature (Fajner *et al.*, 1982; Nocentini *et al.*, 1988; Pinelli and Magelli, 2000).

In this paper the behaviour of the asymmetric blade disk turbine (Chemineer BT-6) in the dual configuration is studied. Gas hold-up, mixing time, liquid phase modelling and mass transfer coefficients are considered. Comparison with other impellers is also provided.

*Present address: Fluent Inc., Lebanon, NH, USA.

EXPERIMENTAL

Equipment and Experimental Conditions

The main body of the experiments was conducted in a cylindrical tank ($T = 0.48$ m diameter, $V = 173$ l working volume) having an aspect ratio $H/T = 2$. The tank had a flat bottom and four vertical $T/10$ baffles. Agitation was provided with two identical BT-6 turbines of standard design (swept diameter $D/T = 0.45$) mounted on the same shaft. (BT-6 is the shorthand notation for the vertically asymmetric blade turbine with the upper section of the blades longer than the lower section, see Figure 1; it is the development of the symmetric concave blade turbine—the so-called Smith turbine.) The lowest turbine was placed at $T/2$ above the tank base, while impeller spacing was equal to tank diameter. In a limited number of experiments the liquid level was varied: it was halved ($H/T = 1$) so that only one turbine was operating for power measurement or increased to $H/T = 2.4$ for selected K_La measurements. The main features of the equipment for the reference $H/T = 2$ configuration are shown in Figure 2.

Additional experiments were performed with triple radial Rushton turbines of standard geometry (either $D/T = 0.32$ or 0.40) and dual and triple hydrofoil impellers of high solidity ratio (HSR-HF, $D/T = 0.40$). For this last case, one of the commercially available, large-blade impellers was chosen, characterized by a solidity ratio equal to 0.85. With dual impellers the vessel aspect ratio was $H/T = 2$ and with triple impellers $H/T = 3$.

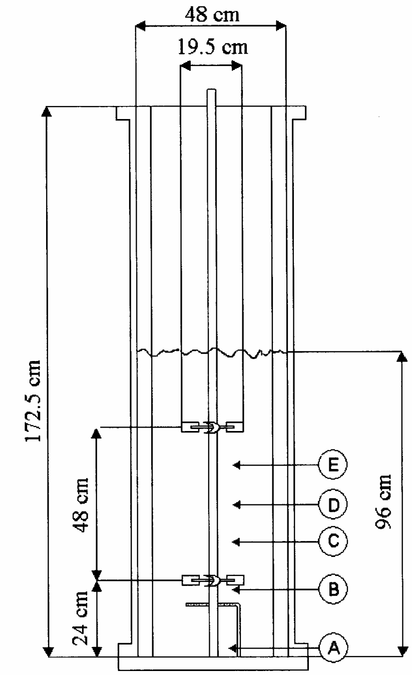

Figure 2. The experimental tank and set-up. (A–E) Liquid measurement/sampling elevations ($Z = 0.04$, 0.22, 0.37, 0.50, 0.62, respectively).

The experiments were performed in semibatch conditions at room temperature and atmospheric pressure. The liquid batch was demineralized water (coalescent system); filtered air was fed to the system through a ring sparger ($D_S/D = 0.7$) located 5 cm below the bottom impeller. The usual working conditions were $N = 2$–$7\,s^{-1}$ (corresponding to a specific power input of 200–2600 W m^{-3} for the BT-6) and $Q_G = 1.2$–$2.5 \times 10^{-3}\,m^3\,s^{-1}$ (which is equivalent to $U_G = 0.007$–$0.014\,m\,s^{-1}$ and 0.47–0.95 vvm for the $H/T = 2$ configuration, based on overall volume); during mass transfer and power consumption experiments, the conditions of 1.5 and 2.0 vvm were also investigated. Under the experimental conditions investigated, flooding did not occur at the bottom impeller, with either the BT-6 or the HSR-HF. Detailed analysis of the various flow conditions was not accomplished in this work. Limited surface aeration was noticed at high rotational speed under ungassed conditions, but it was not characterized as its influence on power draw and K_La was negligible (see below).

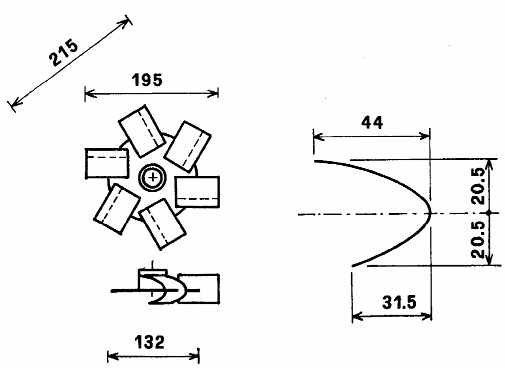

Figure 1. The BT-6 impeller, for clockwise rotation (dimensions in mm).

In addition to the experiments aimed at characterizing the liquid-phase mixing behaviour and measuring the mass transfer coefficients, which will be discussed in the following sections, the following basic parameters were determined: overall power consumption by measuring the restraining torque of the suspended motor and gas hold-up with the visual measurement of liquid level with and without gassing.

Mixing Time Measurements

The mixing behaviour of the liquid phase was investigated by means of the dynamic technique which is usually adopted for determining mixing time—either in the absence or the presence of the gas. A KCl solution was rapidly injected as the tracer about 5 cm below the liquid surface with a three-discharging-point needle. The resulting concentration was detected as a function of time at several vertical positions through conductivity measurements. In the case of no aeration, a standard two-electrode conductivity probe was simply inserted in the vessel for this purpose, while under aerated conditions a small portion of liquid was withdrawn continuously from the measurement point at constant flow rate after separating the gas and subjected to measurement in a continuous mode. The two techniques gave the same result under ungassed conditions (Pinelli *et al.*, 2001).

The experimental curves obtained at several vertical positions (shown in Figure 2) were used for determining the mixing time t_{95}, i.e. the time needed to get 95% homogeneity, as well as for matching the theoretical curves provided by a simple flow model. Several curves were always determined for each operating condition (usually 4, occasionally 7) in order to evaluate the reproducibility of the response curves, the mixing time and the flow model parameter. The error associated to mixing time measurement with dual BT-6 was in the range 5–20% (average 15%), slightly lower with dual and triple HSR-HF.

Volumetric Mass Transfer Coefficients

The steady-state technique based on H_2O_2 decomposition to water and oxygen in the presence of MnO_2 as a catalyst was adopted for the measurement of K_La (Vasconcelos *et al.*, 1997). An aqueous H_2O_2 solution was fed to the vessel at a constant rate. The entering hydrogen peroxide was balanced by its decomposition to O_2 and oxygen transfer to the gas. The volumetric mass transfer coefficient was calculated from:

$$K_La = \frac{\mathrm{OTR}}{(C_L - C_G/m)_{ML}} \tag{1}$$

The oxygen transfer rate, OTR (from the liquid to the gas) is equal to the oxygen production rate (related, in turn, to the feed rate). The log mean oxygen concentration difference was adopted as the driving force, which implies well-mixed liquid phase and gas in plug flow. This combination of simple flow models is considered a suitable approximation for multiple impellers (Nocentini, 1990). (The other possible assumption of well-mixed gas phase would have given a higher estimate of the driving force and, hence, a lower K_La value; the average and maximum difference for the investigated conditions was 6 and 13%, respectively.)

In addition to the impeller rotational speed and gas rate, the following parameters were measured in each experiment: temperature, concentration and feed rate of the entering solution, oxygen concentration of the gas C_G at the inlet, oxygen concentration in the liquid C_L (measured at mid vessel height, position D in Figure 2). The value of C_G at the exit was calculated through an overall mass balance and exhibited a maximum increase of 12% relative to the entering one. The flow rate of the hydrogen peroxide feed was rather small so that the liquid volume V in the vessel could be considered constant over the whole period of the experiment.

For each N and Q_G pair, at least four experiments were performed. All K_La data were normalized to 20 °C for better comparison by means of the relationship: $(K_La)_{20°C} = (K_La)_T\ 1.024^{20-T}$.

THE FLUID DYNAMIC MODEL

The axial dispersion model was used for the analysis of the liquid phase behaviour in a way similar to that used with radial turbines and axial flow impellers (Fajner *et al.*, 1982; Pinelli and Magelli, 2000).

For a batch system, the following mass balance equation and boundary conditions describe the concentration response to a pulse of n_0 moles of a passive tracer into one end of the system for this model (Wen and Fan, 1975):

$$\frac{\partial C_t}{\partial t} = D_{eL}\frac{\partial^2 C_t}{\partial z^2} \tag{2}$$

$$C_t(0, z) = 0 \tag{3}$$

$$-D_{eL}\left.\frac{\partial C_t}{\partial z}\right|_{z=0} = \frac{n_0}{S}\delta(t) \text{ and } \left.\frac{\partial C_t}{\partial z}\right|_{z=H} = 0 \tag{4}$$

The dimensionless solution of the above equations is available in the literature (Siemes and Weiss, 1957).

RESULTS AND DISCUSSION

Power Consumption

The ungassed power number (based on the swept impeller diameter of 0.215 m) of the dual BT-6 system was equal to 3, thus providing the value of 1.5 per turbine. The previously figure of $N_P = 2.4$ reported by Myers *et al.* (1999) for $D/T = 0.3$–0.5 was based on the nominal impeller diameter measured at the disk, which is 0.195 m for the impellers used here (see Figure 1). If the power number is calculated with the same diameter, the agreement between the current results and the literature (Myers *et al.*, 1999) is excellent. For $Re > 3 \times 10^5$ the onset of surface aeration was noticed, which caused very limited decline in N_P values.

On aeration, a very limited drop of P_g value was obtained while varying Q_G up to 2 vvm at constant rotational speed (Figure 3). This result is consistent with that reported (Myers *et al.*, 1999) for a single BT-6. It can be deduced that there is very limited asymmetry in the extent of gas by-passing between single and dual/multiple impellers typical of Rushton turbines (Smith *et al.*, 1987; Nocentini *et al.*, 1988).

The instantaneous value of torque was also recorded and it was noted that its fluctuations were much smaller than those exhibited by hydrofoil impellers pumping downward (Figure 4): the relative intensity of power fluctuation

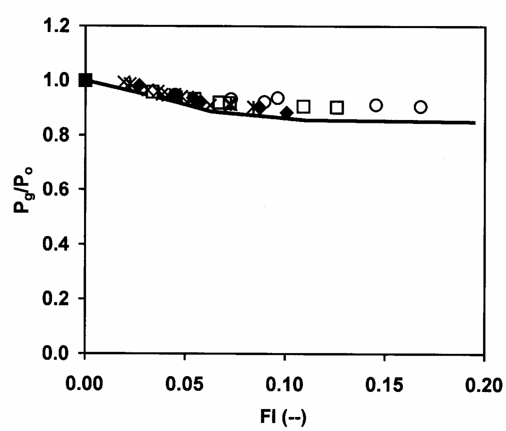

Figure 3. Relative power consumption for BT-6s on aeration. Symbols (data for the dual turbine system): ○, $N = 3\ s^{-1}$; □, $N = 4\ s^{-1}$; +, $N = 5\ s^{-1}$; △, $N = 6\ s^{-1}$; ×, $N = 7\ s^{-1}$; line: average data for a single turbine.

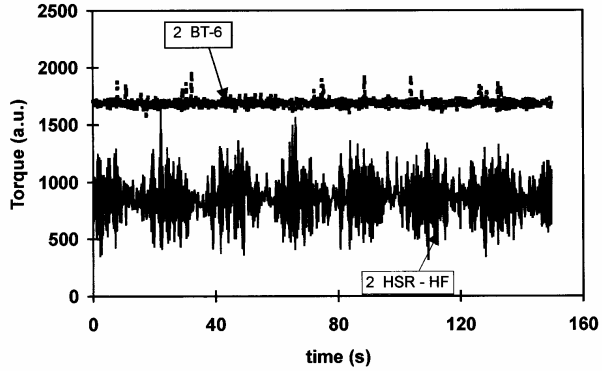

Figure 4. Torque fluctuations for the BT-6 and the HSR-HF impellers (arbitrary units).

(Bujalski *et al.*, 1997) of the former under ungassed conditions is roughly one third of the latter.

Hold-up

The experimental hold-up data for the dual BT-6 system were correlated to impeller speed and gas rate with the usual relationship:

$$\varepsilon_g = A\left(\frac{P_g}{V}\right)^{\alpha}(U_G)^{\beta} \quad (5)$$

and are shown in Figure 5, where the values obtained with dual HSR-HF are also plotted for comparison. It appears that the BT-6s provide slightly higher values than the other system tested—the experimental error for these measurements was in the range 5% (for the higher values) and 15% (for the lower ones). The following best fit values were obtained for the parameters in Equation (5): $A = 0.096$, $\alpha = 0.28$ and $\beta = 0.48$ (average error 3.7%) and were used for the abscissa in Figure 5. Since they are not directly comparable with data reported for other impellers (Pinelli *et al.*, 1994; Bouaifi *et al.*, 1997), the exponents were fixed as $\alpha = 0.24$ and $\beta = 0.65$: the resulting value $A = 0.252$

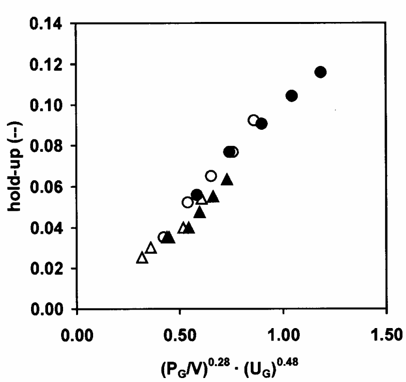

Figure 5. Gas hold-up for dual BT-6 and dual HSR-HF. (○ ●) 2 BT-6; (△ ▲) 2 HSR-HF (open symbols, $U_G = 0.007\,\mathrm{m\,s^{-1}}$ and 0.47 vvm; solid symbols, $U_G = 0.014\,\mathrm{m\,s^{-1}}$ and 0.95 vvm).

(average error 8%) is on the upper side of the range reported by Bouaifi *et al.* (1997) for several dual impeller configurations.

The relationship proposed by Smith (1991) for correlating the data for various configurations

$$\varepsilon_g = A(Re \cdot Fr \cdot Fl)^{\alpha}\left(\frac{D}{T}\right)^{1.25} \quad (6)$$

was also tested with the present data. With the exponent $\alpha = 0.45$, the value $A = 0.005$ was calculated which is 50% higher than that obtained with other configurations (Pinelli *et al.*, 1994).

Mass Transfer Coefficients

The mass transfer coefficient exhibited an increase with N and Q_G. Since it was suspected that surface aeration could affect the K_La values at high rotational speeds and low gas rate, a second set of experiments was performed with the liquid level raised by 18.5 cm above the standard condition shown in Figure 2 (i.e. to $H = 114.5$ cm). The values of the two sets of data did agree within ±5%.

The usual relationship

$$K_La = A\left(\frac{P_g}{V}\right)^{\alpha}(U_G)^{\beta} \quad (7)$$

was adopted to correlate the experimental data for the dual BT-6 system (mostly obtained in completely dispersed conditions) and the following best fit values were obtained: $A = 0.018$, $\alpha = 0.365$ and $\beta = 0.29$. A parity plot of the experimental values with those calculated with this set of parameters is shown in Figure 6. The exponents were also fixed at $\alpha = 0.59$ and $\beta = 0.40$ according to Linek *et al.* (1987) and the constant was found to be $A = 0.0053$ (although with a slightly lower correlation coefficient): this value is in good agreement with that obtained for a Rushton turbine (i.e. $A = 0.00495$). Thus, these data support the well-established independence of the volumetric mass transfer coefficient from the specific impeller type (Vasconcelos *et al.*, 1997; van't Riet, 1979).

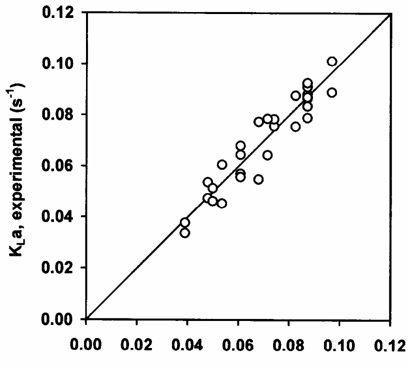

Figure 6. Correlation of K_La data for the dual BT-6 system by means of Equation (7).

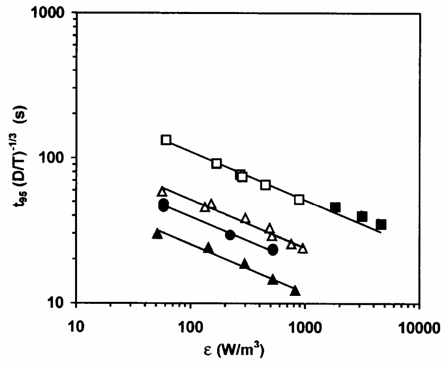

Figure 8. Ungassed mixing time for various multiple impellers ($Z = 0.04$). $T = 48$ cm. Symbols: (●) 2 BT-6; (▲) 2 HSR-HF; (△), 3 HSR-HF; (□), 3 Rushton $D/T = 0.32$; (■) 3 Rushton $D/T = 0.40$. Slope: $-1/3$.

Mixing Time

A first set of measurements dealt with the influence of detection elevation Z on t_{95} value (after a tracer pulse at the vessel top, $Z = 1$). As for radial and axial multiple impellers (Pinelli and Magelli, 2000; Nienow, 1998), the probe responses obtained in the vicinity of the upper turbine exhibited an overshoot, while those recorded in the lower part of the vessel featured a gradual approach to the asymptote. The mixing time data for the ungassed system are plotted in Figure 7. While the typical dependence $t_{95} \propto \varepsilon^{-1/3}$ (which is equivalent to $t_{95} \propto N^{-1}$) is confirmed for this turbine as well, mixing time decreases as the detection height is moved toward the injection point at the top of the vessel (high Z values). It can be observed that only a slight difference exists between the values at $Z = 0.5 \pm 0.13$, this fact suggesting limited staging with respect to what happens with the Rushton turbines (Cronin *et al.*, 1994). The existence of limited, though discernible, compartmentalization was confirmed by the visual observation of the spread of a coloured tracer in the stirred liquid.

It is worth noting that the dimensionless mixing time Nt_{95} evaluated at the bottom of the tank ($Z = 0.04$) is equal to 71, this value comparing well to those found for a number of dual axial-axial and radial–axial impeller configurations (Bouaifi *et al.*, 1997). In spite of the relatively high experimental error on t_{95} measurement (average error 15% for BT-6 and less than 10% for HSR-HF), Figure 8 shows that the ungassed mixing time for dual BT-6 is slightly higher than that for dual HSR-HF. For impeller comparison, the mixing time data were multiplied by $(D/T)^{-1/3}$ to allow for the differences in this geometric parameter, by extrapolating to multiple impellers the results obtained with $H = T$ and single impellers (Cooke *et al.*, 1988; Grenville *et al.*, 1995). For completeness, similar data for triple hydrofoil impellers and Rushton turbines are also included in the same plot. Any difference in behaviour between dual BT-6 and HSR-HF disappears on aeration (Figure 9). This fact seems to be related to the different behaviour of the two impeller types in terms of power draw on aeration. This most likely leads to a smaller reduction in impeller pumping capacity on gassing for the BT-6 than for the other impellers.

Axial Dispersion Coefficient

The experimental mixing time curves for the dual BT-6 system were also interpreted with the model given above. Indeed, the use of flow models provides a more rational

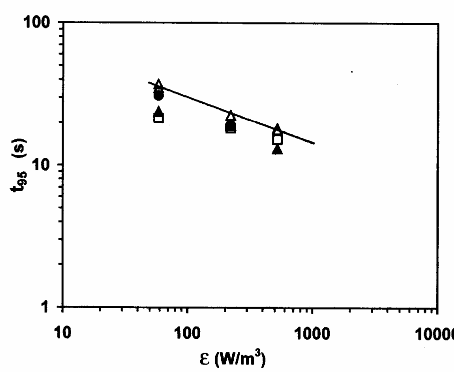

Figure 7. Ungassed mixing time t_{95} as a function of power draw per unit mass at several elevations. $T = 48$ cm, $H/T = 2$, 2 BT-6. Symbols: (△) $Z = 0.04$; (●) $Z = 0.22$; (□) $Z = 0.37$; (▲) $Z = 0.62$.

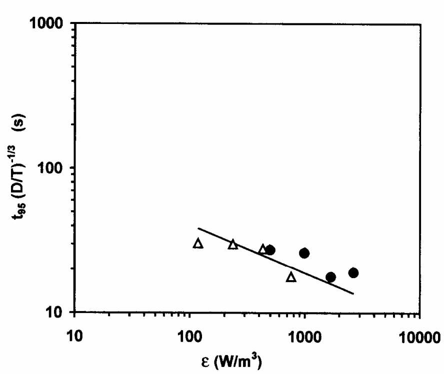

Figure 9. Mixing time $t_{95}(D/T)^{-1/3}$ vs. ε for dual BT-6 and dual HSR-HF in the $T = 48$ cm vessel. $Z = 0.04$; vvm $= 0.47$. Symbols: (●) 2 BT-6; (△) 2 HSR-HF. Slope: $-1/3$.

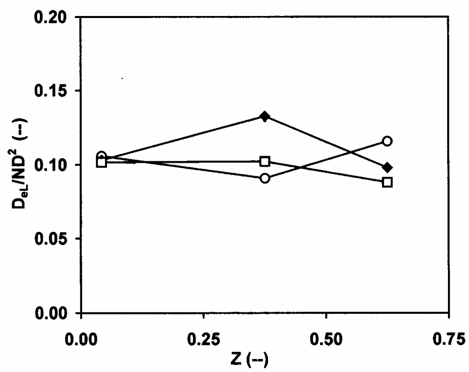

Figure 10. Dimensionless axial dispersion coefficient D_{eL}/ND^2 as a function of elevation Z for the dual BT-6 system; vvm = 0. Symbols: (○) $N = 2\,s^{-1}$; (◆) $N = 3\,s^{-1}$; (□) $N = 4\,s^{-1}$.

means to represent stirred systems than the empirical mixing time. The fit between the experimental and the model curves was quite satisfactory. Despite some scatter, the best fit parameter D_{eL} obtained from the curves is essentially independent of elevation (this is shown in dimensionless form in Figure 10), thus it is proper to characterize the behaviour of the whole equipment.

Consistent with mixing time results, the axial dispersion coefficient is (slightly) higher for the HSR-HF impellers than for the BT-6s under ungassed conditions. However, when data on aeration are compared at equal specific power draw, they become essentially equal: this result can be seen in Figure 11, where a line showing the 1/3 slope of the theoretical dependence (Baird and Rice, 1975; Jahoda *et al.*, 1994) of D_{eL} on ε is also plotted.

CONCLUSIONS

The paper deals with the behaviour of dual vertically asymmetric concave blade turbines in a gas–liquid stirred tank. This system exhibits the same qualitative dependence of hold-up on the working conditions as the other impellers, but the values lay in the upper part of the range. While mixing behaviour, as measured by either the mixing time or the axial dispersion coefficient, is somewhat worse under ungassed conditions compared with other impellers, on aeration it becomes equal to that produced by hydrofoil impellers at equal power draw per unit volume. The BT-6 turbines differ from other commonly used impeller systems, such as Rushton turbines or high solidity ratio hydrofoils, because of their flatter specific power curves. They are, therefore, suited for dispersing gas in reactors and fermenters where a broad range of gas rates is required.

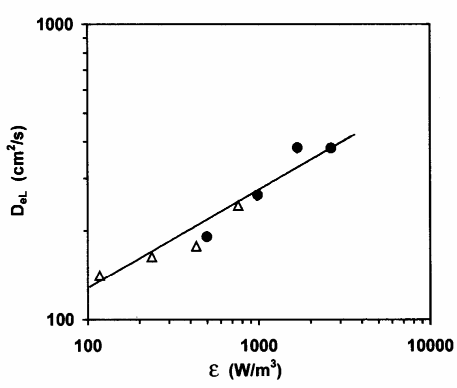

Figure 11. D_{eL} vs. ε for dual BT-6 and dual HSR-HF on aeration. $Z = 0.04$; vvm = 0.47. Symbols: (●) 2 BT-6; (△) 2 HSR-HF; line, 1/3 slope of the theoretical dependence.

NOMENCLATURE

A	constant in Equations (5)–(7)
C_t	tracer concentration in the liquid (kmol m^{-3})
C_G	oxygen concentration in the gas (kmol m^{-3})
C_L	oxygen concentration in the liquid (kmol m^{-3})
D	turbine diameter, based on the swept area (m)
D_{eL}	axial dispersion coefficient in the liquid (m^2 s^{-1})
Fl	$= Q_G/ND^3$, flow number
Fr	$= N^2D/g$, Froude number
g	acceleration of gravity (m s^{-2})
H	liquid height (m)
K_La	volumetric mass transfer coefficient (s^{-1})
N	rotational speed (s^{-1})
P_g	power draw under gassed conditions (W)
P_u	power draw under ungassed conditions (W)
Q_G	gas supply rate (m^3 s^{-1})
Re	$= ND^2\rho/\eta$, Reynolds number
S	tank cross sectional area (m^2)
t	time (s)
t_{95}	mixing time determined at 95% homogeneity (s)
T	tank diameter (m)
U_G	$= Q_G/S$, superficial gas velocity (m s^{-1})
V	liquid volume in the tank (m^3)
z	axial coordinate measured from the tracer injection point (m)
Z	$= z/H$, dimensionless axial coordinate measured from the bottom

Greek symbols

α, β	exponents in Equations (5)–(7)
$\delta(t)$	Dirac function (s^{-1})
ε_g	fractional gas hold-up
ε	P/V, specific power draw (W m^{-3})
ρ	liquid density (kg m^{-3})
η	liquid viscosity (Pa s)

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ADDRESS

Correspondence concerning this paper should be addressed to Dr D. Pinelli, DICMA—Department of Chemical, Mining and Environmental Engineering, University of Bologna, viale Risorgimento 2, 40136 Bologna, Italy.
E-mail: davide.pinelli@mail.ing.unibo.it

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