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Aeration and Mixing in Vortex Fermenters

Yusuf Chisti* & Murray Moo-Young

Department of Chemical Engineering, University of Waterloo, Waterloo, Ontario, Canada N2L 3G1

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Abstract: The overall apparent volumetric gas–liquid mass transfer coefficient ($k_L a$) and the mixing time (t_{95}) were determined in a 240 dm^3 vortex aerated fermenter over stirrer speed and air flow ranges of 300–800 rpm and 10–45 normal $dm^3\ min^{-1}$, respectively. The mass transfer data obtained in an aqueous salt solution (2·5 kg m^{-3} NaCl in water) compared well with the measurements in a fermentation medium used in culture of certain microaerophilic bacteria. Over the ranges examined, the gas–liquid mass transfer coefficient depended only on air flow rate; the dependence was linear with flow. Mixing time declined with increasing agitation according to a power-law relationship. The mixing and mass transfer characteristics of the vortex aerated system were compared with that of a 'standard' stirred tank fermenter (27 dm^3). The mixing time variations with respect to agitation rate were remarkably similar for the two types of fermenters examined.

Key words: vortex aeration, bioreactor, stirred tank, mixing, gas–liquid mass transfer.

NOTATION

a	Gas–liquid interfacial area per unit liquid volume (m^{-1})
DO	Dissolved oxygen
d_i	Diameter of the impeller (m)
d_T	Diameter of the tank (m)
k_L	Liquid film mass transfer coefficient (ms^{-1})
N	Rotational speed of the impeller (s^{-1})
P	Total power input (W)
Q	Volume flow rate of air (normal $dm^3\ min^{-1}$)
t_{95}	Mixing time for 5% deviation from complete mixing (s)
t_{95v}	Mixing time per unit liquid volume ($s\ m^{-3}$)
V_L	Volume of the liquid (m^3)

1 INTRODUCTION

Biotechnological production processes based on microbial, animal or plant cells are predominantly aerobic, a supply of oxygen is necessary for satisfactory cell growth and product formation.[1] Many of these operations are carried out in air-sparged stirred tanks, bubble columns or airlift bioreactors.[2] In some applications, particularly when oxygen requirements are relatively low and the fermenter or bioreactor volume requirements are small, vortex aeration can be satisfactorily used to meet the oxygen demand. This method of oxygen supply reduces foam formation as air is not bubbled through the liquid and it avoids bubble-associated cell damage which may be of concern with fragile cells such as some animal and insect cells.[3,4] Despite these advantages and known industrial applications in such microaerophilic fermentations as the production of diphtheria and pertussis vaccines, little data exist on vortex fermenters. The available literature concentrates on the fluid mechanics of vortex formation,[5] prediction of vortex depth[6,7] and power draw for vortexing tanks.[8] The gas–liquid mass transfer characteristics of these systems are relatively unknown, especially for systems larger than 50 dm^3. Similarly, there is insufficient information on the mixing behaviour of these devices. This paper reports on gas–liquid mass transfer and fluid mixing studies in a vortex-aerated stirred tank of 300 dm^3 nominal size. The performance of the vortex system is compared with that of the conventional 'standard' stirred tank geometry[8]

* To whom correspondence should be addressed.

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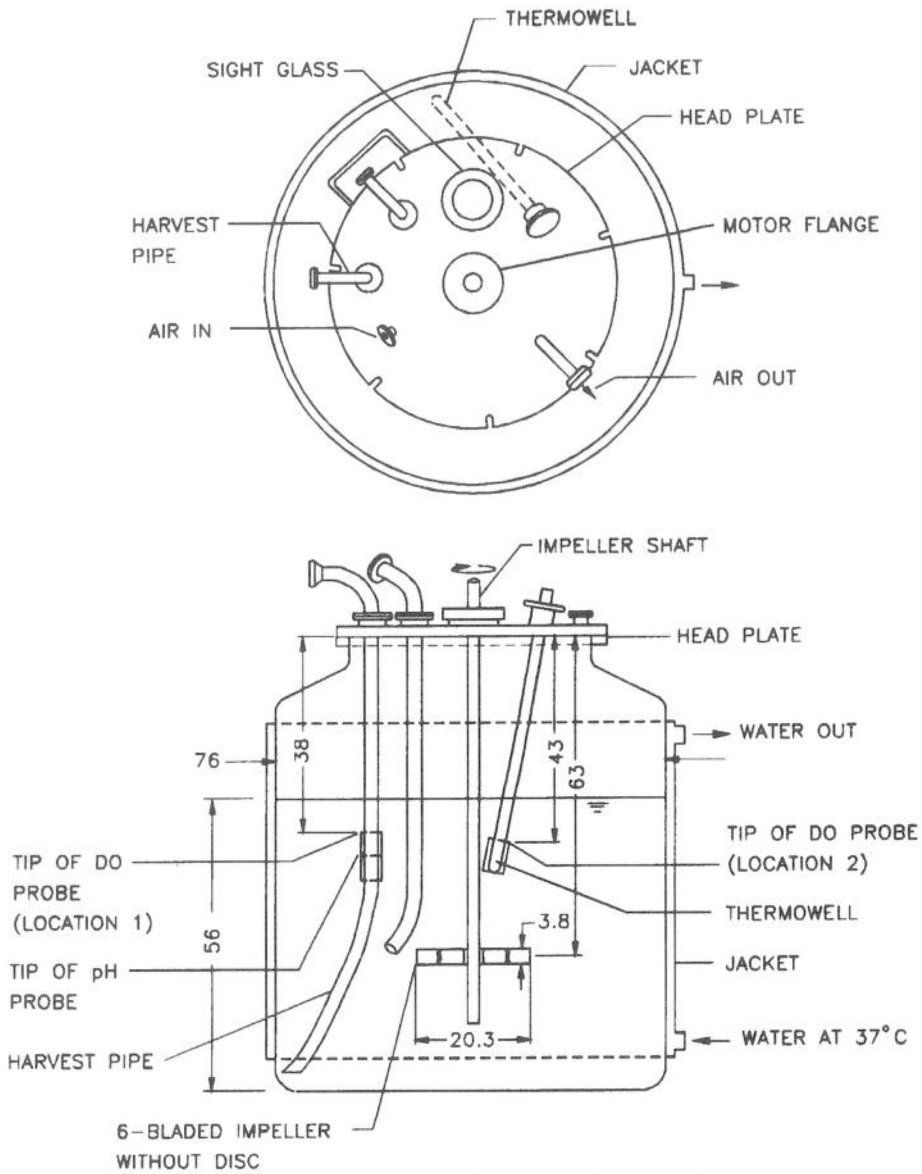

Fig. 1. The vortex fermenter (to scale); dimensions in cm.

which is more familiar to practitioners in the fermentation industry.

2 EXPERIMENTAL

2.1 Bioreactors

The geometric details of the 300 dm^3 (nominal) vortex-aerated industrial fermenter used in this work are illustrated in Fig. 1. A six-bladed turbine was used for agitation in an unbaffled tank which contained other such internals as thermowell, harvest pipes, etc. to effectively reproduce the hydrodynamic behaviour which would occur in actual fermentations with these systems. A working volume (liquid volume) of 240 dm^3 was used in all experiments. Aeration was in the headspace region only, with the air supply connection terminating flush with the inside surface of the head plate of the fermenter. All oxygen transfer to the liquid occurred through the surface of the vortex and through air bubbles entrained when the vortex reached the eye of the impeller. For comparison, measurements were done also in an MBR-Sulzer stirred bioreactor (70 dm^3 nominal), configured to the standard geometry of Fig. 2. This vessel had a liquid volume of 27 dm^3.

2.2 Fluids

Aqueous sodium chloride solution (0·043 kmol m^{-3}) NaCl) simulating the ionic strength effects of the fermentation broths of interest (e.g. diphtheria vaccine broth) was used in most gas–liquid mass transfer work. An actual bacterial fermentation medium (Table 1) was employed in one experiment to validate the data obtained

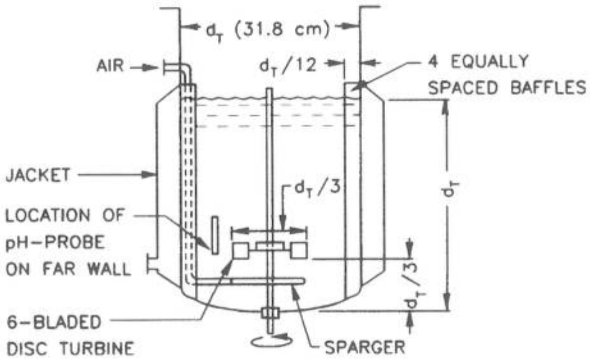

Fig. 2. The 'standard' stirred tank fermenter; dimensions in cm.

TABLE 1
Composition of the Fermentation Medium (in 240 dm^3)

Component	*Amount* (*g*)
Monosodium glutamate	2570·0
L-Proline	58·0
Hydrolysed starch	240·0
Hydroxymethyl aminomethane	365·0
Sodium chloride	600·0
Potassium phosphate monobasic	120·0
Magnesium chloride	24·0
Calcium chloride	480·0

in the salt solution. All mass transfer runs were performed at 36±2°C.

Mixing studies were done in tap water (Ref. 9 for composition) at ambient conditions (*c.* 22°C). Prior to use, the water was acidified (pH 2–3) and aerated (20 min) to remove dissolved carbonates.

2.3 Mass transfer measurements

The apparent overall volumetric gas–liquid mass transfer coefficient, $k_L a$, was determined by the transient gassing-in technique[10] adapted for surface aeration. At any preset agitation conditions, the headspace of the vortex fermenter was flushed with nitrogen until a low value (*c.* 2 ppm) of dissolved oxygen was reached in the liquid, the nitrogen supply was replaced with air flow at some predetermined value, and the increase in liquid phase oxygen concentration was followed with time. The dissolved oxygen was measured with a polarographic electrode connected to a dissolved oxygen meter and chart recorder described previously.[11] Two different electrode locations were tested (Fig. 1) to ascertain the equivalence of $k_L a$ values and confirm the 'well mixed' liquid assumption with respect to mass transfer calculations. The mass transfer coefficients were calculated as explained elsewhere.[11] For the low $k_L a$ values typical of vortex aeration, the criterion that the response time of the DO electrode should be $\leqslant 1/k_L a$[12] applied throughout the range of measurements; hence, the response time of the electrode was disregarded in $k_L a$ calculations. The apparent $k_L a$ values reported here combined the headspace gas mixing and mass transfer effects. This approach is satisfactory for the particular types of fermenters used in this work.

2.4 Mixing

Liquid mixing was characterized by mixing time measurements in carbonate-free water by an acid tracer technique.[2] A batch (pulse) of sulfuric acid (*c.* 5 mol dm^{-3} H_2SO_4; 25 or 250 cm^3 depending on fermenter vessel volume) was instantaneously poured on the surface of the liquid in the fermenter. The pH was followed with time. In keeping with the usual practice, the mixing time (t_{95}) was calculated as the time taken by the acid pulse to reach 95% of its equilibrium concentration. The pH change on addition of the acid was from pH 6 to pH 2. After each tracer experiment, sodium hydroxide (*c.* 8 mol dm^{-3} NaOH) was added to the fermenter to readjust the pH to near pH 6. The pH electrodes used had virtually instantaneous response (< 2 s for full response) over the range of the experiments. The locations of the pH electrodes in the two fermenters are shown in Figs 1 and 2.

3 RESULTS AND DISCUSSION

3.1 Gas–liquid mass transfer

3.1.1 *Effects of impeller speed and gas flow*

A plot of the apparent overall volumetric oxygen transfer coefficient ($k_L a$) at various gas flow rates (Q) is shown in Fig. 3. Over the impeller speed range of 300–800 rpm, there was no noticeable effect of agitation speed on $k_L a$. Throughout this range of speeds, the vortex was fully developed and reached the eye of the impeller. Visual observation confirmed that changes in agitation did not affect the specific gas–liquid interfacial area because the shape and size of the vortex were not affected, implying that the k_L term in $k_L a$ was insensitive to agitation rate over the range tested. This was because increase in agitation increased swirling, but the velocity gradients inside the liquid near the surface were apparently not altered very much. Hence, the explanation for the insensitivity of $k_L a$ to impeller rotational speed in fully developed vortex flow. Upon reducing the impeller speed to a value so low that only a flat vortex formed at the surface of liquid, extremely low (and consequently imprecisely measured) $k_L a$ values were obtained. Clearly, in the range of impeller speeds from zero to the point at which the vortex became fully developed (i.e. it reached

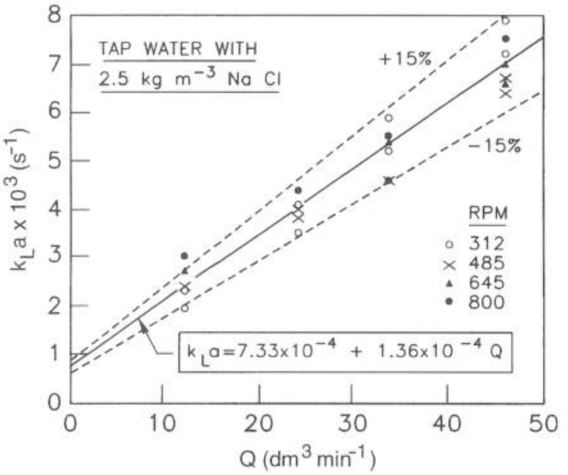

Fig. 3. The overall volumetric mass transfer coefficient (k_La) as a function of air flow (Q) in the vortex fermenter in salt solution. No effect of agitation speed (312–800 rpm) is seen.

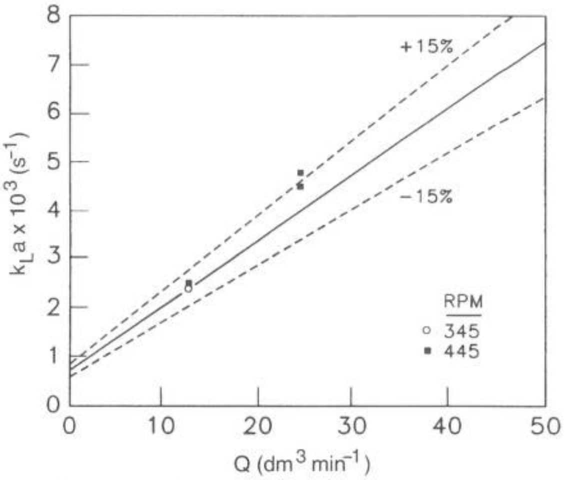

Fig. 4. Mass transfer coefficient vs air flow rate in vortex fermenter: effect of electrode positioning. Solid line represents eqn (1); the data points were obtained at electrode location 2 (see Fig. 1) for two different impeller speeds.

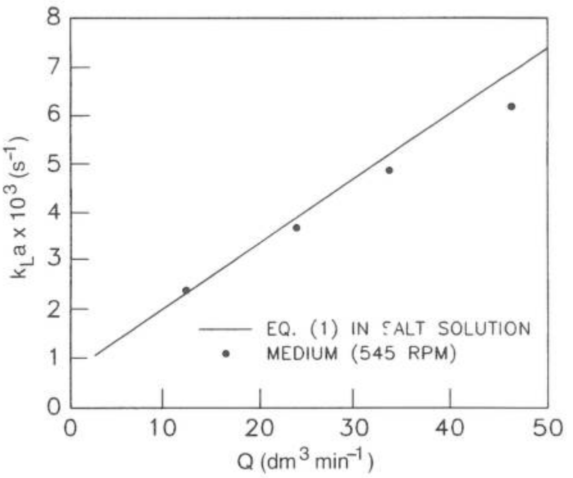

Fig. 5. Mass transfer coefficient vs air flow rate in vortex fermenter: comparison of mass transfer in salt solution (solid line) with that in the fermentation medium.

the eye of the impeller), the k_La was a function of impeller speed; however, the k_La values in this regime were not only too low to be measured with certainty, the regime was also of little practical interest. For example, at an impeller speed of 82 rpm a k_La value of $9{\cdot}8\times10^{-5}\pm2{\cdot}5\times10^{-5}$ s^{-1} was determined for an air flow rate of 33·3 normal dm^3 min^{-1}. Note that surface aeration without a sufficiently developed vortex is insufficient to support even the very low oxygen demands of animal cell cultures in reactors larger than 20 dm^3.

The apparent mass transfer coefficient increased with increasing air flow rate in the head region of the fermenter. The k_La data in salt solution were correlated (Fig. 3) with flow according to

$$k_La = 7{\cdot}33\times10^{-4}+1{\cdot}36\times10^{-4}Q \qquad (1)$$

which applied irrespective of the agitation rate. As shown in Fig. 3, the data fitted the equation within $\pm15\%$ with a correlation coefficient of 0·97.

3.1.2 *Effect of positioning of oxygen electrode*

The data shown in Fig. 3 were determined with the dissolved oxygen probe located on the downstream side of the harvest pipe, 0·38 m below the head plate (probe location 1 in Fig. 1). The electrode tip was unshrouded to allow for good movement of fluid, and the tip pointed upward to eliminate any bubble impingement effects. To ascertain that the position of the electrode in the fermenter did not affect the measured k_La, and that the liquid was indeed 'well-mixed' for mass transfer purposes, a second probe location was tested (Fig. 1). The k_La values determined with the dissolved oxygen electrode placed on the downstream side of the thermowell (unshrouded probe with tip upward), 0·43 m vertically below the head plate of the reactor, are shown in Fig. 4. Even at the two relatively low values of the impeller speeds, when mixing was expected to be relatively poor, the measured k_La values at the second electrode location agreed closely with eqn (1), demonstrating the liquid phase to be fully mixed for mass transfer calculations purposes (Fig. 4).

3.1.3 *Effect of the type of fluid*

Equation (1) for k_La was determined in salt solution (2·5 kg m^{-3} NaCl in water) which was considered to be a good simulation of the fermentation medium (Table 1) for the mass transfer study. Some k_La measurements were done in uninoculated fermentation medium to compare with the results obtained in the simulation salt solution. The k_La data obtained in the fermentation medium at the impeller speed of 545 rpm are shown in Fig. 5 along with the solid line of eqn (1). The oxygen mass transport properties of the medium and the salt solution were equivalent.

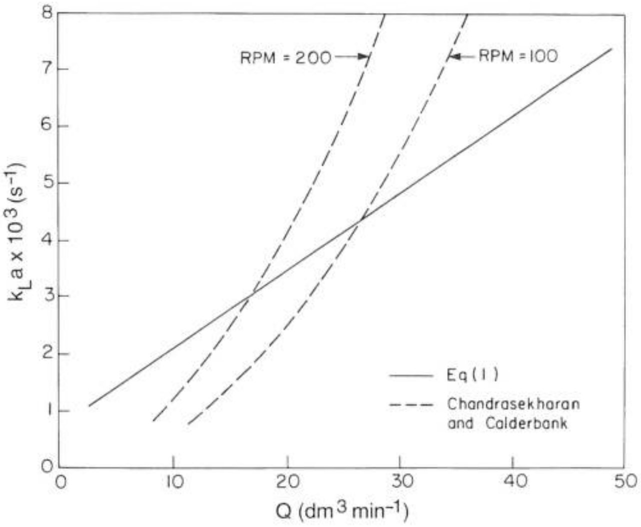

Fig. 6. Mass transfer coefficient vs air flow rate: comparison of eqn (1) for the vortex fermenter with eqn (2) for stirred tanks (dashed lines).

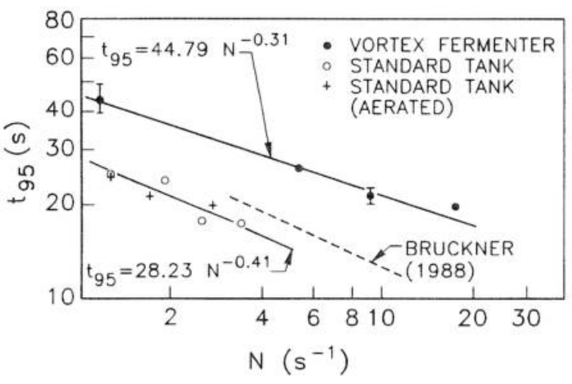

Fig. 7. Mixing time vs impeller speed.

3.1.4 *Comparison with standard stirred tanks*

Figure 6 compares the mass transfer performance of the vortex fermenter with that reported for standard stirred tanks.[13] The equation,

$$k_L a = \frac{0{\cdot}0248}{d_i^4}\left(\frac{P}{V_L}\right)^{0{\cdot}55}\left(\frac{Q}{60\,000}\right)^{0{\cdot}55/\sqrt{d_i}} \qquad (2)$$

developed by Chandrasekharan and Calderbank[13] is shown as dashed lines in Fig. 6. The solid line (Fig. 6) represents eqn (1) for the vortex fermenter. As shown in Fig. 6, at any fixed agitation rate, the mass transfer coefficient due to submerged aeration in standard stirred tanks is more sensitive to air flow than the apparent coefficient in vortex aerated fermenters. Note, however, that at equal agitation and air flow rates in the two types of fermenters, the apparent $k_L a$ in the vortex device can be higher (Fig. 6). Because at equal impeller Reynolds numbers, the power consumption in the unbaffled vortex system is usually less than in the baffled standard fermenters, the vortex devices may have better mass transfer economies than the more traditional reactors for fermentations with low $k_L a$ demands.

3.2 Liquid mixing

The mixing time (t_{95}) vs impeller speed (N) behaviour of the vortex and the standard fermenters is shown in Fig. 7. For the vortex device, the mixing time declined with agitation rate as follows:

$$t_{95} = 44{\cdot}79N^{-0{\cdot}31} \qquad (3)$$

Equation (3) fitted the data with a correlation coefficient of 0·97; the spread of the data about the mean values is shown by vertical bars for a few representative cases (Fig. 7). Earlier observations[14] in a smaller (total volume) vortex aerated fermenter had produced the mixing time correlation,

$$t_{95} = 28{\cdot}1405 - 2{\cdot}622N + 0{\cdot}108N^2 \qquad (4)$$

which is plotted in Fig. 7 as a dashed line for comparison with eqn (3) developed in this work. The two equations showed a similar trend with respect to agitation rate. Compared with the vortex device used in this study, mixing times were shorter in Bruckner's vessel even though the liquid volumes in the two reactors were identical at 240 dm^3. Moreover, the pumping capacity of the impeller used by Bruckner[14] must have been *c.* one-third of that in this work because the ratio (d_i, Bruckner/d_i, this work)^3 was *c.* 0·32. Because Bruckner's[14] vessel had a significantly higher aspect ratio (*c.* 1·1) than the 0·7 of this study, the results in Fig. 7 implied that the mixing efficiency in the axial direction exceeded that in the radial direction in vortex agitation.

The mixing time data in the standard tank also showed improved mixing with increasing agitation according to:

$$t_{95} = 28{\cdot}23N^{-0{\cdot}41} \qquad (5)$$

(correlation coefficient = 0·91). Note that the exponents on the agitation terms (N) in eqn (3) and eqn (5) are remarkably close for the two different types of fermenters.

In terms of the absolute mixing times for any agitation speed, the baffled tank was a superior mixer in comparison with the unbaffled, vortex agitated reactors. Because the volume of the standard tank (27 dm^3) was much smaller than that of the vortex device, comparisons based on absolute mixing times are not meaningful here. When the specific mixing time (mixing time per unit liquid volume, t_{95v}) was used as the criterion for comparison (Fig. 8), the mixing performance of the vortex device seemed better. Although, the standard stirred tank configuration was used as a base for evaluation of the vortex fermenter, the two systems were not geometrically similar. Thus, at 0·74 the aspect ratio

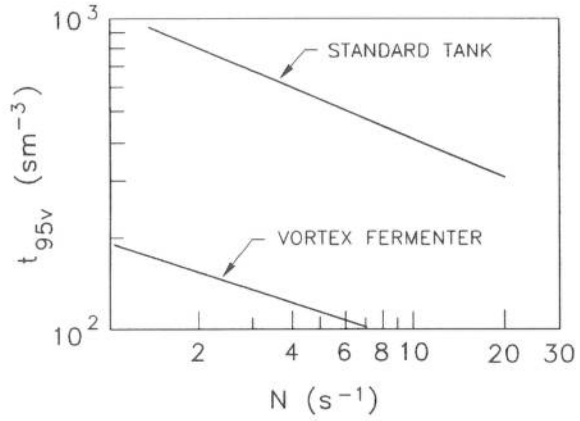

Fig. 8. Specific mixing time vs impeller speed.

of the vortex fermenter was less than that of the standard tank (Fig. 2), and this led to relatively poorer specific mixing performance in the latter device. Usually, for geometrically identical tanks of otherwise standard configuration, the vortexing, unbaffled, tank has poorer mixing efficiency than the non-vortexing tank.[8] A low aspect ratio of 0·74 was used because in fermentation applications it provides a higher surface-to-volume ratio for oxygen transfer than would be possible with the unit ratio used in standard stirred tanks. In the standard tank, air sparging (18·8 normal dm^3 min^{-1}) did not affect the mixing time because the air flow was relatively low (Fig. 7). Normally, aeration of the liquid is known to cause deterioration of mixing performance.

4 CONCLUSIONS

The apparent overall gas–liquid volumetric mass transfer coefficient ($k_L a$) and liquid mixing were evaluated in a 300 dm^3 (nominal) vortex fermenter as functions of the principal operating parameters. The performance of the vortex system was compared with that of the more traditional stirred tank fermenters. The main conclusions are:

1. In the vortex fermenter, the apparent overall volumetric gas–liquid mass transfer coefficient ($k_L a$) was insensitive to the impeller speed over the range 300–800 rpm; however, $k_L a$ increased with air flow rate in the head region of the fermenter.
2. The location of the dissolved oxygen electrode did not affect the $k_L a$ value in the vortex fermenter; hence, for mass transfer purposes, such fermenters up to 300 dm^3 may be considered 'well-mixed' in the liquid phase at agitation rates $\geqslant$ 300 rpm.
3. The aqueous salt solution was a satisfactory simulation of the fermentation medium for mass transfer investigations.
4. For microaerophilic fermentations the vortex fermenters can potentially yield better mass transfer economies than standard stirred tanks.
5. The mixing time in the vortex fermenter declined with increasing agitation speed according to a power-law relationship; very similar behaviour was observed in the 'standard' stirred tank.

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