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Free surface oxygen transfer in large aspect ratio unbaffled bio-reactors, with or without draft-tube

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ABSTRACT

It is widely accepted that animal cell damage in aerated bioreactors is mainly related to the bursting of bubbles at the air–liquid interface. A viable alternative to sparged bioreactors may be represented by uncovered unbaffled stirred tanks, which have been recently found to be able to provide sufficient mass transfer through the deep free surface vortex which takes place under agitation conditions. As a matter of fact, if the vortex is not allowed to reach impeller blades, no bubble formation and subsequent bursting at the free-surface, along with relevant cells damage, occurs.

In this work oxygen transfer performance of large aspect ratio unbaffled stirred bioreactors, either equipped or not with an internal draft tube, is presented, in view of their use as biochemical reactors especially suited for shear sensitive cell cultivation.

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1. Introduction

Gas–liquid stirred vessels are widely employed to carry out aerobic fermentations as well as chemical reactions involving a gas reagent and a liquid phase. For low viscosity liquids, the cylindrical vessel walls are typically equipped with swirl-breaking baffles, aimed at improving mixing performance. In fact, in the absence of swirl-breaking baffles, the relative velocities between the highly swirling liquid and stirrer blades are generally lower than those observed in baffled vessels: this results into smaller pumped flow rates and, in turn, into poorer mixing than in baffled tanks [1].

Despite allegedly being poorer mixers than baffled vessels, unbaffled stirred tanks are recently enjoying a growing interest in the process industry, as they provide significant advantages in a number of applications where the presence of baffles is undesirable for some reason [2]. This is for instance the case of crystallizers, where the presence of baffles may promote encrustations [3], or in food and pharmaceutical industries, where vessel cleanness is a topic of primary importance [4]. In bioslurry reactors for soil remediation processes, sufficient oxygen transfer may be guaranteed by the central vortex formation as an alternative to the adoption of gas spargers, which are intrinsically more troublesome [5]. As a matter of fact, solid particles could cause wear of the sparger holes, or the particles may form a muddy solid residue that blocks sparger holes. Such phenomena have adverse effects on the performance of the sparger and favor the adoption of unbaffled reactors or gas inducing impellers. When dealing with robust cells and non-foaming systems, liquid aeration can be further increased by adopting rotational speeds large enough to promote gas ingestion from the surface or by using a self-ingesting device [6–8].

Notably, when a suspended solid phase is present, higher values of the solid–liquid mass transfer coefficients may be obtained in unbaffled vessels, at the same value of mechanical power dissipation [9]. Also the mechanical power required to achieve complete suspension is found to be smaller than in baffled vessels [5,10].

As regards shear sensitive cell cultivations, mechanical agitation and especially bubble bursting at the free surface, unavoidably associated to air sparging, can cause cell death [11,12]. In unbaffled vessels a free surface deep vortex takes place under agitation conditions [13]. At agitation speeds such that the free-vortex bottom does not reach the impeller plane (subcritical conditions), no bubbles are dispersed in the liquid phase and therefore the cell damage associated with bubble bursting is avoided altogether. Under such conditions, the oxygen mass transfer that takes place through the vortex surface was shown to be sufficient for typical animal cell cultures [14]. This feature clearly makes unbaffled vessels potentially advantageous for shear sensitive cultures (e.g., animal cell or filamentous mycelia cultures), as well as for many foaming gas–liquid systems that share the need to avoid bubble dispersion in the liquid phase, provided that process rates, and relevant gas consumption

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Nomenclature

a	Gas–liquid interfacial area per unit volume of dispersion [m^{-1}]
C	Impeller clearance [m]
$C_L{}^0$	Initial oxygen concentration in the liquid phase [$kmol\ m^{-3}$]
C_L	Instantaneous oxygen concentration in the liquid phase [$kmol\ m^{-3}$]
$C_L{}^*$	Equilibrium oxygen concentration [$kmol\ m^{-3}$]
D	Impeller diameter [m]
Fr	Foude number $Fr = \frac{N^2 D}{g}$ [−]
H	Liquid height [m]
k_L	Oxygen mass transfer coefficient [$m\ s^{-1}$]
$k_L a$	Volumetric mass transfer coefficient [s^{-1}]
N	Rotational speed [rpm]
N_{crit}	Critical rotational speed [rpm]
N_p	Power number [−]
OTR	Oxygen transfer rate
P	Power input [W]
P/V	Specific power input [$W\ m^{-3}$]
Re	Reynolds number $Re = \frac{\rho_L N D^2}{\mu}$ [−]
SDPM	Simplified dynamic pressure method
t	Time [s]
t_0	Initial time
T	Tank diameter [m]
V	Tank volume [m^3]

Greek letters

ΔH	Vortex depth [m]
μ	Liquid viscosity [Pa s]
ρ_L	Liquid density [kg/m^3]

needs, are compatible with the relatively small gas transfer rates achievable [14]. In the above works fairly "standard" unbaffled vessels were investigated, especially as it regards vessel aspect ratio (always equal to one). Considering the encouraging results there obtained, one might wonder whether geometry changes such as the adoption of larger aspect ratios and/or the installation of an internal draft tube, may give rise to advantages in the realm of bioreactor applications. As a matter of fact, increasing liquid height is expected to increase the maximum agitation speed at which the agitator may be operated without gas ingestion, that should positively affect OTR, though on the other hand the liquid volume to feed is increased as well, so that this geometry change might result in either improved or worsened culture oxygenation with respect to standard aspect ratio vessels. Another geometry change that might improve culture oxygenation is the introduction of a draft tube, that should improve mixing rates while shortening vortex height, so allowing the adoption of larger agitation speeds and relevant improved OTR, still in the absence of bursting bubbles and associated cell damage effects. On the basis of the above considerations, in this work oxygen transfer performance of large aspect-ratio unbaffled stirred reactors, either equipped or not with an internal draft tube, is investigated in view of their use as biochemical reactors for shear sensitive cell cultivation.

2. Experimental

The investigated reactor is depicted in Fig. 1. It consisted of a flat bottomed cylindrical tank with an internal diameter of 280 mm and an height of 1450 mm (Fig. 1a). In some configurations (Fig. 1b and c) a removable internal cylindrical concentric draft-tube with internal diameter of 194 mm, thickness of 3 mm and length of 500 mm, off-spaced from vessel bottom by 60 mm, was mounted inside the vessel by means of suitable supports and rod spacers. A 0.095 m diameter six-flat-blade (blade height = 19 mm) hub-mounted turbine was installed on the 17 mm dia shaft, below the draft-tube, at a clearance C = 30 mm. Notably the purpose of placing the stirrer as close as possible to vessel bottom was that of maximizing water height above it and therefore maximizing the agitation speed achievable before vortex reached the impeller and air started to be dispersed in the liquid phase. For the small clearance impeller a turbine with radial action was deemed to be a better choice than an axial turbine.

Stirrer shaft was driven by a 1200 W DC motor (Mavilor MSS-12), equipped with tacho and speed control unit (Infranor SMVEN 1510) so that rotational speed was maintained constant, with a maximum deviation of 0.1% from the set point. Rotational speeds ranged from 100 to 1100 rpm in order to explore different fluid-dynamic regimes occurring inside the unbaffled stirred reactor. The vessel was filled with deionized water. Depending on the experimental run, liquid eight under no agitation conditions was either 280 mm (H = T), or 560 mm (H = 2T), or 840 mm (H = 3T) above vessel bottom. In the cases of H = 2T and H = 3T both the configurations with and without draft tube were investigated. It is worth noting that in the case H = 2T the liquid level at rest coincides with the draft tube brim. This was chosen to see whether, while in operation, the more complex flow field in the vicinities of the brim could result in improved mass transfer performance.

Power consumption was measured by monitoring the temperature rise due to agitation power input [7,8]. All temperature dynamics showed a remarkably constant slope. For simplicity, in the computation of power dissipation the heat capacity of vessel walls, shaft and impellers were neglected in front of that of the water mass. Also, heat exchange through vessel walls was neglected, on the basis of the very small temperature differences between water and ambient air (smaller than 0.5 °C, as each run was started from equilibrium conditions with the surroundings). This allowed to directly convert the observed temperature rise speed (e.g., 3.07 10^{-4} °C/s at 850 rpm) into the relevant specific dissipation rate (1.56 W/kg). It can be stated that power consumption so computed is slightly smaller than the actual value, because of the already mentioned simplifications; however, the resulting underestimation was considered to be negligible for engineering purposes. The total agitation power was finally estimated by multiplying specific power dissipation by the system water mass (e.g., 34.5 kg, H = 2T).

The volumetric mass transfer coefficient, $k_L a$, was assessed via unsteady-state experiments by means of the simplified dynamic pressure method (SDPM) [15], a technique that was found to be particularly suitable for $k_L a$ assessment in culture media and fermentation broths [14]. In this method the driving force for oxygen absorption is obtained by a sudden change of vessel pressure, with no need for sudden gas phase composition changes. For a perfectly mixed system, if $k_L a$ and the interfacial gas concentrations are identical for all bubbles at any time, Eq. (1) is obtained:

$$k_L a = -\frac{1}{t - t_0} \ln\left(\frac{C_L^* - C_L}{C_L^* - C_L^0}\right) \quad (1)$$

where C_L^0, C_L and C_L^* are the initial (time zero), instantaneous (at time t) and interfacial (viz equilibrium) oxygen concentration in the liquid phase, respectively. Eq. (1) shows that, if the above hypothesis are reasonably satisfied, plots of $\ln(C_L{}^* - C_L)$ versus t should result into straight lines with a slope equal to ($-k_L a$). This is not rigorously true, due to the contemporaneous nitrogen transport effects, but these can be accounted for by applying a correction factor to the measured slopes, as described in Scargiali et al. [15].

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