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A new simple approach for the scale-up of aerated stirred tanks

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ABSTRACT

The scale-up of stirred tanks to very large size is challenging because measurements and correlations are mainly developed for small-scale apparatus. This paper presents the basic hydrodynamics of two-phase stirred tanks. The hydrodynamics in large-scale reactors are shown to be mainly a function of the superficial gas velocity in the system. For this reason scale-up by keeping the superficial gas velocity constant is suggested in comparison to scale-up by constant volumetric aeration per volume (vvm). It is shown that in order to achieve adequate mass transfer, large-scale stirred tanks, especially bioreactors, must operate mainly in the heterogeneous regime and therefore the correlations developed in small-scale and homogeneous regime are not directly applicable for scale-up. New simple ways to predict gas holdup, interfacial area and mass transfer in stirred tanks are presented and shown to follow experimental values. The model requires very little data as basis and can therefore be used in the initial stages of reactor design.

Keywords: Scale-up; Aeration; Mass transfer; Holdup; Stirred tank; Bioreactor

1. Introduction

A vast amount of literature has been published in previous years on several aspects of stirred tank design. Stirred tanks have been extensively studied to be used as bioreactors. However, most of the literature in the field deals with small-scale equipment, with only a few studies performed with industrial-sized reactors (Vrábel et al., 2000; Noorman, 2011). Due to financial concerns, experimental studies are usually only possible in laboratory or pilot scale, where the volume of a reactor is at maximum around 1 m^3.

Large bioreactors are currently used for the production of expensive specialty chemicals such as proteins and specialty sugars. Energy optimization of reactor performance for these high value products has not been needed. However, recent interest has been pointed toward various microbial systems as sources of fuels and bulk chemicals (Blanch, 2012). Citric and lactic acid are already being produced in large scale. In order to make such a bioprocess profitable, the scale of the reactor needs to be hundreds of cubic meters while the energy consumption should be low.

Aside from anaerobic ethanol producing yeasts, most microbes grow aerobically. Therefore, they require large amounts of oxygen to grow efficiently. This leads to the use of aeration in bioreactors in order to supply the required oxygen. For the oxygen to be available to the microbe, mass transfer from the gas to the liquid phase must take place. The rate of dissolution of oxygen into liquid is typically described by $-dc/dt = k_L a(C_L - C^*)$, where k_L is the mass transfer rate, a the interfacial area available for mass transfer, C the current concentration of oxygen in the liquid and C^* the equilibrium concentration. Therefore the process is limited by the mass transfer resistance, which decreases k_L, the interfacial area as well as the saturation concentration. Mass transfer is often found to limit the productivity of bioreactor systems (Garcia-Ochoa and Gomez, 2009).

Abbreviations: CFD, computational fluid dynamics; OUR, oxygen uptake rate; OTR, oxygen transfer rate.

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Nomenclature

A'	cross-sectional area of reactor (m^2)
A	surface area of gas slug (m^2)
a	mass transfer area ($m^2\,m^{-3}$)
b	radius of gas slug (m)
C_L	concentration of oxygen ($mol\,l^{-1}$)
C_s	system dependent coefficient of mass transfer correlation (1)
$c_{current}$	current value in CFD cell
c_{ave}	time averaged value in CFD cell
D_L	diffusion coefficient ($m^2\,s^{-1}$)
d_b	bubble size (m)
d_{slug}	slug diameter (m)
g	gravitational acceleration ($9.81\,ms^{-2}$) ($m\,s^{-2}$)
H	height of reactor (m)
h	height of gas slug
K	power law consistency index (Pa s)
k	H/T ratio
k_L	mass transfer coefficient ($m\,s^{-1}$)
k_La	volumetric mass transfer coefficient (s^{-1})
m	mass flow to reactor ($kg\,s^{-1}$)
n	power law index
P	power input (W)
p_0	reference pressure at NTP (Pa)
r	radius
T	reactor diameter (m)
t	times
U_T	terminal velocity ($m\,s^{-1}$)
V	reactor volume (m^3)
V_{slug}	volume of gas slug (m^3)
V_{gas}	volume of gas phase in reactor (m^3)
vvm	volume per volume aeration rate (min^{-1})
v_S	superficial gas velocity ($m\,s^{-1}$)
v_{trans}	transition superficial gas velocity ($m\,s^{-1}$)
Q	gas volume flow ($m^3\,s^{-1}$)
Q_{trans}	transition volume per volume aeration rate (s^{-1})
x_{O_2}	molar fraction of oxygen

Greek symbols

α	exponent of mass transfer correlation (1)
β	exponent of mass transfer correlation (1)
γ	exponent of mass transfer correlation (1)
θ_w	wake angle of gas slug (°)
ε	turbulence dissipation ($W\,kg^{-1}$)
μ_a	apparent viscosity (Pa s)
μ_L	liquid phase viscosity (Pa s)
μ_G	gas phase viscosity (Pa s)
ϕ	gas holdup
ρ	density ($kg\,m^{-3}$)
σ	surface tension ($N\,m^{-1}$)

A stirred tank is often the system of choice for bioreactors. Adding a second phase to the reactor complicates the system hydrodynamics (Moilanen et al., 2006). When a single phase is present, the fluid flow in the reactor is induced by the impellers. The flow inside these systems is relatively easy to predict. Once aeration is introduced, the gas flow hinders the effect of the impellers and the bubbles themselves alter the flow field. The flow conditions inside a two-phase stirred tank reactor can be divided into homogeneous and heterogeneous (Gezork et al., 2000, 2001). These regimes have also been identified for bubble columns (Kantarci et al., 2005; Yang et al., 2010). The flow field in a stirred tank is considered homogeneous when it is defined by the mixing and heterogeneous, when the gas flow defines the flow field. It is shown in this article, that even though stirred tanks are generally utilized, the large scale equipment is usually operated in the heterogeneous regime. Impeller flooding is closely related to the transition of the flow regime from homogeneous to heterogeneous. Flooding is defined as a state when the impeller does not spread gas horizontally, but there is an axial flush of gas through the impeller plane (Warmoeskerken and Smith, 1985). The transition to flooding is a function of the impeller rotation and aeration rate. When the impeller is flooded, the gas is not distributed efficiently over the whole cross section of the vessel, dead areas are created and mass transfer is not efficient. However, transition to flooding is not the same as transition to the heterogeneous flow field (Gezork et al., 2000). The onset of heterogeneous flow field may take place even when the impeller is not flooded at high aeration and stirring rate. Nevertheless, flooding of the impeller seems to lead to a heterogeneous flow field even at relatively low aeration rates. The onset of flooding can be estimated by correlation of the Froude number with the ratio of the impeller to the tank radius (Warmoeskerken and Smith, 1985) or by a mechanistic model by Paglianti (2002).

Flooding is an unwanted phenomenon in stirred tanks. Nienow (1998) presents a broad discussion on the use of Rushton turbines and other radial flow impellers along with the issue of flooding. Several correlations exist for the prediction of impeller flooding for Rushton turbines. However, even with the possibility to predict flooding and therefore avoid this circumstance, the use of Rushton turbines in industrial scale is problematic. The extensive power loss exhibited by Rushton turbines in aerated conditions has received much discussion and correlations exist to predict the gassed power input. The large difference in the gassed power input leads to situations where stirrer motors must be designed for variable speeds. Up-pumping axial flow impellers are seen as an alternative. They exhibit very good dispersion capabilities, very small losses of power due to aeration and are less likely to flood at high aeration (Nienow, 1998).

The state of scale-up and design of large bioreactors is at this moment, even with the existing research, mainly based on correlations, best practices and rules of thumb (Noorman, 2011; Garcia-Ochoa and Gomez, 2009; Humphrey, 1998; Takors, 2012; Junker, 2004; Posch et al., 2013). Bioreactors are often optimized in laboratory scale. Traditionally scale-up has relied on gradual increasing the scale of experiments from laboratory to bench scale (1–10 L), bench to pilot scale (50–300 L) and finally to production scale. The main criterion for scale-up of bioreactors is the oxygen transfer rate (OTR), which is optimized at every scale. A scale-up ratio of 1:10 is typical. For scale up to $100\,m^3$, this would mean a pilot plant of $10\,m^3$ would be necessary. This is, of course, not practical, so larger scale-up ratios are utilized. Four different approaches to scale-up are generally recognized (Garcia-Ochoa and Gomez, 2009): fundamental methods, semi-fundamental methods, dimensional analysis and rules of thumb. Fundamental methods consist of the physics based modeling of bioreactor systems. Modeling gives us a possibility to predict the behavior of large scale equipment without the absolute need for measurements at those scales. One such modeling method is Computational Fluid Dynamics, CFD. Many approaches to scale-up through