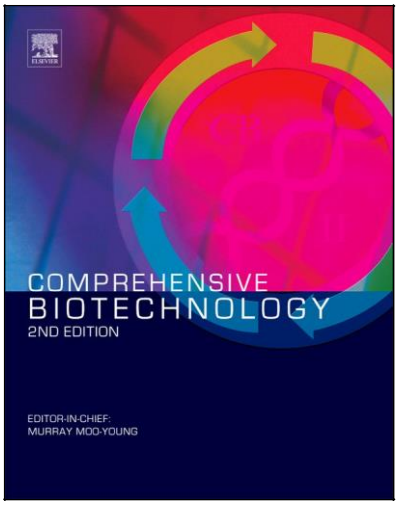

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2.16 Airlift Bioreactors

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Glossary

airlift (bio)reactor A configuration of pneumatically agitated (bio)reactor where injection of a gas stream takes place into a special section of the reactor known as the riser, which causes the reactor broth to circulate between the riser section and a section containing a lower (or negligible) volume of gas phase known as the downcomer.

downcomer A section of an airlift (bio)reactor that contains a much lower (or negligible) amount of gas phase than the riser section of the reactor.

pneumatically agitated (bio)reactor A configuration of agitated (bio)reactor where injection of a gas stream serves for both mixing and exchange of substrates and products.

riser A section of an airlift (bio) reactor where the gas stream is injected.

Nomenclature

Reactor Geometry

H	reactor height (m)
d	reactor diameter (m)
A_d	cross-sectional area of downcomer (m^2)
A_r	cross-sectional area of riser (m^2)
l	distance between the riser and the downcomer (m)
W	width of the gas–liquid separator duct (m)
h_L	height of the gas-free liquid (m)
h_D	gas–liquid dispersion height (m)
A_b	area available for flow under the draft tube (m^2)
A_{de}	minimum area of the downcomer entrance for efficient gas disengagement (m^2)
d_0	diameter of the sparger orifice (m)
d_r	riser diameter (m)
x	liquid height in the connecting section (m)

Liquid Phase

V_L	volume liquid phase (m^3)
ρ_L	density liquid phase ($kg\ m^{-3}$)
σ	surface tension of the cultivation medium ($N\ m^{-1}$)
μ_{eff}	effective viscosity of the cultivation medium (Pa s)

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μ_L	viscosity of the liquid phase (Pa s)
K_T	form friction loss coefficients for the top zone (–)
K_B	form friction loss coefficients for the bottom zone (–)
U_L	mean liquid circulation Velocity (m s^{-1})
U_{Lr}	liquid velocity in the riser (m s^{-1})
U_{LH}	liquid velocity in the connecting duct (m s^{-1})
U_{Ld}	liquid superficial velocity at the entrance of the downcomer (m s^{-1})
U_c	mean liquid circulation velocity (m s^{-1})

Gas Phase

ε_G	overall gas holdup (–)
ε_{Gr}	gas holdup in riser (–)
ε_{Gd}	gas holdup in downcomer (–)
V_G	volume of gas phase (m^3)
Q_m	molar gas flow rate (mol s^{-1})
T	temperature of the gas phase (K)
U_{Gr}	superficial gas velocity in the riser (m s^{-1})
U_g	superficial gas velocity based on the reactor cross-sectional area (m s^{-1})
U_b	bubble rise velocity (m s^{-1})

Solid Phase

V_S	volume of suspended solid phases (m^3)
ε_{Sr}	solid holdup in riser (–)
ε_{Sd}	solid holdup in downcomer (–)
ρ_S	density of the solids (kg m^{-3})
d_P	particle diameter (m)

Others

P_G	power input (W)
P_h	headspace pressure (Pa)
γ	average shear rate (s^{-1})
$N_{G/L}$	mass-transfer rate (mol m^{-3} s^{-1})
C^*	liquid-phase saturation concentration of the transferring substrate in equilibrium with the gas phase (mol m^{-3})
C_L	concentration of the transferring substrate in the bulk liquid phase (mol m^{-3})
K_L	gas–liquid mass-transfer coefficient (m s^{-1})
a	specific gas–liquid interfacial area (m^{-1})
a_p	specific liquid-solid interfacial area (m^{-1})
d_{32}	Sauter mean gas bubble diameter (m)
d_b	mean gas bubble diameter (m)
d_{max}	maximum stable bubble size (m)
$N_{L/p}$	mass-transfer rate from the liquid to the dispersed particles (mol m^{-3} s^{-1})
K_p	mass-transfer coefficient of the dispersed phase (m s^{-1})
C_p	substrate concentration in the solid/liquid interface (mol m^{-3})
D	substrate diffusivity in the liquid phase (m^2 s^{-1})
g	gravitational acceleration constant (m s^{-2})
R	universal gas constant (J K^{-1} mol^{-1})
We	Weber number (–)
Fr	Froude number (–)
ω, ν	empirical parameters in eqn 6 (as appropriate)
Ω	empirical parameter in eqn 13 (m^{-1})
ψ	empirical parameter in eqn 18 (–)
k, α, β, δ, and φ	empirical parameters in eqns 19 and 20
Sh	Sherwood number (–)
Sc	Schmidt number (–)
Re	Reynolds number (–)
$\prec$, χ, and θ	empirical parameters in eqn 23 (as appropriate)
ε	energy dissipation rate from Kolmogoroff's theory of local isotropic turbulence (m^2 s^{-3})
Q_H	heat-transfer rate (W)
h_T,	overall heat-transfer coefficient (W m^{-2} °C^{-1})
A_H	heat-transfer area (m^2)
T_s	temperature of the heating/cooling surface (°C)
T_b	temperature of bulk fermentation medium (°C)
$\wp$	thermal diffusivity of the liquid phase (m^2 s^{-1})
τ	shear stress (Pa)
A	consistency index (–)
n	flow index (–)
c	proportionality constant in eqns 30 and 32 (W s mol^{-1})
Q_x	microbial heat production and 32 (W s mol^{-1})
Q_{O_2}	oxygen uptake rate (mol m^{-3} s^{-1})
t_m	mixing time (s)
θ_m	dimensionless mixing time (–)
t_c	mean cycling time (s)
E_z	overall axial dispersion coefficient of the liquid phase (m^2 s^{-1})
Bo	Bodenstein number (–)
L	distance between tracer injection and detection points (m)
L_c	length of the circulation loop (m)
C_t	instantaneous tracer concentration (mol m^{-3})
$C_{t\infty}$	equilibrium tracer concentration (mol m^{-3})
λ	dimensionless time (–)
p	dimensionless distance (–)
t	instantaneous time (s)
L_t	distance traveled by the fluid at time t (m)

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2.16.1 Introduction

Pneumatically agitated bioreactors take advantage of the injection of a gaseous stream (often air) to provide mixing and mediate transfer of gaseous substances (i.e., O_2 and CO_2) with the liquid phase. However, unlike in classical pneumatically agitated reactors where liquid mixing is random (i.e., bubble column), the specific design of airlift reactors (ALRs) causes the liquid to circulate between two interconnected zones known as the riser and the downcomer [1]. The riser and the downcomer are connected by a specific reactor base allowing for liquid circulation and by a gas–liquid separator at the top. Under typical operation conditions, air is injected below the riser section and the removal of gas in the separator generates a mean density gradient between the riser and downcomer zones that causes the liquid broth to circulate (Figure 1). The function of the gas separator is to support efficient gas–liquid disengagement. The fraction of gas introduced in the downcomer section depends on design and operational variables. This fraction has a significant effect on fluid dynamics and, consequently, reactor performance.

The focalized introduction of energy for mixing in classical bioreactors generates large shear gradients that cause cells to experience mechanical stress in areas of high turbulence and suboptimal solutes concentrations (i.e., O_2, CO_2, H^+, and toxins etc.) and or temperature conditions in areas of low turbulence [2]. By contrast, liquid circulation between the riser and the downcomer (rather than gas injection) is the main contributor to fluid dynamics in ALRs. Because liquid circulation is caused by the gradient between the average fluid densities in the two reactor sections, there is no focal point of energy dissipation and shear forces are very homogeneous within each section, causing less cellular stress. The ALRs also supposedly support higher mass-transfer rates per energy input than classical systems and transfer efficiency (i.e., the amount of O_2 transferred per power input) is much less affected by power input in ALRs than in classical systems. The two main advantages of ALRs described here explain why these systems are often preferred for the cultivation of shear-sensitive mammalian and plant cells or during wastewater treatment applications requiring efficient energy use (aeration costs represent roughly 50% of the energy costs during domestic wastewater treatment).

Research and development on ALRs has hitherto focused in demonstrating the potential of this system in new applications or modeling the complex relationships between design and operational parameters and fluid dynamics and mass transfer. Many experimental and mechanistic models that can describe ALR operation and performance are thus available [3]. However, the validity of these models is too often limited to specific applications or reactor configurations. For this reason, only the most relevant, widely accepted, and generic models are presented here in order to illustrate how design and operational parameters influence fluid dynamics and mass-transfer properties.

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