GEOMETRIC FACTORS IN AIRLIFT MIXING: A RISER-DOWNCOMER INTERPLAY STUDY WITH A THREE-PHASE VISCOUS MIXTURE

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ABSTRACT

Airlift bioreactors, advantageous for microbial cultivation and enzymatic reactions, operate efficiently without mechanical stirrers due to their pneumatic design. The internal draft tube directs aerated bubbles, creating two flow regions: the riser and the downcomer, with mixing efficiency tied to the circulatory velocity influenced by tube geometry. Altering the tube's diameter and inclination significantly impacts mixing, critical when adding solids and viscous liquids that alter fluid dynamics. Conducted in a 1L airlift reactor with varying internal draft tube geometries, experiments assessed mixing times at 45°C with a 2 L/min gas flow, using grape seed oil and immobilized lipase. The continuity equation analysis for the riser and downcomer highlighted the influence of tube geometry on mixing times, with the optimal internal tube diameter identified as 93.9% of the sparger diameter ($Ø_S$), or 5.31 cm ($Ø_R$), for minimizing mixing time under tested conditions. The flow model, enhanced by a differential evolution algorithm, emphasized the riser backflow's role, significantly affecting downcomer velocity and reactor circulation. This novel model, demonstrating a coefficient of determination (R^2) of 0.957, marks a substantial improvement over existing models by incorporating backflow dynamics in this context. The ability to predict mixing times more accurately facilitates enhanced airlift reactor performance, presenting a methodology adaptable to various operational conditions and geometric configurations.

Keywords: Mixing time. Airlift reactor. Downcomer flow. Riser geometry. Three-phase mixture.

1 INTRODUCTION

Airlift bioreactors present an advantageous alternative to conventional stirred tanks in microbial cultivation and enzymatic reactions, offering efficient mixing without the mechanical complexity inherent to stirrers. With a pneumatic design, these reactors feature an internal draft tube directing aerated bubbles from a sparger, creating two main flow regions: the riser and downcomer. The efficacy of mixing is closely tied to the velocity between these zones, influenced by the tube's geometry. Strategic geometric optimization aims to decrease mixing time, crucial when adding solids and viscous liquids that alter fluid dynamics. This enhancement in mixing is designed to increase substrate contact, thereby leading to improved conversion rates. Thus, this investigation aims to evaluate how different internal tube geometries affect mixing times in a three-phase viscous mixture. By establishing a quantitative relationship between mixing time and flow velocity across various geometrical configurations, we seek to identify the optimal conditions for conducting enzymatic reactions in airlift bioreactors, ensuring effective mixing and contact.

2 MATERIAL & METHODS

To characterize the pneumatic system for enzymatic reactions with a solid phase comprised of Immobilized lipase from *Rhizomucor miehei* (Novozymes A/S - Bagsvaerd, Denmark - Lipozyme RM IM®), experiments were conducted in a 1L airlift reactor with diameter ($Ø_T$) of 0.908 m using grape seed (*Vitis vinifera* L.) oil as the liquid phase with 2% (v/v) solids loading. Maintaining a constant height (H_R) of 0.105m, the riser's bottom diameter ($Ø_R$) varied from 5.45 to 7.45 cm, while the inclination angle (θ) ranged from -2 to 10 degrees. The sparger employed was of a cross type, featuring a diameter ($Ø_S$) of 5.45 m, equipped with holes of 0.5 mm in diameter, spaced 2 mm apart.

Mixing characterization involved determining the mixing time (t_m – seconds) at 45°C with a gas flow rate of 2 L/min to replicate reaction-specific conditions. The temperature pulse technique was employed using 60 mL of grape seed oil heated to 85°C acting as the thermal pulse. This approach measured the time required for the temperature to reach 95% of its final, equilibrium value. This process was monitored using three DS18B20 temperature sensors located in the disengagement zone, downcomer, and riser. The gathered mixing time data were then used to adjust the flow model employing a differential evolution algorithm in Python.

3 RESULTS & DISCUSSION

For the purpose of refining a flow model and assessing the mixing within an airlift reactor operating at a steady fluid flow rate (Q – m³/s), the continuity equation, delineated in Equation 1, was examined individually for both the riser (Q_R) and downcomer (Q_D) components. In line with the principles of momentum conservation, the mean downcomer area (A_D – m³) is defined by its hydraulic diameter, given by the difference between the total reactor area (A_T – m³) and the mean riser area (A_R – m³). Consequently, alterations in the riser's cross-sectional area induce inversely proportional velocity variations in the downcomer (U_{LD} – m/s).

$$Q_R = Q_D \Longrightarrow U_{LR}A_R = U_{LD}A_D \Longrightarrow U_{LR}A_R = U_{LD}(A_T - A_R) \qquad (1)$$

In airlift bioreactors, the riser superficial liquid velocity (U_{LR} – m/s) is governed by a flow driving force, which is influenced by the rate at which gas is introduced into the reactor. This driving force is particularly contingent upon the gas holdup within the riser (ε_{GR}) – this being under the no bubble recirculation regime [1,2]. Furthermore, the premise is based on the insight that the bulk density of the three-phase mixture within the riser ($\rho_{R\ bulk}$) can be effectively approximated by the mean density of the riser ($\rho_{R\ mean}$), as elucidated in Equation 2, once the density of solids (ρ_S) and the liquid (ρ_L) can be approximated as the same. Consequently, the driving force is predominantly attributed to the discrepancy between the gas (ρ_G) and liquid densities [3].

$$\rho_{R\ bulk} \approx \rho_{R\ mean} = \rho_{GR}\varepsilon_{GR} + \rho_{LR}(1 - \varepsilon_{GR}) \quad (2)$$

This equation holds true when the two-phase (oil/air) mixture is homogeneous up to a certain level, depending on the size and coalescence of the bubbles, flow rate, and the relative volume occupied by the two phases (void fraction) [4,5]. In examining the scenario where the diameter of the riser is enlarged, while maintaining a constant airflow, an augmentation in the volume available for the liquid is expected. Consequently, the approximate overall riser density begins to markedly deviate from the values of local bulk density. Therefore, the effectiveness of the flow driving force diminishes due to reduced gas holdup in the riser. This reduction is directly proportional to the discrepancy between the sparger (Ø_S – m) and the riser mean diameter (Ø_R – m), which defines the volumetric difference between the ascending bubble column and the liquid adjacent to the inner wall of the riser [6]. Hence, the riser fluid backflow factor (σ_R) is introduced to quantify the reduction in liquid flow through the downcomer due to return phenomena, a consequence of the riser void fraction's impact on the driving force and on the resultant liquid backflow turbulence, contrary to the intended directional flow towards the downcomer. This backflow occurs as a result of the mixture's heterogeneity within the riser, which leads to a loss of downcomer flow velocity. The factor σ_R links velocity changes in the downcomer to circulation velocity variations in the riser, according to Equation 3, by modifying Equation 1. It's defined by a normalized sigmoid function based on the reactor's external column diameter (Ø_T – m) in Equation 4.

$$U_{LD}=U_{LR}(1-\sigma_R)A_R/(A_T-A_R) \quad (3)$$

$$\sigma_R = \left(1 + e^{\left[\frac{-\gamma(\text{Ø}_R-\text{Ø}_S)}{(\text{Ø}_T-\text{Ø}_S)}+\frac{\psi}{\text{Ø}_R}\right]}\right)^{-1} \quad (4)$$

Here, γ represents the system's susceptibility to liquid return via the riser, with a greater value indicating heightened sensitivity to diameter variations near the critical diameter (value where its derivative is the maximum). In other hand, ψ pertains to the critical diameter for return; a larger value signifies an increased diameter at which backflow phenomena is observed. The combined influence of these parameters offers enhanced flexibility in tuning the model to fit experimental observations. As shown in Figure 1a, the data points within the black curve represent the conditions tested experimentally. The liquid's ascent velocity through the riser is dependent on the ratio of two elements, α and β, which denote the flow's driving force (buoyancy-driven force that is attributed to the gas bubbles) and frictional losses, respectively, as depicted in Equation 5. Further examination of the terms associated with frictional forces reveals that this relationship is shaped by the coefficient of frictional loss at the bottom (K_B) [7] and the Fanning friction factors for both the riser (f_R) and downcomer (f_D) [8], as elucidated in Equation 6.

$$U_{LR} = [\alpha/\beta]^{0.5} \quad (5)$$

$$\frac{\alpha}{\beta} \propto \frac{\text{driving force}}{\text{friction losses}} = \frac{2gH_R\varepsilon_R}{K_B\left([1-\varepsilon_R]^{-2} + \left[\frac{A_R}{\{A_T\text{-}A_R\}}\right]^2\right) + 4H_R\left(\frac{f_R}{\text{Ø}_R} + \frac{f_D}{\text{Ø}_T - \text{Ø}_R}\left[\frac{A_R}{\{A_T\text{-}A_R\}}\right]^2\right)} \quad (6)$$

Upon resolving Equations 3 to 6, one can observe the impact of backflow on U_{LD}, leading to a decrease in the reactor's overall circulation speed by diminishing the fluid flow within the downcomer. This phenomenon is depicted in Figure 1b, where U_{LD}^* denotes the standard model for downcomer velocity, without σ_R consideration.

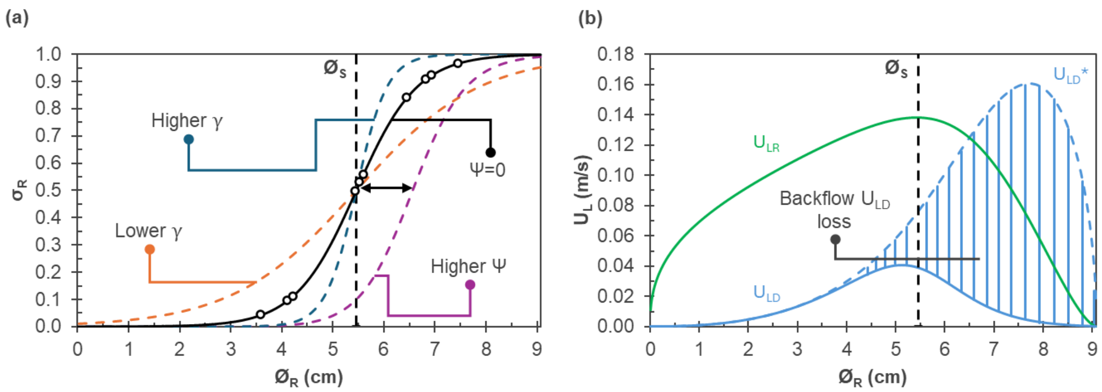

Figure 1 (a) Example of σ_R across different Ø_R. (b) U_{LR} and U_{LD} calculated by Equations 3 and 5.

In our analysis, the mixing time within the airlift reactor is determined by the sum of the inverses of the velocities in the riser (U_{LR}) and downcomer (U_{LD}). This method emphasizes the principle that mixing efficiency is predominantly influenced by the slower segment of flow. This directly addresses the inverse relationship between velocity and mixing time, ensuring the calculated t_m accurately reflects the collective influence of varying flow velocities on the reactor's mixing capability. This method involves adjusting the proportionality constant Ω to align with the experimental mixing time, as demonstrated in Equation 7.

$$t_m \propto \frac{1}{U_{LR}} + \frac{1}{U_{LD}} = \Omega\left(\frac{1}{[\alpha/\beta]^{0.5}} + \frac{1}{[\alpha/\beta]^{0.5}[1-\sigma_R]\,A_R/(A_T\text{-}A_R)}\right) \tag{7}$$

The tuning of the mixing time model and the experimental data is shown in Figure 2, that illustrates the application of both the existing model disregarding fluid backflow and our newly proposed model. This comparative analysis resulted in our model achieving a coefficient of determination (R^2) of 0.957. Specifically, the optimization process adjusted the proportionality constant Ω to 1.7457, the backflow sensitivity factor γ to 6.051, and set the parameter ψ to zero, with the friction factors for both the riser and downcomer optimized to lie within the range of 0.05 to 0.35. The adjustment of ψ to zero suggests that the system's critical diameter for fluid dynamics analysis coincides with that of the sparger, highlighting the significance of the sparger's dimensions in influencing fluid movement. Additionally, gas holdup in the riser (ε_R), was preliminarily estimated to fall between 0.25 and 0.4 for various internal tube configurations, highlighting the influence of tube geometry on gas holdup levels. These adjustments and findings emphasize the robustness and adaptability of the proposed model in accurately simulating the fluid dynamics and mixing efficiency of airlift reactors under varying operational conditions. Thus, the flow model analysis reveals that the optimal internal tube diameter, which minimizes mixing time under the examined conditions, corresponds to 93.9% of $Ø_S$, equivalent to an $Ø_R$ of 5.31 cm (A_D/A_R of 2.133).

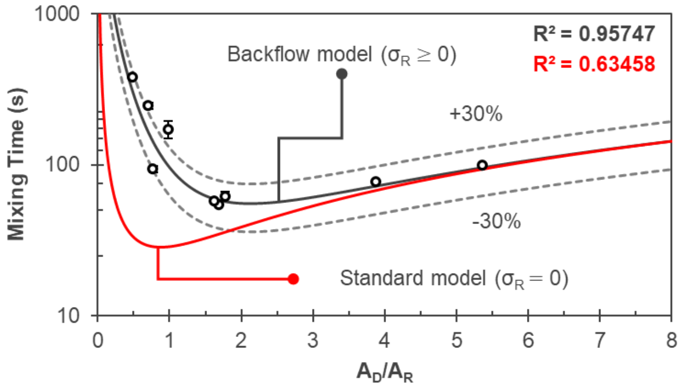

Figure 2 Mixing time model as delineated in Equation 7, plotted against the ratio of downcomer to riser area (A_D/A_R).

4 CONCLUSION

This investigation introduces a nuanced model that elucidates the flow dynamics within airlift reactors in viscous three-phase systems, specifically addressing the downcomer and riser sections and effectively integrating backflow as a pivotal component. The results highlight backflow's critical influence on diminishing downcomer velocity, especially pronounced with the enlargement of riser diameters. This is particularly relevant for configurations with low $H_R/Ø_R$ ratios, where space and cost considerations impact scalability. The incorporation of the backflow factor into the predictive framework significantly refines the accuracy of mixing time estimations, thus propelling advancements in airlift reactor design and operational optimization.

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