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**Diego Abreu-López, Abhishek Dutta,
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Argáez**

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MULTIPHASE FLOWS IN MATERIALS PROCESSING

Mass Transfer Study of a Batch Aluminum Degassing Ladle with Multiple Designs of Rotating Impellers

DIEGO ABREU-LÓPEZ,¹ ABHISHEK DUTTA,²
 JOSÉ LUIS CAMACHO-MARTÍNEZ,³ GERARDO TRÁPAGA-MARTÍNEZ,³
 and MARCO A. RAMÍREZ-ARGÁEZ^{1,4}

1.—Department of Metallurgical Engineering, School of Chemistry, National Autonomous University of Mexico (UNAM), Edificio “D” Circuito de los Institutos s/n, Cd. Universitaria, C.P. 04510 Mexico City, Mexico. 2.—Department of Materials Engineering, KU Leuven, Kasteelpark Arenberg 44 bus 2450, Heverlee, 3001 Leuven, Belgium. 3.—CIATEQ, Avenida del Retablo no. 150 col. Constituyentes Fovissste, C.P. 76150 Querétaro, Qro., Mexico. 4.—e-mail: marco.ramirez@unam.mx

A mathematical model has been developed to describe the mass transfer in a batch aluminum degassing ladle utilizing the rotor-injector technique. The model uses a Eulerian algorithm for the multiphase system and a dispersed renormalization group k - ε model to account for the turbulence in the system. The model was employed to evaluate the relative performance of the degassing kinetics of four impeller designs, i.e., nozzles with and without notches, and with four and five blades. The former two designs are commercially available, while the other two were proposed by the authors of this study. The results were validated against experimental measurements of the oxygen concentration in a physical model, and the flow pattern obtained was compared with image velocimetry. The model could satisfactorily predict the liquid velocity, gas holdup, vortex size and degassing kinetics, for the first time. The local kinetic parameter $k_L a$ predicted using the rigid model was found to be an important characteristic in terms of the degassing kinetics and depends on the bubble size distribution and the stirring of the liquid in the vessel. The proposed impeller designs help to distribute the bubbles more uniformly throughout the ladle, increasing momentum transfer from the impeller to liquid and thus improving gas-liquid mass transfer.

INTRODUCTION

The rotor-injector technique in aluminum degassing is one of the most widely used metallurgical processes for gas purging to remove hydrogen dissolved in the molten metal to obtain high-quality melt for casting.^{1,2} Hydrogen is the only gas that is soluble in molten aluminum, so control of the dissolved hydrogen in the melt is of major concern in foundries. Degassing is an important refining operation that can be achieved by injecting gases through lances, porous plugs, and rotating impellers. These impellers, driven by mechanical energy, stir the melt vigorously and create a large number of small gas bubbles, which, while ascending through the molten aluminum, absorb, carry, and eventually remove hydrogen and other impurities

from the melt inside the ladle. Efficient interaction between the gas and liquid phase is possible due to the capability of mixing the vessel content and dispersing gas bubbles into molten metal promoted by the gas flow rate, angular velocity of the impeller, and rotor design. This interaction determines the degree of degasification, depending on the mass transfer characteristics. According to Zhang et al.,³ different models have been proposed to predict the degassing of aluminum. The simplest such models are based on the mass balance of dissolved hydrogen, where it is assumed that degassing is controlled by the mass transfer at the liquid side of the gas-liquid interface, but the total interfacial area is required for these models. Thus, an overall volumetric mass transfer coefficient $k_L a$ can be obtained from this model and is the parameter typically

reported. However, $k_L a$ is a combination of two parameters, namely the mass transfer coefficient k_L and interfacial area concentration a between dispersed gas bubbles and molten metal. These are the two principal kinetic parameters of the gas–liquid interaction for degassing, being functions of the local fluid flow pattern, turbulent dispersion, and gas holdup distribution in the vessel, although the whole process is also influenced by the rotor-injector impeller design.^{3,4} Saternus et al.^{5,6} studied the gas bubble dispersion and mixing characteristics with different impeller designs using physical modeling and remarked on the importance of testing impeller designs in the laboratory before industrial testing. Several experimental studies that are useful to obtain fit correlations based on mass balances, process characteristics, and geometrical configurations to determine the global $k_L a$ in similar stirred systems can be found in literature.^{7–10} Such mass transfer coefficients, used for characterizing phase resistances, are usually calculated by means of empirical correlations. However, Gimbun et al.¹¹ and Zhang³ advised caution regarding use of these kinds of correlations, because they are only valid for the original system studied. Given that these correlations are not based on physical principles to understand the phenomena, they have a limited window of applicability and cannot be applied to any other system or scaled-up vessel.

There are several empirical models to predict local mass transfer coefficients for multiphasic systems where the bubble diameter is known. Classical mass transfer models include the two-film model,¹² penetration theory model,¹³ surface renewal model,¹⁴ film penetration model,¹⁵ and eddy cell model.¹⁶ The latter proposes that the surface renewal rate can be estimated from the contact time between gas and liquid depending on turbulent large eddies. Calderbank¹⁷ assumed a moving interface where liquid flow relative to the bubble controls the surface renewal and the contact time is calculated from the slip velocity between gas and liquid. Frossling¹⁸ proposed a model based on laminar boundary value theory, which Alves et al.⁹ modified by considering a rigid bubble with a static interface.

However, industrial measurements in situ and adequate prediction of kinetic parameters remain very challenging. In light of this, use of a numerical tool such as computational fluid dynamics (CFD) with appropriate validation through a physical model is a useful approach to predict degassing kinetics. Also, fluid dynamics can be validated against measures of fluid flow characteristics, commonly by applying the particle image velocimetry (PIV) technique, as discussed by several authors.^{19–22} CFD has become a powerful tool for research and development in multiphase flow systems, since detailed computational information about flows can now be obtained at a fraction of the cost of corresponding experiments.²³ CFD

models offer the advantage of predicting local values of the mass transfer coefficient and area concentration based on fundamental theories. Recently, some authors^{11,24,25} have reported predictions of local values of k_L and a using CFD in stirred tanks for water–oxygen systems, but in the majority of these cases, the outlet used was a flat free surface with degassing boundary condition instead of using the shape of the vortex. For the degassing ladle investigated herein, such use of a flat free surface as a boundary condition would be far from reality.²⁶

In the present study, degassing kinetics was predicted by implementing a gas–liquid mass transfer model using a numerical approach through CFD. The results obtained were then validated against experimental measurements of oxygen concentration in a physical model, and the flow pattern obtained with PIV. The model was used to evaluate the relative performance of the degassing kinetics of four impeller designs, two of them commercially available and the other two proposed by the authors.

GAS-LIQUID MASS TRANSFER

Mass transfer of species A from one phase (say liquid) to another phase (say gas) involves the following sequential steps:

1. The solute is transported by convection and diffusion from the bulk of the liquid phase to the gas–liquid interface.
2. The solute is picked up or absorbed by the gas phase at the gas–liquid interface.
3. The absorbed solute is transported from the interface into the bulk of the gas.

The term “interface” refers to the geometrical plane or surface of contact between two phases. The concentration of A should be larger in the bulk of the liquid than at the gas side of the interface, thereby providing a positive driving force for mass transfer toward the gas. Similarly, the concentration of A should decrease from the gas side of the interface to the bulk of the gas, ensuring a favorable driving force in the gas phase as well. It is important to note that steps 1–3 are facilitated by turbulence in fluid media. As the time scale of attainment of interfacial equilibrium is just a fraction of a second, the assumption of interfacial equilibrium is valid in most situations. However, there are cases in which equilibrium does not exist at the interface and interfacial resistance occurs. Since interfacial resistance is negligible in most cases, resistances to mass transfer are caused only by the two phases in contact. Lewis and Whitman¹² visualized that two stagnant fluid films exist on either side of the interface and that mass transfer occurs through these films, in sequence, by purely molecular diffusion. Beyond these films, the concentrations in each phase are equal to the bulk

concentrations. That is, convection occurs in the bulk of the liquid phase, but as the solute reaches a thin layer close to the bubble, diffusion controls the mass transfer, at which point classic mass transfer theory becomes important.^{12–15} Under convective conditions, a functional relationship for mass transfer exists, which can be written as $Sh = f(Re, Sc)$. The two principle dimensionless groups of relevance to mass transfer are the Sherwood and Schmidt numbers. The Sherwood number can be viewed as describing the ratio of convective to diffusive transport. As suggested by Froosling,¹⁸ for rigid spheres and for high Reynolds numbers, the local Sherwood number can be expressed as

$$Sh = 0.6Re^{1/2}Sc^{1/3} \quad (1)$$

Using this equation, the dimensionless number can be substituted by its definition to obtain a mass transfer rigid model for computation of the local mass transfer coefficient k_L^{rigid} (see Eq. 7 later). This model assumes that the bubbles are rigid, i.e., that a laminar profile is developed around the spheres. The mass transfer coefficients in agitated systems, as considered herein, can be correlated by the combination of the above-mentioned Sherwood, Reynolds, and Schmidt numbers. Using the proposed correlation (1), it is possible to correlate a host of experimental data for a wide range of operations. The coefficients of the relation also can be determined from experiment.^{27,28}

MATERIALS AND METHODS

Physical Model Development

An acrylic full-scale physical model of an industrial prototype was used to simulate removal of gaseous hydrogen from aluminum by measuring the deoxidation kinetics of water. According to Tzanakis et al.,²⁹ deoxidation of water is similar to degassing of aluminum, therefore representing an acceptable approximation. Although it is impossible to satisfy the dynamic similarity criteria for both the Reynolds and modified Froude number in aluminum using a water model, since the flow is dominated by inertial forces more than by the drag of the ascending bubbles, it is valid to scale the model using only the Reynolds number.²⁰ Table I presents the dimensions of both the physical model and industrial prototype, the physical properties of the fluids, as well as the kinematic and dynamic similarity criteria. Figure 1a–h depicts the four impellers analyzed in this study, seen from different sides, namely two commercial designs (impellers A and B with similar designs, except that impeller B has notches, Fig. 1a, b) and two further designs previously proposed, which do not have nozzles but blades, four in the case of design C and five for design D^{19,20,30} (Fig. 1c, d). The aim of this study is to select an impeller that degasses faster than the commercial designs. The gas–liquid hydrodynamics was determined using PIV equipment (Dantec

Dynamics®) with a dual-power Litron LDY302 laser (527 nm wavelength) at 65% of its power and a high-speed camera (Speedsense M320, 2MP Ametec) with a 550-nm filter to avoid capturing reflections from the bubbles, absorbing only light emitted by seeded 20- μm polyamide particles painted with Rhodamide B. Longitudinal planes at the center of the vessel were measured as presented by Mancilla et al.,²⁰ but additionally in this study, four transversal planes were analyzed by PIV at heights shown in Fig. 1i. A total of 700 photographs were taken in single-frame mode, and the images were statistically processed with the aid of Dantec Dynamics Studio v4.2 software. A detailed description of the measurement technique can be found in the works of Camacho et al.¹⁹ and Mancilla et al.²⁰ The deoxidation kinetics was measured by saturating with dissolved oxygen (DO) at around 6 ppm in water and eliminating it by purging nitrogen gas ($\sim 99.99\%$ industrial purity) at a rate of 10 L/min while operating at 400 rpm; these conditions correspond to a dispersed gas regime in the similar system. An oximeter (Hanna Instruments HI 98194, USA) was used to measure the evolution of the DO during the degassing operation, and all the impellers were tested in degassing processes carried out in triplicate. The experimental results for DO were normalized and fit using the following first-order decay equation with TableCurve 2DTM v.01 software:

$$C^* = A + Be^{-Ct} \quad (2)$$

where C^* is the dimensionless oxygen concentration, t is time in seconds, C is a constant that includes the global kinetic parameter $k_L a$, and the constants A and B are parameters of the kinetic equation model.

Numerical Model Formulation

Assumptions

The main assumptions for the mathematical model are as follows:

- An isothermal system without thermal gradients is used, since this kind of vessel promotes homogeneous baths.
- The physical properties of gas and liquid are constant; fluids are incompressible and Newtonian.
- Liquid and gas phases are considered to be continuous and dispersed, respectively.
- Gas bubbles are assumed to be rigid and spherical with constant 1-mm diameter.
- All walls are considered to be impermeable and no slip, and a standard wall function is used to connect the laminar region to the turbulent core of fluid near to the walls.
- Turbulence is taken into consideration only in the continuous liquid phase, represented by the dispersed renormalization group (RNG) $k-\epsilon$ model.

Table I. Dimensionless numbers, dimensions, and physical properties of the physical model and prototype

Parameter	Prototype	Physical model
Height of liquid	0.5 m	0.5 m
Ladle diameter	0.5 m	0.5 m
Impeller diameter	0.166 m	0.166 m
Impeller height from ladle bottom	0.166 m	0.166 m
Reynolds number (Re)	185,260 (220 rpm)	185,260 (400 rpm)
Modified Froude number (Fr')	0.0001582	0.0009263
Density of liquid	2345 kg/cm ³ (at 800°C)	1000 kg/m ³
Viscosity of liquid	0.0013 kg/m s (at 800°C)	0.001 kg/m s
Density of gas	1.78 kg/m ³	1.138 kg/m ³
Viscosity of gas	2.125×10^{-5} kg/m s	1.663×10^{-5} kg/m s
Gas diffusivity	$D_{H-Al} = 2.7 \times 10^{-7}$ m ² /s	$D_{O_2-water} = 2.10 \times 10^{-9}$ m ² /s

Fig. 1. Impellers studied in this work: (a, e) commercial design (design A), (b, f) commercial design with notches (design B), (c, g) proposed design with four blades (design C), and (d, h) proposed design with five blades (design D), in side (a–d) and bottom view (e–h); (i) Schematic of system and transversal planes with heights H1, H2, H3, and H4 captured by PIV.

Fluid Flow and Turbulence Model

The numerical model for the gas–liquid hydrodynamics was built under a Eulerian frame of reference using the multiple reference frame (MRF) approach to represent the rotating parts of the system (shaft and impeller) together with the nonrotating part (fluid domain of degassing unit).

The multiphase flow considers Newtonian and incompressible fluids; steady state and turbulence were taken into account only in the liquid phase. To describe the hydrodynamics, a complete set of conservation equations for each phase was stated, including the continuity and Navier–Stokes equations. Turbulence in the liquid phase was described through the dispersed RNG $k-\epsilon$ model, which is

deemed appropriate for such swirling flows. The RNG model was developed using the RNG method by Yakhot et al.³¹ to renormalize the Navier–Stokes equations and account for the effects of smaller scales of motion. This approach is similar to the standard $k-\varepsilon$ model, but since it was developed from a theoretical basis, model constants are adjusted for energy dissipation rate equations. It was also selected because it gives satisfactory predictions of hydrodynamics and turbulence quantities in stirred vessels.^{30,32–34} The boundary conditions represented the impermeable and no-slip conditions at the moving and static walls and the standard wall function treatment of the turbulence; also, a gas inlet was set at the impeller nozzle, and an opening in the atmosphere boundary allowed gas to leave the domain. Details of the hydrodynamic model including drag forces, turbulence production due to bubbles, and other important source terms in the governing equations, as well as a complete set of boundary conditions, can be found in the study of Abreu-López et al.³⁰

Species Model for Dissolved Oxygen

Dissolved oxygen in water is transferred into gas bubbles at the gas–liquid interface. Representation of this process requires solution of the conservation equations of DO in the liquid and gas phase, respectively, as mentioned below:

$$\begin{aligned} \frac{\partial}{\partial t} (\alpha_l \rho_l w_{O_2,l}) + \nabla \cdot (\alpha_l \rho_l \bar{V}_l w_{O_2,l}) \\ = \nabla \cdot \alpha_l \left(\rho_l D_{O_2,l} + \frac{\mu_t}{Sc_t} \right) \nabla w_{O_2,l} + R_{O_2,l} \end{aligned} \quad (3)$$

$$\begin{aligned} \frac{\partial}{\partial t} (\alpha_g \rho_g w_{O_2,g}) + \nabla \cdot (\alpha_g \rho_g \bar{V}_g w_{O_2,g}) \\ = \nabla \cdot \alpha_g \rho_g D_{O_2,g} \nabla w_{O_2,g} + R_{O_2,g} \end{aligned} \quad (4)$$

where α_i and ρ_i are the volume fraction and density of phase i , w_{O_2} and D_{O_2} are the mass fraction and diffusion coefficient of oxygen, while the subscripts “l” and “g” stand for liquid and gas, and Sc_t is the turbulent Schmidt number with a default value of 0.7. The source terms representing the mass of oxygen transferred from the liquid to the gas, $R_{O_2,l}$ and $R_{O_2,g}$, having the same value but different sign, are as follows:

$$R_{O_2,l} = -R_{O_2,g} = k_L a (\rho_{O_2,l}^* - \rho_{O_2,l}) \quad (5)$$

where $\rho_{O_2,l}^*$ and $\rho_{O_2,l}$ are the final and instantaneous oxygen concentration (kg/m^3), k_L is the mass transfer coefficient, and finally, a is the interfacial area concentration, defined as

$$a = \frac{6\alpha_g}{d_g} \quad (6)$$

where d_g is the bubble diameter.

In this study, several approaches to compute k_L were used,³⁵ but the best agreement between measured and predicted kinetics was obtained when using the rigid model proposed by Frossling.¹⁸ By replacing the Reynolds and Schmidt dimensionless numbers from Eq. 1 and clearing mass transfer coefficient, the rigid model is computed as

$$k_L^{\text{rigid}} = c \sqrt{\frac{|\bar{V}_l - \bar{V}_g|}{d_g}} D_{O_2,l}^{2/3} v_l^{-1/6} \quad (7)$$

where $|\bar{V}_l - \bar{V}_g|$ and v_l are the slip velocity and liquid kinematic viscosity, respectively, and c is a model constant with value of 0.6, although numerical values of this constant between 0.42 and 0.95 can be found.^{27,28} The numerical model was solved in the commercial fluid dynamics software ANSYS Fluent v18.2. The mass transfer computation was done locally in each cell using the rigid model through a user-defined function (UDF).

RESULTS AND DISCUSSION

Model Validation

Figure 2 compares the experimental (top) and predicted (bottom) gas distribution in a central longitudinal r – z plane for impellers A, B, C, and D (from left to right in the figure), for a degassing operation at 10 L/min and 400 rpm. Clearly, the model is in excellent agreement with the experimental results for all the impellers, since the real bubble distribution is well represented by the model in each case. Even features such as the vortex size and the large bubble produced at the bottom of the impeller are well captured by the model, and their sizes for each impeller are properly predicted. In terms of process analysis, impeller C, with four blades, forms the largest bubble at the bottom, creates the biggest vortex, and promotes the best bubble dispersion, closely followed by impeller D, and these characteristics are quantitatively far from those of impellers A and B. Impeller C exhibited the strongest pumping effect, arising as a pressure drop at the bottom of the impeller to counterbalance centrifugal forces. This pumping effect explains both the formation of the bubble at the bottom part of the impeller and the vortex size; the greater the pumping effect, the larger the vortex and the size of the bubble at the bottom. The dispersion of the bubbles is explained by the ability of the impeller to eject liquid radially, which is called the discharge of the impeller, and is closely related to the pumping effect and the momentum transfer from the impeller to the liquid. As seen in Fig. 2, designs A and B did not promote the dispersion of bubbles and pumping effect as much as designs C and D, thus the latter designs show better performance.

The measured (top) and predicted (bottom) streamlines of liquid at the longitudinal r – z plane

Fig. 2. Comparison between (a–d) experimental and (e–h) predicted gas holdup in an r – z plane with impellers operating at 400 rpm and 10 L/min, for impellers A (a, e), B (b, f), C (c, g), and D (d, h).

for rotors A, B, C, and D are shown from left to right in Fig. 3. The agreement between experiment and simulation is reasonably good, with measurements in the lower part of the vessel describing in all cases a clockwise loop below the impeller height. In the upper part of the vessel, however, the agreement is not so good, and both the measurements and simulations may be failing due to the presence of the vortex, which causes instabilities and reflections, making it difficult to acquire PIV data. The simulations predict a counterclockwise loop above the line of the impeller, not seen in the measurements (except for impeller C). These uncertainties in the upper part of the vessel do not detract in any

way from the efficacy of the simulations, since the axial and radial components of the liquid velocity in that part of the vessel at the plane shown in Fig. 3 are much lower than the angular component (not seen in Fig. 3). Comparison of the measured and predicted angular velocities is thus necessary to validate the model and analyze the momentum transferred from the impeller to the liquid (Fig. 4). Figure 4 shows radial profiles of the liquid velocity magnitude at the four heights indicated in Fig. 1i for all the impellers. Only at H1 is the profile complete, since this height is below the impeller, whereas for the other heights above the impeller, it blocks the view of the camera so that the profiles are

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Fig. 3. Comparison between streamlines experimentally obtained by PIV (a–d) and numerically predicted (e–h) in an r - z plane for impeller designs A (a, e), B (b, f), C (c, g), and D (d, h).

incomplete. The agreement is excellent for all heights for impellers A and B, but just good for the impellers with blades (C and D). This may be due to the nature of the mesh used in the MRF approach, which consists of a structured grid in the nonrotating zone connected through an interface with an unstructured grid in the rotating part.³⁰ Blades transfer momentum to the fluid in designs C and D, while notches and nozzles do so in designs A and B. The blades are farther from the nonrotating zone than the notches and nozzles are, so numerically momentum transfer from the rotating to nonrotating part is easier in designs A and B than in designs C and D. Besides, the unstructured grids for the designs with blades (designs C and D) represent more fluid volume

with low-quality tetrahedral grids due to the particular shape of the blades and the conical section of these rotors, which further explains the differences observed between the measurements and simulations for designs C and D. Quantitatively, the unstructured grids represent about 15% more fluid volume for the designs with blades (designs C and D) than for those without blades (designs A and B). The percentage of fluid volume in the rotating zone is 54.9%, 54.7%, 70.4%, and 68.0% for design A, B, C, and D, respectively. Moreover, note that tetrahedral grids require approximately twice the storage space and CPU time compared with hexahedral tessellations of the same vertices because of the greater number of edges, although these additional edges do

Fig. 4. Radial profiles of liquid velocity magnitude for impeller designs A, B, C, and D at different heights H1, H2, H3, and H4 (see Fig. 1i).

not contribute significantly to improvement in solution accuracy.³⁶

From this analysis, it is clear that the impellers with blades transfer momentum more efficiently to the liquid. The maximum liquid velocities of 1.6 m/s reached at the impeller height of impellers C and D are comparatively higher than those of 1.0 m/s reached with impellers A and B. It can be thus concluded that, at the same rotational speed of 400 rpm, impellers C and D transfer momentum to the liquid more effectively and consequently promote better-stirred baths compared with impellers A and B.

The values of the kinetic parameter $k_L a$ predicted by the rigid model for impellers A, B, C, and D are shown in Fig. 5. The scales are kept exactly the same for comparison purposes. The local kinetic

parameter contours are similar in shape to the gas fraction contours presented in Fig. 2. Clearly, this parameter is the most important characteristic in terms of degassing kinetics, but it depends closely on the bubble distribution due to the surface area concentration for mass exchange (a), and also depends on the stirring of the liquid to promote convection in the liquid and consequently a fast mass transfer rate (k_L). Therefore, from the previous analysis of Figs. 2, 3, and 4, it becomes evident that the impellers with blades (C and D) promote better bubble distribution, better stirring, a stronger pumping effect, and more turbulence (for example, the turbulent dispersion for each impeller can be seen in the study of Abreu-López et al.³⁰). Thus, they present more homogeneous and higher mass

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Fig. 5. Contours of the local kinetic parameter $k_L^{\text{rigid}} a$ in an r - z plane obtained with the rigid model of mass transfer for impeller design: (a) A, (b) B, (c) C, and (d) D.

transfer coefficients throughout the ladle compared with impellers A and B.

Finally, Fig. 6 shows the predicted (lines) and measured (symbols) dimensionless DO evolution for all the impellers tested in this study. As noted, the agreement between the simulated and predicted mass transfer is excellent in all cases. Since the final steady-state DO concentration at the end of the process, i.e., at 1400 s, is almost the same for all the impellers, it can be concluded that the rotor design is not as significant a variable as the gas flow rate or rotation speed, as also observed by Hernandez-Hernandez et al.¹⁰ Furthermore, at very long times, when the liquid is depleted of dissolved gas, the degassing kinetics is no longer controlled by mass transport processes but by so-called chemical control³⁷; thus at very long times, the degassing kinetics is similar regardless of the rotor design. Consequently, the performance of each rotor can be distinguished only at short and medium times by the slope of the degassing curves obtained for each rotor.

The analysis above (Figs. 2, 3, 4, 5, 6) confirms that impeller C (four blades) promotes the fastest degassing kinetics, closely followed by impeller D (five blades), then impeller B and impeller A, with the worst performance. This can be confirmed by calculating the global kinetic parameter $k_L a$ according to Eq. 2 for the results presented in Fig. 6. The numerical values of the global $k_L a$ for the experiments and predicted results are presented in Table II. A numerical comparison of the improvement of using design C instead of each of the other impellers is also reported. The values of the global $k_L a$ obtained from the predictions and experiments show good agreement, with error of less than 6.5%.

Fig. 6. Comparison of deoxidation kinetic curves obtained experimentally and theoretically for all impellers at 400 rpm and 10 L/min.

The results in Table II show that the most relevant improvement for degassing kinetics occurred when replacing impeller A with impeller C, achieving an improvement of 24% in the global mass transfer coefficient, while replacing impeller B with impeller C resulted in an improvement of 12.5%. Regarding the difference in performance between designs D and C, the latter showed slightly better performance by 1.6%.

Although all the impellers operate under the same conditions (10 L/min and 400 rpm), the designs with blades (C and D) increased the pumping effect, transfer more momentum to the liquid stirring, and improve the mixing, creating a more turbulent flow and dispersing more gas over the entire degassing unit, in comparison with the

Table II. Global kinetic parameter k_{La} obtained by fitting degassing data to Eq. 2 for experiments and simulations

Impeller design	Experimental k_{La} (s^{-1})	Simulated k_{La} (s^{-1})	Difference between numerical and experimental results (%)	Improvement in global k_{La} values when using design C with respect to each other design (%)
A	$0.002738 \pm 2 \times 10^{-5}$	0.002707	1.1	24.0
B	$0.003151 \pm 1 \times 10^{-5}$	0.003066	2.7	12.5
C	$0.003601 \pm 4 \times 10^{-5}$	0.003836	6.5	—
D	$0.003544 \pm 5 \times 10^{-5}$	0.003752	5.9	1.6

impellers with nozzles (designs A and B). This is because the momentum transfer from the impeller to the liquid is better. Nevertheless, it must be noted that the present study involves the assumption of a uniform bubble size in the momentum exchange mechanism of the interphase drag coefficient, which directly affects the prediction of the local mean velocities and gas holdup.

CONCLUSION

A numerical model was developed for a batch aluminum degassing ladle, being able to predict satisfactorily the liquid velocity, gas holdup, vortex size, pumping effect, mass transfer coefficient, and kinetics for four different impeller designs. The model was validated using a physical model with dissolved oxygen for the degassing kinetics. Impeller designs with blades, in particular that with four blades, were found to have superior performance compared with commercially designed impellers with nozzles, since those with blades promote turbulence-induced stirring, a more homogeneous gas distribution, a greater pumping effect, and more efficient momentum transfer compared with nozzle designs. In terms of overall performance, the designs tested in this study at the same operating conditions can thus be ranked, from best to worst, as follows: C (four blades), D (five blades), B (nozzles with notches), and A (nozzles).

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