Modeling of Mixing in Ladles Fitted with Dual Plugs

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A physical and mathematical modeling study has been carried out to investigate mixing in a gas stirred ladle fitted with dual plugs, located diametrically opposite at \( \pm \frac{1}{2} R \) positions. While conductivity measurement technique was applied to record 95% mixing times, mathematical modeling was carried out via the commercial CFD package FLUENT® wherein, a two phase flow calculation procedure based on discrete phase approach was adapted. It was demonstrated that numerically predicted flow and mixing times in general agree reasonably well with the corresponding experimental measurements.

In addition to the above, a relatively simple, quasi single phase flow calculation procedure, developed in-house, was also applied to predict mixing times and thereby, assess an earlier work. It is shown that the quasi single phase model, despite its simplicity, is reasonably effective in simulating mixing phenomena in such system. A comparison between different modeling approaches vis a vis experimental measurements is also illustrated in the text.

KEY WORDS: dual plug stirred ladles; flow; mixing; discrete phase modeling; quasi single phase modeling; experimental measurements; comparison.

1. Introduction

The efficiency of many chemical processing operations carried out in the present day steelmaking ladles are intricately related to mixing phenomena. Mixing enhances chemical reactions by bringing in reactants together and removing products from reaction sites. In addition, it also influences the extent of thermal and particulate in-homogeneities within the ladle. It is therefore desirable to ascertain the extent of mixing, in order to evaluate the process performance of Argon or Nitrogen stirred ladles. Justifiably therefore, the study of mixing in gas stirred ladle system has received much attention over the years. Accordingly, numerous physical and mathematical modeling investigations have so far been carried out and reported in the literature.\(^1\)

Mixing is essentially a “convection+diffusion phenomenon” and therefore, elaborate flow calculation is a prerequisite to the modeling of mixing. To this end, a wide variety of computational procedures have so far been applied and these have accordingly resulted in to: quasi single phase,\(^2,3\) Eulerian two phase\(^4,5\) and Eulerian-Lagrangian two phase\(^6,7\) (also known as the discrete phase modeling approach) models respectively. Multiphase flow modeling embodying the VOF (Volume of Fluid) technique has also been applied recently\(^8\) to model flow and mixing in a slag covered ladle. These\(^2-8\) as well as a vast majority of the model studies reported in the literature,\(^9,10\) as one would note here, were primarily restricted to axisymmetrical and/or asymmetrical gas injection configuration in which, a single nozzle/plug located at the base of the vessel was used to supply the gas. As such, not much information on fluid flow and mixing on dual plug/nozzle stirred ladles is available in the literature.\(^9,10\)

Nearly a decade back, Joo and Guthrie\(^9\) reported a computational study of mixing in a dual plug stirred ladle embodying the popular quasi single phase calculation procedure. In their study, drastic simplifications were made and concepts in general applicable to axisymmetrical gas stirred systems were applied. Despite such, it was demonstrated\(^9\) that quasi single phase model produces results that are in reasonable agreement with experimental measurements. Barring this,\(^9\) not much information on fluid flow and mixing phenomena in dual plug stirred ladles is available in the literature. It is to be emphasized here that stirring in ladles with two porous plugs is relevant and is becoming increasingly popular particularly for relatively bigger vessels, in order to achieve gentle but rapid mixing as well as to promote better slag/metal intermixing, and to avoid explosive degassing effects under vacuum. Moreover, mathematical modeling of liquid steel processing operations are currently being increasingly carried out embodying commercial CFD packages. In such a context, it is important to report new experimental results as these could be used to validate the CFD codes. Given such, the relevance of the present work is readily apparent.

Consequently, the purpose of the present study has been to carry out a physical and mathematical modeling investigation on mixing phenomena in gas stirred ladles fitted with dual porous plugs. Towards this, details of the computational and experimental procedures together with results derived from a 0.20 scale water model of a 210T refining
ladle are reported in the subsequent sections.

2. Present Work

In the present study, mathematical modeling was carried out adopting two conceptually different approaches namely; the discrete phase and the quasi single phase procedures; the former via the commercial CFD package FLUENT® and the latter via a three dimensional, turbulent flow calculation procedure developed in house. Parallel to these, experiments were carried out on a 0.20 scale water model of a 210 T industrial ladle and 95% mixing times were measured over a wide range of operating conditions (viz., liquid depth, gas flow rate etc.). In this section, a summary of the computational and experimental procedures is presented.

2.1. Experimental

Mixing times were measured in a cylindrical vessels (I.D. = 0.60 m) in which, water was agitated by injecting air or \( N_2 \), in 1 : 1 proportion, through a pair of nozzles located at the bottom of the vessel at the mid bath radius position. Prior to monitoring mixing, air/\( N_2 \) was bubbled into the water bath at the desired flow rate for a few minutes to ensure the stability of the flow in the vessel as well as to remove any inhomogeneities present in the bath. The gas flow rates were so chosen to ensure gentle stirring condition as are typically encountered in actual ladle metallurgy steel-making operations (e.g., ~0.001 to 0.015 Nm\(^3\)/t·min). While lower end flow rates are typically used for bath homogenization (i.e., of thermal and/or material), relatively large gas flow rates are used to achieve enhanced slag–metal reactions. However, as the primary objective of this study was to ascertain the general adequacy of mixing models, consequently it was deliberately decided to employ only a small fraction of the above mentioned range of gas flow rates, such that some meaningful experimental data, required for validation of the mathematical models can be derived.

A conductivity probe supplied with a digital conductivity meter (Eutech make, CYBERSCAN 200) was employed to record changes in the local ion concentration of a pulse tracer (NaCl or \( H_2SO_4 \)) added directly to a point on the liquid free surface lying midway between the ‘eyes’ of the two surfacing plumes (e.g., the axis of symmetry). The change in local ion concentration around the probe tip was measured through the changes in water’s electrical conductivity and recorded manually via the digital conductivity meter. Thus, data on conductivity were collected typically every 2/3 s and simultaneously stored in a computer using the Cyber Comm Portable software®. On addition of tracer, considerable oscillations in the conductivity value were observed which essentially resulted from the periodic fluctuations of the amount of tracer passing through the probe tip. For each experiment, the recording of tracer response was carried out until the concentration of the tracer in the bath was considered to have reached a homogeneously mixed value. Experimental data were subsequently analyzed through MS-Excel® software and a plot between conductivity vs. time was produced from which, corresponding mixing times were estimated. These are defined in the present context as the time required for the monitoring point concentration to fall continuously within a 5% deviation of the well mixed/homogeneous value. By keeping the probe tip immersed in the slowly moving region of the vessel (i.e., near the junction of the bottom and side walls lying on the principal diameter of the vessel; see also later), measured mixing times were interpreted as the bulk mixing times. The present approach therefore ensured that by the time the monitoring point reached a degree of 95% homogeneity, the bulk of the liquid was practically homogeneous. A set of at least five measurements were made for each experimental condition and thereby, an average mixing time was determined. The maximum variation in successive measurements was found to be always less than 15%.

2.2. Computational

Numerical computation of flow and mixing in the dual plug stirred ladle was primarily carried out adapting a discrete phase modelling approach. In addition, a three dimensional, turbulent flow model developed in house (e.g., Advanced CFD models) was also applied. This latter approach, was considered in this study to assess the adequacy of the relatively simple quasi single phase calculation procedure vis-à-vis advanced CFD models and thereby, make an assessment of a study reported earlier. A brief summary of the two calculation procedures relevant to the present investigation is presented below.

2.2.1. Discrete Phase Modeling

This procedure in essence is similar to a combined Lagrangian–Eulerian calculation procedure adapted many a times to compute trajectories of bubbles/particles in a gas stirred ladle system. In the present work, an unsteady, three dimensional turbulent flow calculation scheme was first configured and the liquid or the continuous phase continuity and momentum conservation equations solved in conjunction with three different bubble trajectory equations (e.g., force balance expressions along three coordinate directions) in a mutually coupled fashion.

To model turbulence within the flow system as well as to track the discrete phase (viz., the bubbles) in a stochastic manner, the standard coefficient \( k-e \) turbulence model was incorporated in the calculation scheme. In the beginning of a typical calculation, bubbles of known diameter (estimated from an appropriate correlation; see later) was introduced in to the domain through the two plugs/nozzle and their position tracked as a function of time till the gas bubbles reached free surface where these were assumed to have escaped from the system. A large number of such bubble trajectories were typically computed. Subsequent to this, the continuous phase momentum balance equations were solved for a pre-determined number of iterations. The continuous phase and the discrete phase equations were solved sequentially repeatedly and the calculation was terminated once both the continuous phase momentum balance and the trajectory equations simultaneously reached convergence. A threshold scaled residual value on each variable, which in the present work was less than equal to \( 10^{-4} \), was applied to monitor convergence.

The flow, turbulence model and the trajectory equations were solved first to get to a steady state velocity field. To this end, the volume average velocity within the flow domain was computationally monitored and this served as an
indicator of the flow stabilisation in the system. Following convergence of flow, discrete phase and turbulence model equations, the species transport model was enabled to simulate the tracer dispersion or mixing phenomena in the system. In this, a pulse tracer addition (e.g., for a period of 2 to 3 s) was mathematically simulated and numerical computations carried out embodying a time step size of 0.25 s, till a homogeneous concentration was recorded at the monitoring point (identical to the actual location of the measuring probe). Following such, the variation of concentration with time at the monitoring point was evaluated to estimate the associated 95% mixing time. During numerical computation, the volume average mass fraction of the tracer was carefully monitored at every 10 s time interval to check on the overall mass balance on the added tracer such that an accurate estimate of mixing time results. A schematic of the discrete phase calculation procedure is shown in Fig. 1. The following assumptions were invoked in the discrete phase model, viz.,

1. The flow system is essentially isothermal, turbulent, three dimensional and unsteady.
2. The presence of an upper buoyant phase was ignored and the bulk liquid–air interface was assumed to be flat and mobile.
3. Discrete mono size bubbles were assumed to form at the nozzle/plug tip. Thus the size of the bubbles forming at the nozzle or orifice was assumed to be known a priori from the available correlation reported in the literature, e.g.,

   \[ d_b = 0.35 \left( \frac{Q^2}{g} \right)^{0.2} \]  

   (for symbols, see nomenclature). Furthermore, the size of a bubble was estimated on the basis of the ambient flow rate and was considered to remain invariant during its rise through the liquid.
4. Bubble–bubble interactions were ignored. Standard drag law for non-spherical geometry (e.g., spherical cap) was typically applied.
5. As a first approximation, the turbulence generated within the plume by the bubbles as well as the lateral lift forces acting on the bubbles were ignored.

The initial and boundary conditions applied to solve the discrete phase model equations are illustrated schematically in Fig. 2.

2.2.2. Quasi-singe Phase Modelling

The quasi single phase model for flow calculation is
based on the continuum approach, in which, the gas liquid mixture in the upwelling plume is considered to rise like a homogeneous fluid but with a reduced density (viz., \( \rho_{\text{mix}} = \alpha \rho_g + (1-\alpha) \)). Conceptually, by embodying a spatially dependent, buoyancy force per unit volume term (= \( \rho_g \alpha \)) in the axial direction momentum balance equation of a single phase, three dimensional calculation procedure, flow phenomena in gas stirred ladle systems can be conveniently predicted. Figure 3 provides the essence of the quasi single phase modeling approach in the form of a schematic. There, as seen, key to the mathematical model is a set of four partial differential equations, the continuity, the axial direction momentum balance equation, the radial direction momentum equation and the azimuthal direction momentum balance equation respectively. However these do not form a closed set, because \( \alpha \), the gas volume fraction and the effective viscosity, \( \mu_{\text{eff}} \) require prior specification. The gas volume fraction, \( \alpha \), as pointed out already, appears in the axial direction momentum balance equation through which the free convection effect due to gas injection phenomena is incorporated while, \( \mu_{\text{eff}} \) (deduced from the \( k-\varepsilon \) turbulence model) appears in all three momentum balance equations and accommodates the effect of momentum transfer via turbulence phenomena. The relevant partial differential equations of flow, central to the quasi single phase calculation procedure, are rather well known and consequently, not re-iterated here.\(^{2,3}\) Similarly, the details of the present steady, three dimensional, turbulent flow calculation procedure is available elsewhere.\(^{12}\)

To estimate the gas volume fraction within the rising two phase plume, a knowledge of the average rise velocity of the two phase mixture is required. To this end, an algebraic model for average plume rise velocity \( U_p \) was applied. From such rise velocity and a knowledge of the physical dimension of the two phase plume region (i.e., the average radius, \( R_c \), determined experimentally through video photography), the gas volume fraction (can be readily estimated from the principle of volume continuity. To this end, two distinct possibilities were examined, namely,

\[
\alpha = \frac{Q L}{\pi R_c^2 L} \quad \text{for no slippage between bubbles and liquid} \quad (2)
\]

and

\[
\alpha = \frac{Q - \pi R_c^2 \alpha(1-\alpha) U_s}{2 \pi \int_0^R r U_p dr} \quad \text{with slip} \quad (3)
\]

in which, \( U_p \) is the average plume rise velocity and is given by:

\[
U_p = 3.1 Q^{0.33} L^{-0.25} R^{-0.38} \quad (4)
\]

It is to be mentioned here that the macroscopic plume model as given above was derived for an axisymmetrical gas stirred ladle system (viz., ladle fitted with a centrally placed plug/nozzle) and therefore, may not be applicable to the present system in a rigorous sense. It is instructive to note here that a practically identical version of the quasi single phase flow calculation procedure, derived on the basis of zero/no slip, was adapted earlier by Joo and Guthrie.\(^9\)

3. Results and Discussion

A series of experiments were carried out in the water model ladle (\( D = 0.6 \) m) in which, 95% bulk mixing times were measured as a function of gas flow rates. Since mixing time vs. gas flow rate relationships are known to be specific\(^1\) to the operating flow regimes, consequently a range of gas flow rates was applied. In Fig. 4, a log–log plot between 95% bulk mixing times and gas flow rates is presented. There, along the line of numerous previous studies, two distinct line segments have been fitted through the experimental data points. These indicate that initially, mixing time decreases rather sharply (e.g., \( \tau_{\text{mix}} \approx Q^{-0.58} \)) with in-
creasing net gas flows, up to about $2 \times 10^{-4} \text{m}^3/\text{s}$ (or 12 L/min). Thereafter, the decrease in mixing times with gas flow appears to be somewhat less pronounced ($\tau_{\text{mix}} \propto Q^{-0.35}$). In the inset of the same figure, equivalent observation$^1$ derived from axi-symmetrical gas stirred ladle systems is also shown. The striking similarities between the two set of experimental results is readily evident. It is instructive to note here that for the dual plug stirred ladle, beyond a gas flow rate of 12 L/min the exponent on $Q$ is practically equivalent to $1/3$. Based on the information documented in the literature,$^1$ flow phenomena in the dual plug stirred system beyond this critical gas flow rate (12 L/min) can be taken to be essentially dominated by the inertial and gravitational forces ($\text{viz.}$, Froude dominated). In addition to the measurements shown in Fig. 4, the time taken by the tracer to reach the probe tip (termed henceforth as the first tracer response time) was also recorded for each experiment. These together with 95% mixing time data were subsequently applied to validate mathematical model predictions.

Prior to carrying out elaborate mixing time calculations and drawing a comparison between experimental measurements and numerical prediction, it was decided to evaluate the general adequacy of the present discrete phase modelling to gas stirred ladle systems. To this end, as our first step, flow phenomena in a gas stirred ladle ($L=0.21 \text{ m}$ and $D=0.30 \text{ m}$ and $Q=0.33 \times 10^{-4} \text{m}^3/\text{s}$) fitted with a central nozzle were considered and the discrete phase model applied in both axisymmetric as well in full 3-dimensional modes to deduce the steady state velocity field in the system. Experimental data$^{14}$ on flow obtained via Particle Image Velocimetry (PIV) as well as continuous video photography were applied to evaluate the predicted results. The PIV system is described in detail in Ref.$^{15}$ The Video recording technique, on the other hand, was identical to the one adapted earlier by Mazumdar and Guthrie.$^{16}$ In Figs. 5(a) and 5(b), predicted velocity magnitude on the central vertical plane (between $r=0$ and $r=R$) is shown together with corresponding experimental measurements. For the sake of a comparison among various flow measuring techniques, previously reported experimental data$^{17}$ derived via LDV (Laser Doppler Velocimetry) have also been included in Figs. 5(a) and 5(b). On the basis of such, the following observations can be made:

- The model predictions agree reasonably well with experimental measurements in the bulk of liquid.
- The axisymmetric and the full 3-dimensional predictions are somewhat different. This might as possibility indicate that a full 3-D simulation of axisymmetric gas injection operation is able to capture some inherent asymmetric feature of a rising bubble plume. Many experimental studies have already confirmed that even with a centric gas injection nozzle, flow phenomena in such systems are often asymmetrical due to the short term and long term wandering of the bubble plumes.$^1$ It is important to note that centre-line velocities as predicted by the 2-D and 3-D versions of the models are substantially different. However, since the diameter of the plume is only a small fraction of the vessel diameter, such large difference near the axis manifest in to a much smaller variation within the main bulk of liquid. Reliable experimental data on flow within the plume is required in order to make any further inferences.
- Video-recording produces results not much different from those derived via advanced flow measuring devices. This suggests that in the absence of sophisticated flow measuring equipment, videography can be applied in water model investigations to measure flows and thereby, assess the adequacy of fluid flow models.

Fig. 4. The variation of mixing time against gas flow rate in the model ladle ($L=0.60 \text{ m}; D=0.60 \text{ m}$ and plugs at $\pm R/2$).

Fig. 5. Comparison of predicted and experimental flows at two different depths of an axisymmetric gas stirred ladle system$^2$ ($L=0.21 \text{ m}, R=0.15 \text{ m}$ and $Q=2 \text{ L/min}$ ($6.666 \times 10^{-4} \text{m}^3/\text{s}$): (a) at $z/L=0.14$ and (b) at $z/L=0.87$ respectively.
On the basis of such, video recording of flow was carried out to determine flow fields along the principal central vertical plane of the dual plug stirred ladle systems. Thus, many flow measurement exercises along the central vertical plane were carried out in our laboratory for a diverse range of nozzle configurations (e.g., \( R/H_{11006} \), \( R/2 \), etc.), gas flow rate, liquid depth etc. On the basis of such, velocity magnitude at different depths in the ladle were derived and compared with equivalent numerical predictions. As a typical example, the comparison between experimental and predicted velocity fields for \( R/H_{11006} \) plug position is illustrated in Figs. 6(a) through 6(c). There, reasonable agreement between prediction and experiment is readily evident. Each data point in Fig. 6, as one would note here, is the result of 10 discrete measurements. It is to be emphasized here that apart from the results in Figs. 5 and 6, experimental data presented in various tables and figures of the present work strictly conform to a ladle fitted with dual plugs, placed diametrically opposite at \( R/2 \) locations. Evidently, the agreement between experimental and predicted flows as illustrated in Figs. 5 and 6 provide indication that the discrete phase model is sufficiently robust and simulates flow phenomena in gas stirred ladle systems with reasonable degree of certainty.

The results presented so far were derived embodying a relatively fine grid system (average grid spacing \( \sim 0.015 \) m), an uniform bubble diameter estimated from Eq. (1), a seemingly small time step size of 0.25 s together with 60 number of continuous phase iterations per discrete phase iteration. These numerical parameters were kept identical for simulation of flow and mixing times for \( R/2 \) position of the porous plugs, as discussed in the subsequent paragraphs. However, since mixing is intricately related to fluid flow and turbulence phenomena, therefore, it was decided to computationally investigate the sensitivity of mixing time to the precise choice of various numerical parameters mentioned above. Thus, mixing time for a typical experimental condition was computed for different values of the various model (or numerical) parameters and the result thus obtained are summarised in Table 1. There, it is readily evident that predicted results within the range of values studied do not differ dappreciably. These further imply that numerical parameters applied to derive results presented in Figs. 5 and 6, produces results that are practically independent of nodal configurations, both in space and time and hence, can be conveniently applied to estimate mixing times in ladles fitted with dual plugs at \( R/2 \) locations.

In addition to the above, the influence of the amount of tracer as well the type of tracer inlet boundary condition (applied to solve the species conservation equation; see Fig. 1) on mixing times were also investigated computationally. Numerical results, like many other previous experimental studies, confirmed that the predicted 95% mixing times are independent of the amount of tracer added to the bath. Referring back to Table 1, it is instructive to note that the results presented there were derived by prescribing a fixed tracer concentration or mass fraction at the inlet plane for a predetermined period of time. Alternative to this, some calculations were also carried out by prescribing a fixed mass flow rate of tracer at the inlet plane. In this latter approach, a meaningful velocity as well as mass fraction were simultaneously prescribed at the inlet. The influence of the duration of tracer injection as well as the two types of inlet boundary condition on the predicted time vs. concentration

Table 1. Predicted 95% mixing times for different numerical, control and operating parameters (\( L=0.60 \) m, \( D=0.60 \) m, \( Q=2.5\times10^{-4} \) m\(^3\)/s (or 15 L/min) and plugs at diametrically opposite at \( R/2 \) locations).

| Parameter | Range of values of the parameters | Computed range of 95% mixing time, s | Response time, s |
|-----------|----------------------------------|--------------------------------------|----------------|
| Grid size | \( 0.015-0.020 \) m               | 31.25 to 29.75                       | 11             |
| Drag law  | Spherical vs. spherical cap       | 31.25 and 37.75                      | 11 and 13      |
| Number of continuous phase iteration | 60-100                            | 31.25-31                            | 11             |
| Overall time step | 0.25-2 sec                  | 31.25-36                            | 11-13          |
| Bubble diameter | 3 to 10 mm             | 31.25-34                            | 11-12          |
curve is illustrated in Fig. 7. There, as seen, predicted 95% mixing times are practically independent of the duration of tracer injection as well as the type of tracer inlet boundary conditions considered. It is important to note here that the equilibrium tracer concentrations in Fig. 7(a) as well as in Fig. 7(b) as predicted via different computational approaches are somewhat different. This is expected, since numerically, different amount of tracer has been introduced in to the system for each specific condition.

In Figs. 8(a) through 8(b), the predicted tracer contours on the principal central vertical plane of the dual plug stirred ladle (viz., at ±R/2 locations) are shown at four different instants. There, the first figure corresponds to an instant immediately following the pulse addition of the tracer. Qualitatively, these figures show that following the pulse addition of the tracer above the bath (i.e., along the vessel’s axis), initially the tracer tends to get caught up between the surfacing plumes. Thereafter, the tracer is advected towards the bottom of the vessel and disperses via the three dimensional, turbulent convection current prevalent in the system. It is also apparent that the region contained between the two surfacing plumes as well as the junction between the side and bottom walls are by far the slowly moving regions in the system and are therefore, likely to mix last. These computed results, it is instructive to note here, agree very closely with the observations derived from a dye tracer injection (i.e., KMnO₄) study. On the basis of such, the conductivity probe in the present work was always kept immersed on the principal central vertical plane, in the vicinity of the bottom and side wall of the ladle, such that measured mixing time

Fig. 7. Numerically predicted transient tracer concentration profiles at the monitoring location of the ladle for (a) two different duration of tracer injection and (b) two different types of tracer inlet boundary conditions.

Fig. 8. Computed mass fraction of the added inert tracer in the dual plug stirred ladle (plugs at diametrically opposite ±R/2 locations) at various instants of time: (a) at the instant of tracer addition, (b) after 10 s, (c) after 20 s and (d) after 40 s.
could be interpreted as the bulk mixing time. Clearly Figs. 8(a) through 8(d) show that in about 30–35 s time, the bath becomes practically homogeneous.

In Table 2, discrete phase model predictions are directly compared with experimental 95% mixing times for different gas flow rates. There, it is readily evident that except for the smallest gas flow rate applied, the agreement between the two is in general excellent. At the smallest gas flow rate, it is likely that the bath in not highly turbulent. However the \( k-e \) turbulence model is truly applicable to high Reynolds number flows. This together with experimental error seem to be the possible reasons for the observed discrepancy in the low gas flow regime. Interestingly, results presented in Table 2 also appear to indicate that generation of turbulence within the rising plume as well as lateral lift forces on the bubbles while are important towards accurate prediction of rise velocity and plume’s spread in gas stirred ladles, nonetheless are not critical to the prediction of mixing phenomena in such system.

As mentioned already, flow phenomena in the water model ladle \( (L=0.60 \text{ m}, D=0.60 \text{ m}) \) and plugs at diametrically opposite at \( R/2 \) locations) were also predicted via the quasi-single phase calculation procedure. In this, a \( 16 \times 20 \times 38 \) (along radial, axial and angular directions respectively) grid system was applied to numerically predict the turbulent flow phenomena. Subsequently, predicted flow and turbulence parameters were applied to solve the species conservation equation from which, 95% mixing times were estimated. As pointed out in Sec. 2.2.2, two different versions of the flow model were employed; one considering bubble slippage and the other ignoring the same \( (\text{viz., Eqs. (2) and (3))} \). These have accordingly lead to two different set of estimates, of both flow and mixing times. Results thus obtained are summarised in Table 3 in which, numerically predicted 95% mixing times (for slip as well no-slip) are shown together with the corresponding experimental measurements. In general, the predicted mixing times are somewhat longer than the experimental values. Furthermore, as seen, the predicted mixing times derived via the “no-slip” version of the quasi single phase model are somewhat shorter \( (\text{viz., mixing is faster}) \) than those deduced considering bubble slippage. This is expected since the convection current generated within the vessel is higher for no slip condition \( (\text{e.g., the associated gas voidage is relatively large}) \) than for slip. Considering the extent of experimental uncertainty and the simplistic nature of the quasi single phase model, the agreement between measurement and prediction as illustrated in Table 3, can be considered reasonable. The present work thus appears to indicate that quasi-single phase model, despite their simplistic nature, can be applied to deduce a first hand estimate of mixing times in such systems, as has been suggested earlier by Joo and Guthrie.9)

Finally, a direct comparison of the two calculation procedures \( \text{vis a vis} \) experimental measurements is illustrated in Fig. 9. There, the two phase model is clearly seen to be superior to the quasi single phase procedure. To this end, it is instructive to recognize here that despite many simplistic assumptions \( (\text{viz., mono size bubble, flat and mobile free surface etc.}) \), the two phase calculation procedure has been able simulates mixing in the system fairly accurately. From the view point of practical utility however, the scope of the model, needs to be broadened further to accommodate the presence of an upper buoyant phase (slag), since slag rather than “no-slag” is typical of industrial practice. Computational work in this direction embodying the VOF
4. Conclusions

A combined physical and mathematical modeling investigation of mixing in a dual plug stirred ladle has been carried out. From the present work, the following conclusions can be drawn:

(1) Mixing time beyond a flow rate of 12 L/min decreases nearly in proportion to a third power to gas flow rate. This suggests that flow phenomena in the system beyond the critical flow rate is essentially Froude dominated.

(2) A discrete phase, turbulent flow model has been found to simulate experimental 95% mixing times with reasonable degree of certainty. Computationally, a fine grid size, a seemingly small incremental time step size together with a spherical drag law are necessary to accurately simulate the mixing phenomena.

(3) Turbulence production within the rising plume, lateral lift forces on the bubbles etc. while are known to be of considerable importance to flow modelling in such system, have been found to have practically no bearing on the predicted mixing times.

(4) Quasi single phase models based on the concept of no slip was found to produce only reasonable results. The predictive capabilities of this class of flow model were shown to be somewhat limited in comparison to those of the relatively more advanced two phase model.

(5) The present study suggests that a quasi single flow model can be applied to derive a first hand estimate of mixing in such gas stirred systems as has been anticipated earlier.

Nomenclature

- $d_b$: Bubble diameter (m)
- $L$: Depth of liquid in the ladle (m)
- $Q$: Gas flow rate (corrected to mean height and temperature of the liquid) (m$^3$/s)
- $R$: Radius of the ladle (m)
- $R_c$: Average radius of the plume (m)
- $U_p$: Average plume velocity (m/s)
- $U_s$: Slip velocity (m/s)
- $\alpha$: Gas volume fraction in the plume
- $g$: Acceleration due to gravity (m/s$^2$)

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