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Fluid flow phenomena in bottom gas-stirred ladles with top layer: Part I. Fluid flow

By Alberto N. Conejo[*1], Shin-ya Kitamura[*2]

In order to define the optimum number and location of injection devices to optimize fluid flow in metallurgical ladles it is necessary to understand the phenomena associated with bottom gas injection such as mass transfer, mixing, the role of the top layer as well as bubble behavior. The research work carried out in the previous 35 years is reviewed. This part covers the fundamentals aspects of fluid flow phenomena involving a top layer, such as mixing time, slag emulsification due to gas injection and ladle eye formation.

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Keywords: *fluid flow, mixing time, ladle eye, mass transfer, gas injection*

1 Introduction

The development of the ladle furnace process represented a break point in the history of steelmaking. Previously, steel was produced in the basic oxygen fumace (BOF) or the electric arc furnace (EAF) combining multiple operations. With the new process, the BOF and EAF assumed a more specialized function: Melting at higher rates. The ladle furnace was converted into a specialized reactor responsible for providing the final quality and chemical composition, also at a higher rate. In order to improve the rate of reactions, the ladle furnace is equipped with stirring by gas injection or electromagnetic means. Without this tool, the ladle furnace would be inefficient, however, in order to improve flow patterns inside this reactor the mechanisms of bubble formation, bubble motion and bubble interaction with the liquid need to be fully understood. Steel opacity inhibits flow visualization, consequently, in order to get that knowledge, small scale water modeling has been employed in the past. More than 35 years of research in this field have produced a vast knowledge, however, in spite of this progress there are many unresolved issues. These issues are mainly related with scaling up to the industrial level the predictions from physical and mathematical models, involving realistic conditions. Most of the studies on mixing phenomena have either neglected the presence of the top slag layer or focused on conditions which deviate from a real process. Too much work has been concentrated on single nozzle and axisymmetric gas injection whereas the real process involves a top layer, more than one nozzle and eccentric injection.

This review has the objective to summarize the progress made on the physical and mathematical modeling of gas stirring in ladles, emphasizing its application to the industrial conditions, in particular on the optimum configuration of injection elements to improve mixing time. In part I the fundamental concepts on fluid flow phenomena involving a top layer are reviewed. Part II will deal with the industrial considerations.

2 Effect of top layer on mixing time

Gas stirring in agitated ladles accomplishes two main roles; mixing and mass transfer. Optimum mixing conditions not necessarily represent optimum conditions for mass transfer. The work

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conducted by Nakanishi *et.al.* [1, 2] in the mid 1970's was a pioneer work which defined a quantitative relationship between stirring energy and mixing time, where mixing time is inversely proportional to the stirring energy elevated to a power "n". His group analyzed gas injection in industrial size reactors. Haida *et al.* [3] and Ying *et al.* [4] compared mixing time with and without a top layer. They found that the presence of the top layer increases mixing time. Figure 1 illustrates the results reported by Haida *et al.* The following equations summarize the results reported by Haida. Equation (1) describes mixing time without slag and Equation (2) mixing time with the presence of a top layer. In this work, the top layer was simulated by doubly layered polystyrene balls of 0.9 g/cm^3.

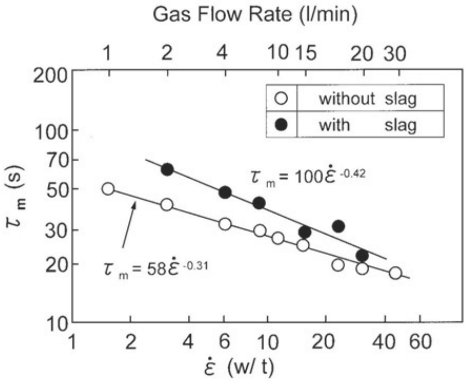

Fig.1 Relationship mixing time and stirring energy [3].

$$\tau_m = 58\varepsilon^{-0.31} \tag{1}$$

$$\tau_m = 100\varepsilon^{-0.42} \tag{2}$$

where τ_m represents mixing time in seconds and ε, the specific stirring energy in W/t.

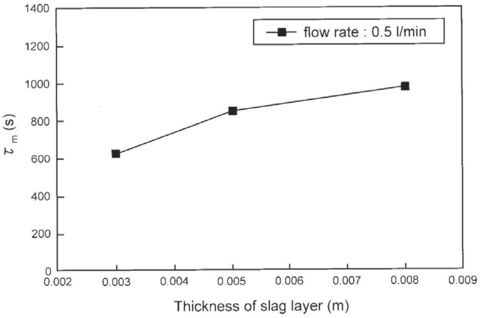

Fig.2 Effect of slag thickness on mixing time [5].

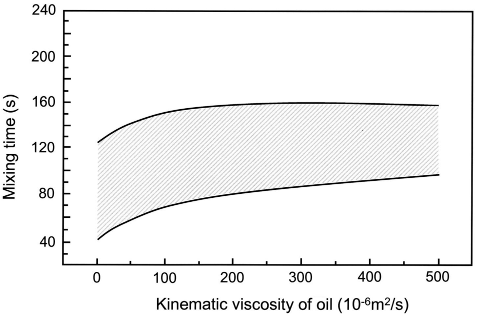

Fig.3 Effect of slag viscosity on mixing time, based on ref. [6].

More recently, a Korean group [5, 6] reported results of mixing phenomena from both physical and mathematical modeling using an injector in a central location, involving a top slag layer, with a density ratio close to one. It is the first numerical approach relating fluid flow and mixing time with a top layer and emphasis on the dynamic behavior of the recalculating loop. The aspect ratio in this investigation was low ($H/D = 0.41$). They found that by increasing slag thickness and slag viscosity mixing time also increases. As shown in Figure 2. This behavior was attributed to a momentum suppression coming from the buoyancy force by rising bubbles in the plume zone. The influence of slag viscosity was investigated using silicone oil. This material can be found in a large of viscosities. They reported an increase in mixing time as the slag viscosity increases, Figure 3 shows the outline of their result. It is observed that at higher flow rates, there is a negligible effect if the kinematic viscosity is above 100×10^{-6}m^2/s.

The previous investigations practically represent the whole amount of experimental research focusing on mixing time involving a top layer. In the past, the vast majority of investigations have neglected the presence of the slag phase, furthermore, the information available with top layer is quite incomplete because those investigations present many limitations; in general, did not include eccentric injection and more important, the density of the top layer was close to that of the underlying phase, whereas in the real process that density ratio is higher than 2. In addition to this, scale up predictions on mixing time from physical modeling to the industrial scale has been practically ignored by the scientific community. Neifer *et al.* [7] reported a comprehensive numerical investigation including bubble gas expansion due to the heating process. This work will be described in more detail in a later section. Ridenour *et al.* [8] also reported mathematical modeling results applied to industrial size ladles. In this work they focused on the influence of slag thickness and density on fluid flow. It was found that the slag increases recirculation and reduces the dead zones located at the bottom corners due to the lower position of the recirculation pattern. More slag improves overall stirring. This finding is in contradiction with previous results which indicate an increase in mixing time due to the presence of the top layer. They also reported a decrease in slag eye by increasing the slag thickness and decreasing the slag density.

Initially mixing time was related only with stirring energy, suggesting that ladle's dimensions and mode of energy input have no influence on mixing. This would be possible if turbulent or eddy diffusion effects dominate over convection phenomena. Mazumdar and Guthrie [9] made a critical assessment of this problem. With reference to the continuity equation, which includes both convection and turbulent diffusion terms, they analyzed two limiting situations, excluding one of those terms in each case. The equation for the mixing time derived for those limiting conditions appeared to satisfy the experimental data, with a perfect fit for the case of dispersion dominated by turbulent diffusion phenomena, however predictions using the simplified version of the continuity equation applied to each limiting condition was inaccurate to represent the experimental data. Such inconsistency indicates that both convection and eddy diffusion are important, consequently ladle dimensions should be included in any analysis of mixing time. The same authors proposed the following relationship to represent mixing time as a function of ladle dimensions.

$$\tau_m = 37\varepsilon^{-\frac{1}{3}} R^{\frac{5}{3}} H^{-1} \tag{3}$$

Where τ_m represents mixing time in seconds, R is the vessel radius in meters and H is the height of liquid in meters.

There is a large amount of relationships involving mixing time with ladle dimensions and stirring energy, however for conditions neglecting the slag layer. There are few investigations defining mixing time as a function of stirring energy and ladle dimensions involving a slag layer.

Mazumdar and Kumar [42] investigated mixing time by dimensional analysis and water modeling involving a top layer. The experimental work was carried out with two nozzles located diametrically opposite at half radius on the consideration of the lowest mixing time (this subject is discussed in more detail in part II). The physical properties of the top layer taken into account in their dimensional analysis were its thickness, molecular viscosity, density and surface tension, however the kinematic viscosity was neglected on the basis of a thin, low mass, low density slag in the real system, conditions which promote that the hydrodynamic conditions within the bulk phase remain dominated by inertial and gravitational forces and not by viscous forces. The obtained functional relationship between mixing time and gas flow rate with and without a top layer was similar. This result was used to proof that the hydrodynamic conditions are dominated by those two forces. The experiments were carried out with three oils; petroleum ether ($\rho = 640\text{kg/m}^3, \sigma = 0.0161\text{N/m}, \mu = 0.00038\text{kg/m}\cdot\text{s}$), Benzene ($\rho = 873\text{kg/m}^3, \sigma = 0.0288\text{N/m}, \mu = 0.00038\text{kg/m}\cdot\text{s}$) and Mustard oil ($\rho = 919\text{kg/m}^3$, $\sigma = 0.035\text{N/m}, \mu = 70\text{kg/m}\cdot\text{s}$). The final expression for mixing time in terms of the physical properties of the top layer is given below.

$$\tau_m = 60.2\varepsilon Q^{-0.33} R^2 H^{-1} h_s^{0.6} \left(\frac{\sigma_s}{\mu_s}\right)^{-0.022} \tag{4}$$

Where: τ_m represents mixing time in sec., Q is the gas flow rate corrected to mean height and temperature of the liquid in m^3/s, h_s represents the slag thickness in m, σ_s is the surface tension of the slag in N/m and μ_s is the molecular viscosity of the slag in $\text{kg/m}\cdot\text{s}$, R and H represent the radius and liquid height of the vessel in m.

Real industrial ladles, in addition to tapered walls, also may have a non-flat bottom in order to improve metallic yield. Mazumdar *et al.* [10] investigated the influence of ladle design on mixing time. They found that the specific geometry of a ladle's bottom influences mixing, the extent however, depends on the gas injection configuration. In all cases, a flat ladle bottom gave the shortest mixing time. Eccentric gas injection produced the shortest mixing time if the nozzles are located at half radius, in comparison with $0.64R$.

Iguchi *et al.* [11–13] have conducted extensive research on physical and mathematical modeling. They reported that the mean flow and turbulence motions in the recirculation region located outside the bubbling jet region is drastically suppressed by the top layer. The phenomenon was attributed to the entrainment of top slag into steel in a real system. They also measured mixing time in a 100% silicone oil bath

$$\tau_m = 1200 Q^{-0.47} D^{1.97} H_L^{-1} \nu_L^{0.47} \tag{5}$$

Where: τ_m represents mixing time in sec., Q is the gas flow rate in m^3/s, D and H the diameter and liquid height, respectively in m, and ν_L is the kinematic viscosity of the liquid in m^2/s.

Mazumdar and Guthrie [14] criticized the previous equation on the basis that mixing phenomena in agitated ladles is controlled by inertial and gravitational forces (i.e. viscous forces are of secondary importance) and a low Reynolds (Re) number in Iguchi's experiments. In the reply [16] to the previous arguments Iguchi *et al.* indicated that the primary purpose of their investigation was to define mixing time of the slag, however their relationship proved to be applicable to conditions with higher Re numbers. In any case, they pointed out the large standard deviations employing both relationships (Iguchi's and Mazumdar's) when collecting a large amount of experimental data from other researchers, therefore, better relationships are required. Yamashita *et al.* [15] reported an equation including the density difference between water and a top layer. This equation indicates that increasing the density difference between slag and steel, increases mixing time.

$$\tau_m = 1910 Q^{-0.217} D^{1.49} H^{-1} \nu_w^{0.37} \left[\left(\rho_w - \rho_o\right)/\rho_w\right]^{-0.243} \tag{6}$$

Where, τ_m represents mixing time in sec., Q is the gas flow rate in m^3/s, D and H the diameter and liquid height, respectively in m, ν_w is the kinematic viscosity of the underlying phase in m^2/s, ρ_w and ρ_o represent the density of the underlying and top phase, respectively, in kg/m^3.

Discussions in water modeling experiments similar to the previous one have been reported in the literature, expressing that the evolution in the comprehension of concepts has not attained its final point. Precisely, one very important point in the discussion is the measurement of mixing time. Mietz and Oeters [17] reported an investigation on the influence of tracer addition and measurement position on mixing time, for both central and eccentric gas injection with one tuyere. They concluded that mixing time depends on tracer injection location due to the presence of dead zones. Krishna Murthy [16] argues that there is only one true mixing time if the system has the same degree of mixing, if there are differences in the degree of mixing agitation has to be continued until the desired extent of mixing is attained in the bath. The idea is correct, however it is almost impossible to define the exact location of dead zones, unless specialized equipment such as PIV is used, therefore, in practical terms is better to refer to the mixing time as local mixing time if the tracer sensor is placed arbitrarily.

There is a large group of investigations analyzing mass transfer, emulsification phenomena and ladle eye formation involving a top layer in physical and mathematical modeling. This work will be reviewed to get a better understanding about the influence of the slag layer on fluid flow.

3 Effect of top layer properties on emulsification phenomena

Poggi *et al.* [18] described two mechanisms to explain emulsion formation due to central gas injection: Each bubble, surrounded by a liquid film, rises to the top layer, then if the top layer has low viscosity, the bubble shatters into fine drops and into larger drops if the viscosity is higher. They used three top layers with density ratios higher than two and viscosities from 0.1-1.1 Pa·s. The volume of liquid carried up into the upper phase and the effect of the top layer on the recirculation velocity was investigated by Guthrie *et al.* [20–23]. They artificially modified the density of the water with additions of $ZnCl_2$, obtaining an equation to describe slag emulsification:

$$R_b \approx 0.19\mu_m \rho_s \mu_s^{-\frac{1}{3}} \Delta\rho^{-\frac{5}{3}} \frac{Q}{d_B} \tag{7}$$

Where: R_b represents the net volume of droplets lifted into the upper phase per unit time, in $\mathrm{cm^3/s}$, μ_m and μ_s represent the viscosity of metal and slag, respectively, in $\mathrm{g/cm \cdot s}, \Delta\rho$ is the density difference between the top and underlying phase, in $\mathrm{g/cm^3}$, Q is the gas flow rate in $\mathrm{cm^3/s}$, and d_B the bubble diameter in cm. The previous equation indicates that emulsification is increased by; (i) decreasing the density ratio between the two liquid phases, (ii) decreasing upper phase viscosity, (iii) decreasing bubble size. Additionally, they observed a drastic reduction in the total kinetic energy due to the top layer. They proved that the effect of interfacial friction on input energy dissipation is negligible, suggesting that such energy dissipation is consumed during the formation of slag droplets and by the potential energy to keep them entrained.

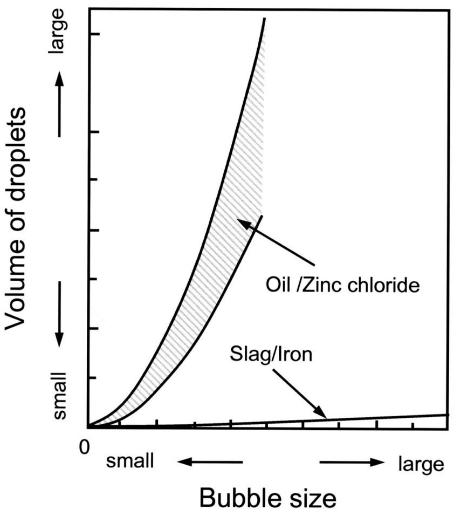

Fig.4 Schematic presentation about the influence of bubble size on the volume of droplets based on ref. [23].

Figure 4 summaries their results as well as predictions for the steel/slag system, using the following data: $\mu_m = 5 \times 10^{-2}$ g/cm s, $\mu_s = 4.5$ g/cm s, $\rho_\mathrm{s} = 3.0$ g/cm^3 and $\Delta\rho = 3.8$ g/cm^3. According with these predictions, slag emulsification in the slag/iron system is negligible, if the bubble size is below 20 mm.

Mietz *et al.* [24] reported an equation to define the critical velocity for slag emulsification, which indicates that decreasing slag density also decreases the critical velocity. Kim *et al.* [25] suggested a more practical approach defining a critical gas flow rate for slag emulsification, valid for vessels with an aspect ratio equal to one.

$$Q_{crit} = 3.8 \times 10^{-3} H^{1.80} \left(\frac{\sigma \Delta\rho}{\rho_s^2} \right) \tag{8}$$

Where, Q_{crit} is the critical gas flow rate in l/min, σ represents the interfacial tension in dyn/cm, H is the liquid height in m, ρ_s is the density of the slag, in g/cm^3. This equation has a similar form to that reported for the critical velocity. Predicted values from this equation for a 200 t steelmaking ladle, at an operating temperature of 1773K, with densities for the slag and steel of 3 and 7 g/cm^3, and a range of interfacial tensions from 400 to 1500 dyn/cm, give a critical gas flow rate from 17-26 Nm^3/h. If interest is focused in inclusion flotation, the gas flow should be less than the Q_{crit} but if desulphurization is more important, then, it should be higher. Emulsification is important to enhance mass transfer [26], however a strong emulsification may create problems with permanent droplet entrainment in liquid steel. Sahajwalla *et al.* [27] proposed an empirical criteria to define the critical was flow rate in terms of the mixing power density, equal to 0.8 W/kg. This criterion is more general, covering ladles of any size.

Kim and Fruehan [28] reported an abrupt increase in the mass transfer parameter at approximately 5 l/min, equivalent to 5 W/t, attributed to oil droplet entrainment into the bulk liquid phase. An increase in mass transfer due to increasing the gas flow rate has been reported by several investigators. It is usually expected higher mass transfer coefficients as mixing time decreases, however, the central injection reported higher values of the mass transfer parameter and this position is not the one with the shortest mixing time. The reason of this behavior was given in terms of the movement of the top layer. Central injection promotes homogeneous stirring of the top layer; this is in contrast with linear off center arrangements of tuyeres, which produce stagnant zones in the top layer. Therefore, optimum conditions to minimize mixing time not necessarily represent the optimum conditions to maximize mass transfer.

4 Effect of top layer on ladle eye formation

For the last 10 years, measurements of ladle eye have been reported [29–40]. The ladle eye area is produced when the slag is pushed away during the exit of the injected gas. It is also called exposed eye or open spout area. Several relationships involving slag thickness, density difference between the two liquid phases, gas flow rate and liquid's height with ladle eye have been reported. The subject has been strongly debated among researchers, questioning the concepts and simplifications employed. Yonezawa and Schwerdtfeger [29] reported ladle eye area from water modeling and industrial conditions. In physical modeling they employed a mercury/silicon oil system to simulate the steel/slag system. The whole set of experimental results couldn't be reported in one single relationship. The relationships were given in terms of a dimensionless number defined as slag's Froude Number, expressed in terms of the gas flow rate at the nozzle exit and the slag height:

$$Fr = \frac{Q_b^2}{gh_s^5} \tag{9}$$

Where; Fr represents the slag Froude number, Q_b represents the gas flow rate at the bottom of the ladle in m^3/s, g is the gravity constant in m/s^2 and h_s is the slag thickness in m.

The reported empirical relationship, valid for $0 < Fr < 2000$, is the following:

$$\log\left(\frac{A}{H_L h_s}\right) = a + b\log Fr + c\left(\log Fr\right)^2 + d\left(\log Fr\right)^3 \tag{10}$$

Where A is the average open spout area in m^2, Fr is the slag's Froude number, H is the height if the underlying phase in m, h_s is the thickness of the top layer in m, a, b, c and d are constants which fit the data to the polynomial equation (-0.69897, 0.90032, -0.14578, 0.01560, respectively). Subagyo *et al.* [30] used the previous experimental results to define one single relationship, claimed as an improved version. Yonezawa and Schwerdtfeger [31] replied that such a version could be obtained directly from their original one and it was not better but worst. Mazumdar and Evans [32] criticized both works on the basis that the forces controlling fluid flow in ladles are inertial and gravitational and these forces should be properly expressed in terms of the plume velocity and liquid's height, furthermore, they also questioned that the previous experimental work at low gas flow rates was not under Froude dominated control. Mazumdar and Evans [33] reported one equation using a simplified plume geometry. The ladle eye was related with the conventional Froude number and they concluded that one single relationship could not represent both cold model and industrial conditions. Krishnapishadory and Irons [34] developed a mechanistic model to define the ladle eye area, based on a momentum balance over the control volume located outside the plume region (toroid region) with a thickness h_s. The final result of this analysis, assuming a conical plume, yields the following expression:

$$A_e = H_L^2\left[\alpha + \beta\left(\frac{\rho}{\Delta\rho}\cdot\frac{U_P^2}{(gh_s)}\right)^{0.5}\right] \tag{11}$$

Where: A_e is the area of the ladle eye, $\Delta\rho$ is the density difference between the two liquids, ρ represents the density of the lower phase, α and β are constants which result from plotting the dimensionless ladle eye with the densimetric Froude number, $\left(Fr_D = \frac{\rho}{\Delta\rho}\cdot\frac{U_P^2}{(gh_s)}\right)$. The previous expression was not adequate to represent the whole set of experimental data. To correct this behavior, they re-calculated the areas assuming a plume with a cylindrical shape. This change is supported by previous results from Ebneth and Plunshkell who defined a relationship for the plume radius (R_p) as a function of gas flow rate and height, as follows.

$$R_p = 0.38Q^{0.15}H^{0.62} \tag{12}$$

The previous relationship shows that the plumes do not increase linearly with bath height and become more cylindrical as the bath becomes deeper. Using the previous relationship the non-dimensional ladle eye area (A_e/A_p) produced better results. A relationship involving the density difference was previously reported by Iguchi *et.al.* [35] however it was not capable to reproduce the whole set of experimental data from Yonezawa and Schwerdtfeger. In the experimental work by Krishnapishadory and Irons [34], they analyzed the influence of the aspect ratio, slag thickness and slag viscosity, gas flow rate and nozzle/porous plug position. They criticized the validity of Mazumdar and Evans relationship on the following basis; (i) the ladle eye area is not smaller than the plume diameter, (ii) the ladle eye area with Mazumdar and Evans equation yields an unrealistic minimum at a Froude number of 2, (iii) it does not include the density difference between liquids and (iv) the dome height was calculated neglecting the presence of slag. Mazumdar and Evans replied [36] that their relationship was extrapolated at conditions where the modified Froude number in terms of the slag thickness (Fr_h) ≤ 2 (a condition of no visible slag) and the use of substantially smaller specific energy inputs than those in industrial conditions. Irons *et al.* [37] replied that an any Fr_h value, the correlation provided by Mazumadar and Evans was inaccurate, not only for $Fr_h \leq 2$, furthermore, if the ladle eye were not visible for $Fr_h \leq 2$, they should not have any data from their experiments, which was not the case. What appears evident from this discussion is that Mazumdar and Evans derived a relationship neglecting the presence of the slag and the final result was used to measure a parameter which involves the slag. Figure 5 shows the schematic image of the plume geometry employed by (a) Krishnapishadory and Irons [34] and (b) by Mazumadar and Evans [33].

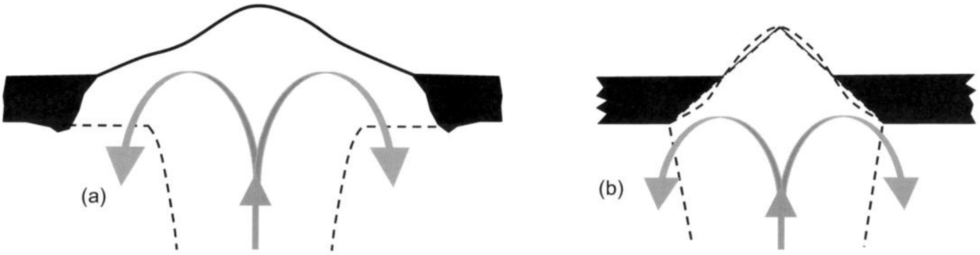

Fig.5 Image of plume geometry according with (a) Irons based on ref. [34] and (b) Mazumdar based on ref. [33].

More recently, Krishnapishadory and Irons [38,39] reported new relationships which appear to satisfy the experimental data from a large group of researchers.

$$A_e^* = -0.76\,(Q^*)^{0.4} + 7.15\,(1-\rho^*)^{-0.5}\,(Q^*)^{0.73}\,(t^*)^{-0.5} \tag{13}$$

Where; ρ^* represents the density ratio of the liquids (ρ_s/ρ_m), Q^* is the dimensionless gas flow rate $\left(Q/g^{0.5}H^{0.5}\right)$ and t^* is the dimensionless slag thickness (h_s/H).

They also found that ladle diameter, size and type of gas injector as well as the physical properties of the gas-liquid system do not play any appreciable role in the spout formation, and furthermore, the presence of slag reduces the spout height. This last result was also previously reported by Trinidade *et al.* [40]. Valentin *et al.* [41] measured ladle eye geometry for industrial conditions. They reported a change from circular to elliptical above 15Nm^3/h and an average ladle eye of 1m^2.

5 Conclusions

Fluid flow phenomena in bottom gas stirred ladles have been reviewed in this work. In the past 35 years a large body of knowledge has been accumulated which has been used to improve the ladle furnace process. The research work has promoted a better understanding of the mechanisms involved in bubble stirring, consequently, several equations have been proposed to quantify mixing time, mass transfer, ladle eye and impurity removal. In spite of this enormous progress there are issues which still require additional research in order to improve the current limitations in both physical and mathematical modeling, such as involving the top layer and scale up laboratory data to industrial conditions.

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