Mathematical modelling of slag dispersion in CLU reactor

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A study on slag dispersion in a model of bottom-blown bath smelting converter was undertaken at room temperature. The aim was to investigate the behaviour of the dispersed slag phase at various operating conditions with very high gas flow rates in low slag volume (10%) systems. For simulation purposes water, paraffin-oil, and air were used to represent the bulk steel, molten slag, and gases respectively in a one-fifth cold model of the commercial 100 t Creusot-Loire-Uddeholm (CLU) converter used in the manufacture of stainless steel in South Africa. Dimensional analysis was done to correlate mathematically the dispersed phase holdup with the operating conditions and the modified Froude number. The overall expression of the holdup revealed that the dispersed slag phase decreased with the vertical distance from the original interface between the liquid phases. The dispersed phase also decreased with the radial distance in the water plume zone toward the sidewall. The holdup was apparently much higher on the side of the reactor where there were no tuyeres. The modified Froude number ensured the correlation between the model and the prototype.

Introduction

During the smelting process, high heat and mass transfer rates occur as a result of the intense mixing and large interface developed between metal and slag. The formation of slag-metal emulsions is of considerable practical importance in pyrometallurgy, particularly during the refining stage. These emulsions create a large interfacial area between the slag and molten metal, resulting in rapid chemical reactions and mass transfer rates in bath smelting, in the basic oxygen furnace (BOF), in the Q-BOP, and in the commercial Creusot-Loire-Uddeholm (CLU) converter.

Only a few investigations have been conducted on metallurgical emulsions in high-strength bottom, top, and combined gas-blown liquid–liquid systems in converters. A number of studies have addressed the emulsification process caused by gas bubbles rising through a slag/metal interface using oil/water analogies by. From the literature it is clear that most of these investigations have focused on liquid–liquid emulsions for low gas flow rates and low slag volumes. Lin and Guthrie (1994) studied the emulsification caused by gas bubbles rising through a slag/metal interface using aqueous modelling techniques. For systems of large differential density with a thick upper phase (*e.g.* in the bath smelting process), it was found that the dispersion of the lower phase into the upper phase was more significant than *vice versa*. Dimensional analysis was used to express the volume of lower liquid carried up into the emulsion per bubble. Lee and Sohn (1996) carried out a cold model study on the effects of various operating conditions on the dispersed phase holdup in liquid-liquid emulsion generated by bottom gas injection. They used water and kerosene to simulate the dispersion of slag in the QSL leadmaking process. The variation of the holdup within the plume was correlated by a single equation involving a set of dimensionless numbers. Akdogan and Eric (2004) carried out a physical modelling study to simulate the slag-metal dispersion induced by a high-rate bottom gas in a one-seventh model of a CLU converter utilized in ferroalloy refining. The dispersed phase holdup was determined at various axial and radial distances, gas flow rates, and tuyere configurations (off-centre, triangular centre, and centre). They found out that off-centre configuration resulted in better dispersion. Most of the previous investigations were basically concentrated on other features of emulsion, such as the nature of the emulsion, drop size distribution, and drop formation, rather on the dispersion of one phase in another. Little work has been done on the modelling of the dispersed slag phase in bath smelting with respect to the positions and side of the tuyeres. The simulated tri-phase (water, paraffin, and air) conditions used in the current work are based mainly by the following criteria:

- The need for inducing the same type of physical phenomena, such as immiscibility, in both reactors (model and prototype), considering that the density difference between steel and slag is 2.5–2.6, and that of water and paraffin-oil is 1.3
- The close physical similarity (Martin *et al.*, 2005)

- The use of water as simulated liquid steel in the isothermal cold model, because of the nearly equal kinematic viscosity of water at room temperature and liquid steel at 1600°C (Fabritius *et al*., 2003)

The objective of the present study was to study the dispersion behaviour in a high-strength bottom-blown CLU converter with off-centre located bottom tuyeres used in the stainless steel industry. The tuyere configuration used in the experiments was different to that in previous studies (Akdogan and Eric, 2004; Nyoka, 2001). Useful information regarding the dispersion behaviour in a stainless steel reactor was obtained by using water as the continuous phase and paraffin-oil as the dispersed phase to represent the metal and slag respectively. For this purpose, a one-fifth scale water model of the bottom-blown CLU converter was used at various injection rates.

Experimental

The experimental set-up comprised a clear cylindrical PVC tank, representing a one–fifth cold model of a 100 t commercial CLU converter used in the manufacture of stainless steel. For simulations purposes, water and paraffin were used to represent the molten steel and slag phase respectively, as shown in Figure 1. The air flow rate was varied from 0.01 m^3/s to 0.019 m^3/s. The heights of water and oil were fixed at 0.4 m and 0.035 m respectively, and the radial distance was in the range of 0 cm (centre line) and 19 cm (wall side).

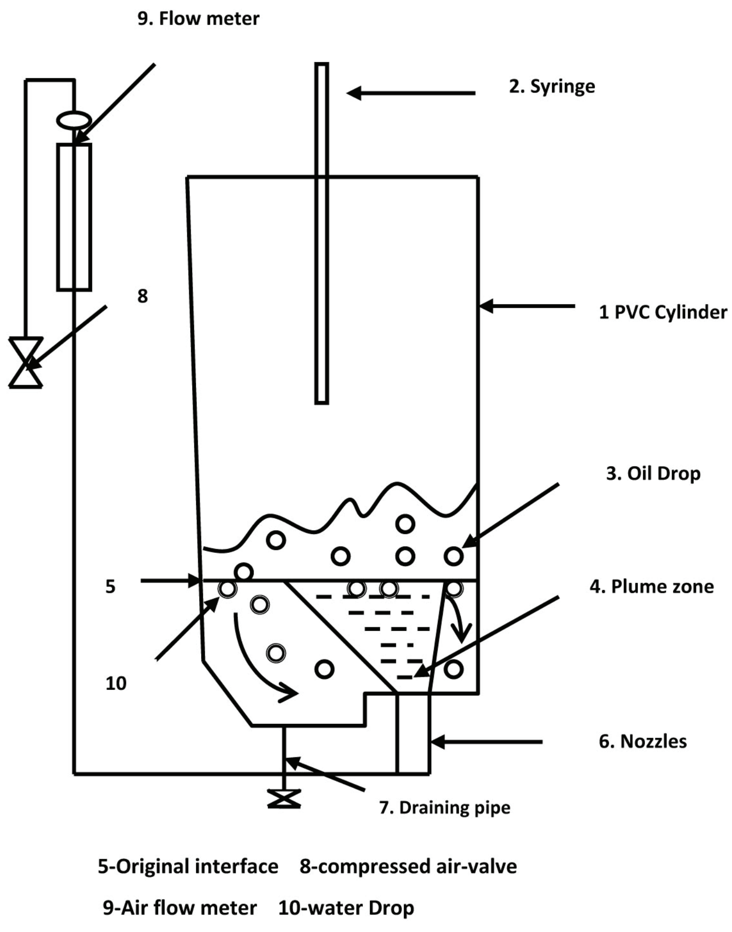

Figure 1. Schematic representation of the experimental set-up

Experiments were run by injecting compressed air into the bath through fives nozzles located at the bottom on the right-hand side of the tank The axial and radial position from where the samples were taken were recorded together with the gas flow rate and the heights of water and paraffin before gas injection. After a half-hour injection period, a sample emulsion was collected rapidly at a particular position by means of specially made syringe fixed firmly by clamps. It was found that the system reached steady state after 20 minutes. Sample emulsions were taken eight times, as shown in Figure 1 and the averaged value of the dispersed slag phase holdup was calculated by means of the relationship:

$$Dph = \frac{V_O}{V_O + V_W} \qquad [1]$$

Mathematical model of the dispersed phase

The operating variables were defined by the axial distance, radial distance, gas velocity, and heights of oil and water. When combined with the physical properties of the liquids, these resulted in a total of nine variables. Only the operating conditions, such as the gas velocity, the height of water and paraffin, and the axial and the radial distances, were varied. The physical properties of the liquids were kept fixed. The dispersed slag phase was defined as a function of the following variables: z, r, v, H_w, H_o, $\Delta\rho_l$, ρ_g , g, μ]. There were three dimensions: length, mass, and time. Application of the Buckingham theorem therefore yielded six dimensionless groups, as shown through the dimensional set in Table I, where, $\Delta\rho_l$ g, and μ represent the density difference of the liquid phases, acceleration due to gravity, and dynamic viscosity respectively.

Table I. Dimensional set of the dispersed phase

	r	z	H_o	H_w	v	$\Delta\rho_l$	ρ_g	g	μ
L	0	0	0	0	0	1	1	0	1
M	1	1	1	1	1	-3	-3	1	-1
T	0	0	0	0	-1	0	0	-2	-1
π_1	1	0	0	0	0	0	0.667	0.333	-0.667
π_2	0	1	0	0	0	0	0.667	0.333	-0.667
π_3	0	0	1	0	0	0	0.667	0.333	-0.667
π_4	0	0	0	1	0	0	0.667	0.333	-0.667
π_5	0	0	0	0	1	0	0.334	-0.334	-0.334
π_6	0	0	0	0	0	1	-1.001	0.001	0.001

The six dimensionless groups that were relevant to the process could be written as follows:

$$\pi_1 = r.\rho_g^{0.6667}.g^{0.3333}.\mu^{-0.6667} \qquad [2]$$

$$\pi_2 = z.\rho_g.^{0.6667} g^{0.3333}.\mu^{-0.6667} \qquad [3]$$

$$\pi_3 = H_o.\rho_g^{0.6667}.g^{0.3333}.\mu^{-0.6667} \qquad [4]$$

$$\pi_4 = H_w.\rho_g^{0.6667}.g^{0.3333}.\mu^{-0.6667} \qquad [5]$$

$$\pi_5 = v.\rho_g^{0.3334}.g^{-0.3334}.\mu^{-0.3334} \qquad [6]$$

$$\pi_6 = \Delta\rho_l.\rho_g^{-1.001}.g^{0.0001}.\mu^{0.0001} \qquad [7]$$

Dividing all the dimensionless groups by π_4 yields the following dimensionless groups.

$$\frac{\pi_1}{\pi_4} = \frac{r}{H_w} \qquad [8]$$

$$\frac{\pi_2}{\pi_4} = \frac{z}{H_w} \qquad [9]$$

$$\frac{\pi_3}{\pi_4} = \frac{H_o}{H_w} \quad [10]$$

$$\frac{\pi_5}{\pi_4} = \frac{v}{H_w}\rho_g^{-0.3333}.g^{-0.6667}.\mu^{0.3333} \quad [11]$$

$$\frac{\pi_6}{\pi_4} = \frac{\Delta\rho_l}{H_w}.\rho_g^{-1.6667}.g^{-0.3333}.\mu^{0.6667} \quad [12]$$

By squaring the dimensional group in Equation [11] and combining with Equation [12], the dimensional group called the modified Froude number (Equation [13]) was defined, which permitted analogies and correlations to be developed:

$$\frac{\pi_5^2}{\pi_6.\pi_4} = \frac{\rho_g}{\Delta\rho_l}\left(\frac{v^2}{g.H_w}\right) \quad [13]$$

Hence, the overall correlation of the dispersed phase was defined as a mathematical function of four dimensionless groups as shown in Equation [14]:

$$\varphi = f\left(\frac{z}{H_w},\frac{r}{H_w},\frac{H_o}{H_w},N_{Fr}\right) \quad [14]$$

Assuming that the dispersed slag phase is defined at the centre line of the bath by φ_{cent}, the corresponding value of the dispersed phase holdup at any position within the bath from the centre can be estimated through a model of the dispersed slag phase. The current model may be defined as a mathematical expression that regroups the value of the holdup at the centre line and all relevant dimensionless groups such as the axial distance from the original interface, the relative height of the liquids, the radial distance, and the modified Froude number. In this case, the axial and radial distances are given as ratios to H_w. By performing nonlinear regression, the equation of the dispersion model was approximated to the following mathematical expression:

$$\varphi_{(z.r)} = \varphi_{cent}\left[1-\left(\frac{r}{H_w}\right)^2\right]^a \quad [15]$$

where

$$\varphi_{centerline} = 0.22\left[(Nfr)^{0.12}\left(\frac{H_o}{H_w}\right)^{0.74}\left[1-\left(\frac{z}{H_w}\right)^2\right]^{0.483}\right]^{1.02} \quad [16]$$

$$a = \left(\frac{z}{H_w}\right)^{-0.3851}(N_{Fr})^{0.2610}\left(\frac{H_o}{H_w}\right)^{1.0661} \quad [17]$$

Results and discussion

The radial dispersed phase was studied by keeping all operating conditions constant and varying the radial distance in the range 0–19 cm. It was found that the dispersed phase decreased with the radial distance from the centre line towards the wall side of the tank, at different axial distances as shown in Figure 2 and Figure 3. That effect could be justified by the decrease in turbulence, and bubbles dispersion from the plume zone ($r = 10$ cm), where nozzles were located towards the recirculation region ($r = 19$ cm). It was observed that viscous forces and drag forces were much more significant in the recirculation zone, and bubbles were rare due to the discontinuity of turbulence and the formation of dead zones. However, Lee and Sohn (1996) pointed out that the decrease in the radial dispersed phase was due to the conical shape of the plume. In the CLU model, the water plume was observed mostly on the right-hand side of the tank, where the tuyeres were located.

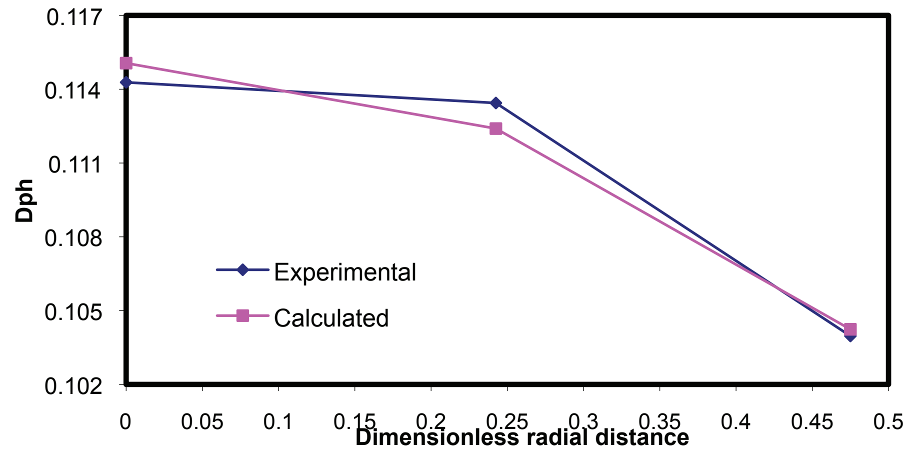

Figure 2. The correlation of radial dispersed phase: experimental and calculated values (z= 25 cm)

Equation [15] was used to represent the Dph at axial positions including the radial variations. The relationship between the experimental Dph values and those calculated by using Equation [15] is illustrated in Figures 2, 3, and 4. The standard error between the two values was 0.0037 which indicates very good agreement between the experimental Dph values and the values calculated by Equation [15] with an R^2 value of 0.97. For comparison, Figure 4 also contains data and correlation from the study of Akdogan and Eric (2004) for the one-seventh scale model of the CLU used in the ferroalloy industry. There is again very good agreement, clearly indicating the success of using Equation [15]. According to the previous model, the dispersed slag holdup increased with the radial distance. Conversely, the opposite effect was observed on the right-hand side of the model converter in the current work. Lee and Sohn (1996) characterized the dispersed phase in the water plume zone as a Gaussian distribution.

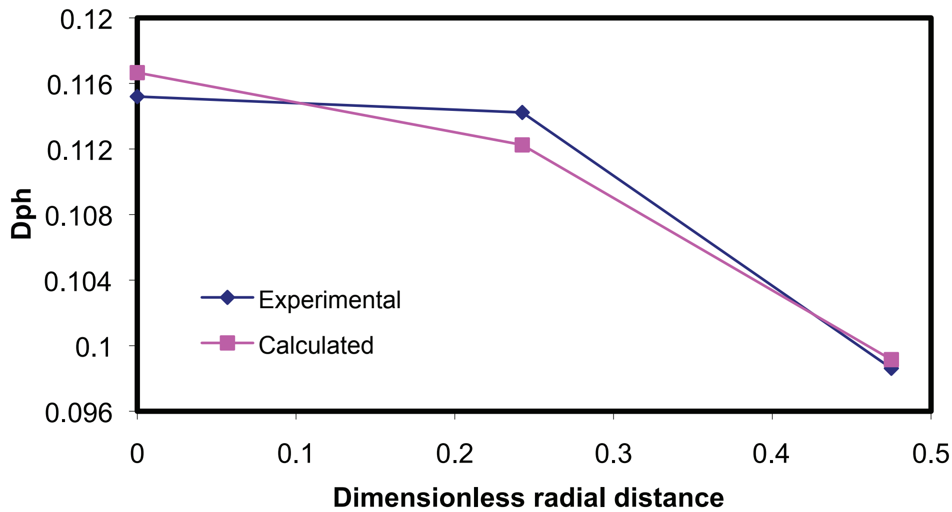

Figure 3. Correlation between experimental and calculated radial dispersed phase ($z = 5$ cm)

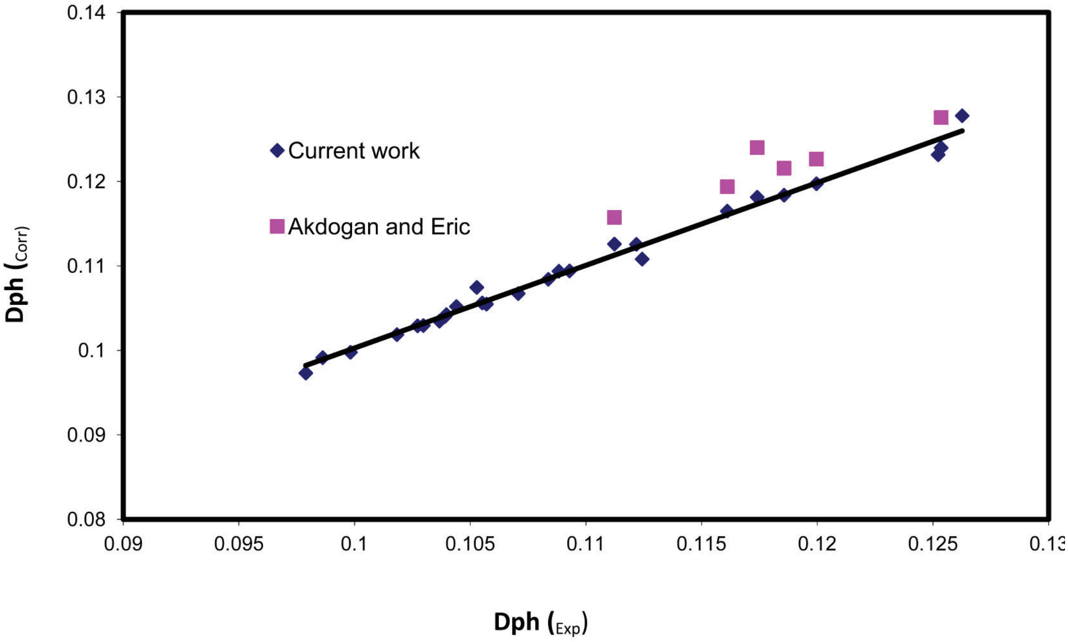

Figure 4. Correlation between the experimental and the calculated results

Industrial application

The physical model developed in the present case has laid the foundation for defining the dispersed slag phase holdup in a CLU converter as a correlation of dynamic forces and dimensionless numbers.

Considering the higher interfacial tension and the density difference between slag and liquid steel as compared to the water and oil system, one should expect the values of the dispersed slag phase of the prototype to be lower than the experimental values. The dispersed slag phase holdup varies from the plume zone towards the recirculatory zone as shown in Figure 2

In the cold model, the interfacial tension between the liquid phases is constant, while it varies with temperature in the real converter. Scheller and Hagemann (2012) pointed out that the strong change in the interfacial tension due to the strong chemical reactions at the steel-slag interface can lower surface tension, and therefore promote the dispersion of slag droplets into the steel.

Conclusions

Using a cold model of a bottom-blown CLU converter used in the stainless steel industry, investigations were carried out to model the dispersed slag phase holdup at various operating conditions. The following conclusions were drawn.

- The experimental results revealed that the dispersed phase holdup increased with increasing gas flow rate, and decreased with increasing vertical (axial) distance from the original interface between the two liquid phases toward the bottom of the vessel
- The dispersed phase holdup also decreased with increasing radial distance, and was inversely proportional to the relative height of the liquid phases. It was apparently much higher on the right side of the reactor, where the tuyeres were located, than on the left side, where there were no tuyeres
- The variation of the dispersed phase holdup with axial and radial distances and gas flow rate was correlated by using the modified Froude number. The experimental values were found to be in very good agreement with the calculated values obtained from the correlation equations.

The physical and mathematical model of the dispersed slag phase holdup developed in this work might provide a useful basis to describe the slag dispersion in actual industrial CLU processes. However, care should be taken when scaling up these findings to real steel-slag systems, as the model does not take into account the effects of chemical reactions in the real system.

Symbols

V_o	Volume of oil in the sample emulsion
V_w	Volume of water in the sample emulsion
□*h*□□□□***Dhp***, Φ	Dispersed phase holdup
μ	Dynamic viscosity
z	Axial distance from the original interface
r	Radial distance
H_o	Height of oil in the bath before air injection
H_w	Height of water in the bath before air injection
ρ	Density
v	Velocity of gas flow rate at the nozzle tip
Fr	Froude number
N_{Fr}	Modified Froude number
L	Length
M	Mass
T	Time
Q	Volumetric gas flow rate

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