Modelling of Liquid Flow in the Blast Furnace.

Theoretical Analysis of the Effects of Gas, Liquid and Packing Properties

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Molten iron and slag flows play a critical role in the blast furnace lower zone, transporting mass and energy, whilst impairing and redistributing gas flow. In turn, molten iron and slag undergo physical and chemical changes, and are redistributed radially during descent to the hearth. Using a 'force-balance' approach, the flows of liquid in the blast furnace were characterised. A consistent set of equations describing liquid holdup, gas-liquid interaction and solid-liquid interaction was developed with reference to previous experimental studies and furnace conditions. The model accounts for the effect of gas, liquid and packing properties on liquid flow, as well as the effect of liquid on gas flow. Interaction between metal and slag phases occurs via a shared gas flow field. Importantly, the model can be applied under both countercurrent and non-countercurrent conditions, where gas can either hinder or enhance liquid flow.

KEY WORDS: blast furnace; liquid flow; packed bed; gas-liquid two phase flow; modelling.

1. Introduction

Molten iron and slag flows play a critical role in the blast furnace lower zone, transporting mass and energy, whilst impairing and redistributing gas flow. In turn, molten iron and slag undergo physical and chemical changes, and are redistributed radially during descent to the hearth. A comprehensive numerical model of the blast furnace must incorporate liquid in its formulation, preferably as two distinct phases of molten iron and slag due to their disparate physical properties. Solution of the liquid flow fields will improve the prediction of other operational variables such as gas temperature and pressure distributions, and allow the analysis of important phenomena such as silicon pickup in molten iron.

The 'force-balance' approach of Gupta *et al.*¹⁾ has been demonstrated to provide a suitable basis for liquid flow modelling in the blast furnace.^{2,3)} Conceptually, it assumes that liquid flows as discrete droplets or rivulets, quickly achieving a steady average percolation velocity based on gas, liquid and packing properties:

$$\vec{F}_l^g + \vec{F}_l^{\text{gravity}} + \vec{F}_l^s \dots\dots\dots(1)$$

where $\vec{F}_l^g$, $\vec{F}_l^{\text{gravity}}$ and $\vec{F}_l^s$ are the gas drag force, gravitational force and solid drag force on liquid, respectively.

The model is capable of predicting liquid velocities in two, or three, dimensions and the computational grid is independent of the particle size. However, effective use of the force-balance approach depends strongly on adequate characterisation of liquid holdup and phase interactions. These

parameters, in turn, depend on the physical properties of gas, molten iron, slag and solid, and microstructural properties such as the size and shape of liquid droplets and the packing structure.

Possibly the most comprehensive analysis of liquid flow under blast furnace conditions was that of Fukutake.⁴⁾ Experiments were performed over an extensive range of conditions relevant to both molten iron and slag flow, providing gas pressure drop and liquid holdup information:

$$h_s^* = (20.5 + 0.263 \cdot C_{p_m})^{-1} \dots\dots\dots(2)$$

$$h_d^* = 6.05 Re_m^{0.648} Ga_m^{-0.485} C_{p_s}^{0.097} N_c^{0.648} \dots\dots\dots(3)$$

$$h_{to} = h_s^* + h_d^* \dots\dots\dots(4)$$

$$h_t = h_{to} + 0.679 h_{to} X_p^2 \dots\dots\dots(5)$$

$$X_p = \frac{\Delta P_w}{\Delta L \rho_l g} \left(\frac{\rho_l g \phi^2 d_p^2}{\sigma (1 - \varepsilon)^2} \right)^{0.3} (1 + \cos \theta)^{-0.5} \dots\dots\dots(6)$$

$$\frac{\Delta P_w}{\Delta L} = \frac{k_1 \left(\frac{1 - \varepsilon}{\phi d_p} + \frac{h_t}{d_l} \right)^2 \mu_g U_{og} + k_2 \left(\frac{1 - \varepsilon}{\phi d_p} + \frac{h_t}{d_l} \right) \rho_g U_{og}^2}{(\varepsilon - h_t)^3} \dots\dots\dots(7)$$

$$d_l = [6.828 \sqrt{X_p} - 0.891]^2 + 0.695 / \sqrt{\rho_l g / \sigma} \dots\dots\dots(8)$$

where h_s^* , h_d^* , h_{to} and h_t are the modified static, modified dynamic, total (without gas flow) and total (with gas flow) liquid holdup, respectively, C_{p_m} , Re_m , Ga_m and C_{p_s} are the

modified Capillary, modified Reynold's, modified Galileo and effective Capillary number, respectively, N_c and X_p are the dimensionless interfacial force and gas pressure drop, respectively, and $\Delta P_w/\Delta L$, ρ_l , g , ϕ , d_p , σ , ϵ , θ , k_1 , k_2 , d_l , μ_g , ρ_g and U_{og} are the wet bed pressure gradient, liquid density, gravity, particle shape factor, particle diameter, surface tension, porosity, contact angle, viscous term Ergun coefficient, inertial term Ergun coefficient, liquid droplet diameter, gas viscosity, gas density and gas superficial velocity, respectively.

In this paper, an attempt will be made to quantify the complex interactions between gas, liquid and solid under blast furnace conditions. Equations (2) to (8) will be re-analysed and modified, based on experimental findings, for application in the force-balance model. The force-balance approach will then be used to describe the effect of gas, liquid and packing properties on liquid flow in the blast furnace.

2. Liquid Holdup Without Gas Flow

Liquid holdup is typically defined in terms of static and dynamic contributions. Static holdup, h_s , is the volume fraction of liquid that remains in a fully irrigated packing after gas and liquid sources are removed. This holdup must be considered in an analysis of blast furnace liquid flow because: it is transported by the mobile packing, contributing to the overall flow; it is typically more significant than dynamic holdup,⁵⁾ thus contributing more strongly to the change in bed voidage and permeability to gas flow; and dynamic holdup cannot be considered in isolation from static holdup since the former establishes the latter and there is constant interchange between the two, with both subjected to the influence of gas drag.

Fukutake⁴⁾ defined a modified static holdup, h_s^* (Eq. (2)), that included an additional contribution, h_f , that is only present during irrigation but was assumed to be independent of liquid velocity:

$$h_s^* = h_s + h_f \quad \text{.....(9)}$$

where h_s is the true static holdup and h_f is the difference between the modified and true static holdup.

Static holdup will be influenced by gravitational, surface and solid-liquid interfacial forces, as encapsulated in the modified Capillary number, Cp_m . Static holdup data extracted from Fukutake's investigation were plotted against Cp_m (Fig. 1) (Eq. (10), $R^2=0.673$). The correlation compared favourably to Eq. (2) ($R^2=0.689$).

$$h_s = (21.0 + 0.305 Cp_m)^{-1} \quad \text{.....(10)}$$

The correlation for static holdup was determined for single-liquid phase flow. Based on an analysis of holdup in the dissected Chiba No. 1 blast furnace, Fukutake *et al.*⁶⁾ proposed that the static holdup under two-liquid phase flow could be obtained by adding the contributions of the two liquids calculated individually. While this approach was adopted for the current analysis, it might be expected that two liquids would compete for occupation of similar static holdup sites such as the contact point between particles.

Dynamic liquid holdup transports the majority of liquid between its generation point in the cohesive zone and the

Fig. 1. Dependence of static holdup on the modified Capillary number.

hearth. It is this liquid, flowing relative to the packed bed, that is governed by the force-balance model. Consistent with the earlier analysis of static holdup, h_f should be added to the dynamic holdup calculated by Eq. (3). However, the assumption that h_f is independent of the superficial liquid velocity is inconsistent with the force-balance model. Therefore, h_f was modified to approach zero as irrigation is decreased, giving Eq. (11). The average superficial liquid velocity in the blast furnace, $U_{ol,BF}$, was incorporated in Eq. (11) to ensure that calculated holdup was similar to Fukutake's results under typical conditions. At higher velocities, the component of dynamic holdup calculated by Eq. (3) becomes dominant, masking any change in h_f . The index, 'n', which should be in the range of zero to one, will be discussed later.

$$h_d = h_d^* + (U_{ol}/U_{ol,BF})^n (h_s^* - h_s) \quad \text{.....(11)}$$

where h_d is the true dynamic holdup and U_{ol} is the superficial liquid velocity.

3. Liquid-Solid Interaction

In the absence of gas flow, the interaction between liquid and solid must be sufficient to retain static holdup stationary relative to the packed bed and dynamic holdup at a steady average percolation velocity under the influence of gravity, according to the force balance model.¹⁾ The gravitational force acting on dynamic liquid holdup is proportional to both the holdup volume and liquid density:

$$F_{l,dynamic}^{gravity} = \rho_l g h_d \quad \text{.....(12)}$$

In the absence of gas flow, this force will be balanced by the bed drag force. Under saturated flow conditions the bed drag force can be described by the Ergun⁷⁾ equation, which may be expressed as:

$$F_{l,saturated}^s = \frac{150}{36} \mu_l \left(\frac{A_{sl}}{V_l} \right)^2 \left(\frac{U_{ol}}{V_l} \right) + \frac{1.75}{6} \rho_l \left(\frac{A_{sl}}{V_l} \right) \left(\frac{U_{ol}}{V_l} \right)^2 \quad \text{.....(13)}$$

where μ_l is the liquid viscosity, A_{sl} is the liquid-solid contact area and V_l is the liquid volume.

It was proposed that the equation be applied to the unsaturated liquid flow in the blast furnace, in which case only the dynamic liquid holdup participates in the force balance:

Fig. 2. Liquid velocity for mercury flow through glass ballotini.

$$F_1^s = \frac{150}{36} \mu_l \left(\frac{A_{sl}}{h_d} \right)^2 \left(\frac{U_{ol}}{h_d} \right) + \frac{1.75}{6} \rho_l \left(\frac{A_{sl}}{h_d} \right) \left(\frac{U_{ol}}{h_d} \right)^2 \dots (14)$$

Little data exists relating to the flow of non-aqueous liquids through packed beds. Mackey⁸⁾ has presented data for the flow of mercury through 10 mm glass sphere packings. This system was of particular interest due to the use of a liquid metal with disparate physical properties from the liquids used to generate Eqs. (2) to (8). Applying Eq. (11) and data from Mackey (with $\varepsilon=0.4$), liquid velocity was predicted as a function of flowrate and compared to experimental results at different values of the index n (Fig. 2). Experimental results were extrapolated through a representative superficial liquid velocity for metal in the blast furnace, $U_{ol,m}$, of $1 \times 10^{-4} \text{ m}^3/\text{m}^2\text{s}$ using the fitted curve of Mackey:

$$U_l = 3.981 U_{ol}^{0.49} \quad (\text{SI units}) \dots (15)$$

Results indicated that the predicted liquid velocity became too high as the superficial velocity was increased if no correction was applied to the dynamic holdup ($n=0$). However in the region of interest to the blast furnace, prediction of velocity, and by implication dynamic holdup, was very good. As superficial velocity increased, application of the h_r correction factor allowed predicted results to better fit the experiments. At $n=0.5$, Eq. (15) and the velocity predicted using Eq. (11) were well matched over the entire range of superficial liquid velocity. Given the high density, surface tension and contact angle of mercury, it therefore appeared reasonable to apply the corrected dynamic holdup (Eq. (11)) in the analysis of blast furnace liquid flow with n set to 0.5.

The effective solid-liquid contact area for dynamic holdup can be determined from the balance between gravitational (Eq. (12)) and bed drag (Eq. (14)) forces:

$$\frac{A_{sl}}{h_d} = \frac{-f_1 + \sqrt{f_1^2 + 4f_2 \rho_l g h_d}}{2f_2} \dots (16)$$

where

$$f_1 = \frac{1.75}{6} \rho_l \left(\frac{U_{ol}}{h_d} \right)^2, \quad f_2 = \frac{150}{36} \mu_l \left(\frac{U_{ol}}{h_d} \right)$$

Applying Eqs. (11) and (16), the solid-liquid contact area was estimated for the system of Mackey. On a bed volume basis, the contact area increased as the amount of liquid in

Fig. 3. Liquid-solid contact area and predicted liquid rivulet diameter for mercury flow through glass ballotini.

Fig. 4. Geometry of liquid rivulet contacting the packing surface.

Table 1. Physical properties of liquids and packing in blast furnace.

Liquid	ρ (kg m^{-3})	μ (Pa s)	σ (N m^{-1})	θ ($^\circ$)	ϕ (-)	d_p (m)	ε (-)
metal	6600	0.005	1.1	90	0.8	0.035	0.4
slag	2600	1.0	0.47	90	0.8	0.035	0.4

the bed increased, as expected (Fig. 3). On a liquid volume basis, the opposite trend was observed. This was consistent with an increase in the characteristic size of the liquid stream, and therefore decrease in surface area to volume ratio, as the irrigation density increased. Assuming that mercury flows as part-cylindrical rivulets over glass spheres, with the geometry determined by the contact angle (Fig. 4), a rivulet diameter, d_r , was estimated for the case of continuous contacting between liquid and packing based on the liquid-solid contact area:

$$d_r = \frac{4 \cos(\theta - \pi/2)}{(A_{sl}/h_d)(\theta + \cos(\theta - \pi/2) \sin(\theta - \pi/2))} \dots (17)$$

The predicted rivulet diameters (Fig. 3) were implausibly large compared to the packing size. In order to bring the estimates into a viable range, it would be necessary to assume that the droplets were not in continuous contact with the packing during descent. This accorded with the observation of 'Type 1' flow behaviour in the experiments of Mackey, where liquid flowed as droplets or rivulets in the packing void space. It also reinforced the macroscopic nature of the force-balance model since liquid in freefall will achieve very different instantaneous velocities from that in contact with packing surfaces.

In the absence of gas flow and using an average superficial liquid velocity for slag, $U_{ol,s}$, of $8 \times 10^{-5} \text{ m}^3/\text{m}^2\text{s}$ together with the physical properties in Table 1, the characteristics of molten iron and slag flow through the blast furnace

Table 2. Estimated properties of liquid flow in the blast furnace.

Liquid	h_s (%)	h_d (%)	h_{10} (%)	U_{0f}/h_d (m/s)	A_{sf}/h_d (m^2)	A_{sf}/V_b (m^2)
metal	1.67	0.28	1.95	0.0351	75.7	0.215
slag	1.75	0.85	2.60	0.0094	73.4	0.623

dropping zone were estimated. Predictions for static (Eq. (10)), dynamic (Eq. (11)) and total holdup, true liquid velocity, and solid–liquid contact areas on a liquid (Eq. (16)) and bed volume basis are presented in **Table 2**.

The average combined holdup of metal and slag in the dropping zone was estimated to be approximately 4.5% in the absence of gas flow. While static holdup was similar for metal and slag, dynamic holdup was greater for the latter as a result of the lower gravitational and higher viscous forces. This was reflected in the lower velocity and higher total liquid–solid contact area estimated for flowing slag. Under operational conditions, gas–liquid interaction, liquid maldistribution, and varying physical properties of the liquids and packing are expected to promote much greater local accumulations of liquid in the blast furnace.

4. Gas–Liquid Interaction

Liquids generated in the blast furnace cohesive zone must descend to the hearth *via* the same packing pore space through which gases ascend from the raceways to the lumpy zone. Consequently, these phases interact during passage, altering their respective flow fields. If this interaction is excessive then furnace stability and/or productivity may be compromised. Similar concerns have prompted numerous studies into loading and flooding phenomena in gas–liquid countercurrent packed bed reactors.^{9–11} However, as for liquid holdup, the unique operating characteristics of the blast furnace require careful consideration. Fukutake⁴) considered some, but not all of the implications. For example, the concept of flooding, which is a limiting condition in the operation of a one-dimensional column, must be reanalysed for the non-uniform packed bed of a blast furnace.¹²

In the one-dimensional system considered by Fukutake,⁴) the total holdup determined without gas flow was recalculated as a function of the gas pressure gradient. In simple terms, gas flowing countercurrent to liquid changed the holdup which, in turn, affected the pressure drop due to a change in bed permeability. Consequently, an iterative calculation was required. Equations were given for liquid holdup (Eq. (5)) and effective liquid droplet diameter, d_1 (Eq. (8)), as a function of dimensionless gas pressure drop, X_p (Eq. (6)), and gas pressure drop as a function of liquid holdup and droplet diameter (Eq. (7)).

The total pressure drop in Eq. (7) is calculated according to an Ergun-type expression. This pressure drop is the result of both packing and liquid drag on the gas phase. The area of interaction between gas and condensed phases is changed from the non-irrigated condition by the combined effect of holdup and the liquid geometrical factor, d_1 . Additionally, pore space for gas flow is reduced by holdup, increasing the average gas velocity through the voids. While gas–solid contact area is decreased due to liquid–solid contacting, gas–liquid contact area contributes to the

Table 3. Estimated areas of interaction between gas and liquid.

Liquid	$6/d_1$ (m^{-1}) $\sqrt{X_p}=0$	$6h_d/d_1$ (m^{-1})	A_{sf}/h_d (m^{-1})	$6/d_1$ (m^{-1}) $\sqrt{X_p}=0.891$	$6h_d/d_1$ (m^{-1})
metal	238	0.68	314	2093	8.51
slag	228	1.94	302	2010	24.37

pressure drop. For the case where liquid is distributed isotropically in the packing:

$$A_s - A_{sl} + A_{gl} = 6[(1-\varepsilon)/\phi d_p + h_t/d_1] \dots\dots\dots(18)$$

where A_s is the packing surface area and A_{gl} is the gas–liquid contact area.

Analysis of the interaction between gas and liquid flows is complicated by the presence of both static and dynamic holdup. In the previous section, these holdups were handled separately. By contrast, Fukutake made no distinction between the two based on their effect on gas flow with neither the coefficients, k_1 and k_2 , nor the area terms in Eq. (7) given to be a function of the relative amounts of static and dynamic holdup. Making this assumption, the gas–liquid contact area for dynamic holdup can be estimated without detailed knowledge of the distribution of static holdup in the packing. Applying Eq. (18) to the dynamic holdup:

$$A_{gl,d} - A_{sl,d} = 6h_d/d_1 \dots\dots\dots(19)$$

where $A_{gl,d}$ and $A_{sl,d}$ are the gas–liquid and solid liquid contact areas for dynamic holdup, respectively.

Applying Eqs. (5), (8) and (19), the difference between gas–liquid and solid–liquid interfacial area was estimated for metal and slag at zero gas flow and for the minimum effective droplet diameter ($\sqrt{X_p}=0.891$), as given in **Table 3**. Using the solid–liquid interfacial areas from Table 2, gas–liquid interfacial areas were also estimated at zero gas flow. The latter were significantly greater than the former, which is consistent with both the high liquid contact angles on coke and the argument that liquids are not in constant contact with the packing surface.

Since there will always be a finite holdup in an irrigated packing, d_1 plays an important role in moderating the calculated effect of liquid on gas flow. Based on the quadratic relationship of Eq. (8), d_1 decreases by an order of magnitude between zero and intermediate gas flow rates, before increasing again as flooding is approached. Qualitatively, the decrease is consistent with the experimental findings of Gardner¹³) for non-wetting flow of water over coated coke, where gas flow caused liquid to disperse into thinner or more numerous and smaller trickles. Additionally, as holdup increases, liquid movement to regions of low gas velocity within a packing will decrease due to saturation of these sites. The transition from decreasing to increasing d_1 can be associated with a change in the type of liquid flow. For example, the flow may progress from droplets to rivulets as the droplets become larger or more numerous and link up, with increased connectivity decreasing their external surface on a liquid volume basis.

The force of interaction between gas and dynamic liquid holdup was estimated by applying Eq. (7) to the fraction of the total area of interaction determined using Eq. (19):

Table 4. Estimated gas–liquid interaction force at $\sqrt{X_p}=0.891$.

Liquid	$F^g (Nm^{-3})$	$F_{gravity} (Nm^{-3})$	$F^g/F_{gravity} (%)$
metal	539	263	205
slag	582	309	188

$$F_{l,d}^g = \frac{k_1 \left(\frac{A_{gl,d}}{6} \right) \left(\frac{1-\varepsilon}{\phi d_p} + \frac{h_t}{d_l} \right) \mu_g U_{og} + k_2 \left(\frac{A_{gl,d}}{6} \right) \rho_g U_{og}^2}{(\varepsilon - h_t)^3} \dots\dots\dots(20)$$

The solution of Eq. (20) requires Eqs. (5), (6) and (8) and coefficients k_1 and k_2 , which were set at 150 and 1.75 to ensure that Eq. (7) would reduce to the Ergun relationship for a dry packing. The gas–liquid interaction force on dynamic holdup for metal and slag at the minimum effective droplet diameter was approximated using the difference between gas–liquid and solid–liquid interfacial area from Table 3, as given in **Table 4**. Since the magnitude of the solid–liquid interfacial area must be positive, this represented a minimum for the estimated force. For comparison, the gravitational force was also calculated using Eq. (12). The calculated interaction force exceeded the gravitational force by a significant margin. Hence, at intermediate gas flow rates where flooding was not expected, the effective area of interaction for dynamic holdup was overestimated using Eq. (19).

A likely cause of the excessive interaction force is the assumption that liquid distributes isotropically in the packing. Therefore, the calculation of the effective area of interaction was altered to allow for an anisotropic liquid distribution, by introducing a modified effective liquid droplet diameter, d_{lm} . The effect of gas flow on the dynamic holdup was decreased by weighting the modified liquid droplet diameter between the liquid droplet diameter calculated with and without gas flow, using the parameter α , which will be discussed later:

$$d_{lm} = \alpha d_l + (1 - \alpha) d_{lo} \dots\dots\dots(21)$$

where d_{lo} is the liquid droplet diameter in the absence of gasflow.

Replacing d_l with d_{lm} in Eq. (19) and solving for the effective area of gas–liquid interaction:

$$A_{gl,d} = 6h_d/d_{lm} + A_{sl,d} \dots\dots\dots(22)$$

Substituting this area term into Eq. (20) with standard Ergun coefficients gave the interaction force between gas and dynamic liquid holdup:

$$F_{l,d}^g = \frac{\left(\frac{h_d}{d_{lm}} + \frac{A_{gl,d}}{6} \right) \left[150 \left(\frac{1-\varepsilon + h_t}{d_w} \right) \mu_g U_{og} + 1.75 \rho_g U_{og}^2 \right]}{(\varepsilon - h_t)^3} \dots\dots\dots(23)$$

where d_w is the common effective packing diameter, as defined by Fukutake:

$$d_w = \phi d_p d_l (1 - \varepsilon + h_t) / [d_l (1 - \varepsilon) + \phi d_p h_t] \dots\dots\dots(24)$$

Table 5. Estimated liquid holdup and gas pressure drop.

Liquid	h_{lo} (%)	h_t (%)	h_{do} (%)	h_d (%)	d_l (m)	$\Delta P/\Delta L_{dry}$ (Nm^{-3})	$\Delta P/\Delta L_{wet}$ (Nm^{-3})
metal	1.95	2.04	0.28	0.30	0.0070	1484	3889
slag	2.60	3.32	0.85	1.08	0.0032	1484	3889

5. Gas–Metal–Slag Interaction

Previously, the interaction between metal and slag was only mentioned in relation to static holdup, since these liquids must ‘compete’ for holdup sites in the same packing. Dynamic holdup is generally considered to be less significant in the furnace due to the low superficial velocities of both metal and slag.⁵⁾ Consistent with the force-balance model, which neglects liquid–liquid interactions on the basis that flow occurs as discrete rivulets or droplets, it was assumed that direct interactions are minimal.

However, metal and slag will influence each other indirectly since the gas pressure field is common and would be higher than for either liquid acting alone. Hence, Fukutake⁴⁾ extended Eq. (7) for application to the blast furnace:

$$\frac{\Delta P_w}{\Delta L} = (k_1 f_3^2 \mu_g U_{og} + k_2 f_3 \rho_g U_{og}^2) / (\varepsilon - h_{t,s} - h_{t,m})^3 \dots\dots\dots(25)$$

where $f_3 = (1 - \varepsilon) / \phi d_p + h_{t,s}/d_{l,s} + h_{t,m}/d_{l,m}$, $h_{t,s}$ and $h_{t,m}$ are the holdup of slag and metal, respectively, and $d_{l,s}$ and $d_{l,m}$ are the liquid droplet diameter for slag and metal, respectively.

Recalculating the effective packing diameter for this system:

$$d_w = \frac{\phi d_p d_{l,s} d_{l,m} (1 - \varepsilon + h_{t,s} + h_{t,m})}{d_{l,s} d_{l,m} (1 - \varepsilon) + \phi d_p d_{l,m} d_{t,s} + \phi d_p d_{l,s} d_{t,m}} \dots\dots(26)$$

The drag force of gas on either flowing liquid should also consider the interaction between metal and slag. Using Eqs. (23) and (25) for liquid i (m: metal, s: slag):

$$F_{d,i}^g = \frac{k_1 f_4 \left(\frac{1 - \varepsilon + h_{t,s} + h_{t,m}}{d_w} \right) \mu_g U_{og} + k_2 f_4 \rho_g U_{og}^2}{(\varepsilon - h_{t,s} - h_{t,m})^3} \dots\dots\dots(27)$$

where $f_4 = h_{d,i}/d_{l,m} + A_{sl,d,i}/6$.

Using an average superficial velocity for gas in the blast furnace lower zone of $1.9 m_g^3 \cdot m_b^{-2} \cdot s^{-1}$, total and dynamic holdup with gas flow and effective liquid droplet diameter for metal and slag, and gas pressure drop, were calculated by applying Eqs. (5), (6), (8) and (25) and the data from Tables 1 and 2. These results are presented, along with liquid holdup in the absence of gas flow and the dry bed pressure drop, in **Table 5**. The interaction between gas and liquid was significant and should be considered in blast furnace modelling, with the gas pressure drop more than doubling for typical dropping zone conditions. While the effect of gas flow on metal holdup was relatively small, the increase in slag holdup was greater than 27%. Holdup is therefore susceptible to changes in operating conditions and should be considered with reference to the variations in packing properties, gas flow and liquid flow that might be

experienced in the blast furnace lower zone.

6. Gas-Liquid-Solid Interaction

6.1. Countercurrent Gas-Liquid Flow

Under countercurrent conditions at gas flow rates below the onset of flooding, the downward gravity force on liquid phase i is balanced by upward acting gas and solid drag forces. Substituting Eqs. (12), (14) and (27), into Eq. (1):

$$\frac{k_1 f_4 [(1 - \varepsilon + h_{t,s} + h_{t,m}) / d_w] \mu_g U_{og} + k_2 f_4 \rho_g U_{og}^2}{(\varepsilon - h_{t,s} - h_{t,m})^3} + A_{sl,i} / h_{d,i} (A_{sl,i} / h_{d,i} f_2 + f_1) = \rho_l g h_{d,i} \dots \dots \dots (28)$$

Equation (28) contains six unknowns (dynamic holdup of phase i , total holdup of metal and slag, effective packing diameter, modified liquid droplet diameter, and liquid-solid contact area of phase i) in a system with known gas and liquid superficial velocities. Holdups can be determined using Eqs. (5), (6), (8), (10), (11) and (25). Liquid droplet diameters found in the preceding step can then be used to determine the effective packing diameter via Eq. (26). Since the gas drag force must approach the gravity force at flooding as liquid velocity approaches zero, it is possible to estimate the parameter α in Eq. (21). The liquid-solid contact area can then be found and, subsequently, the gas-liquid area through Eq. (22).

Local operating conditions in the blast furnace will differ from those assumed in Sec. 5. In particular, a radial variation of gas and liquid mass flux, and packing properties will occur due to factors such as burden distribution, deadman formation and combustion at the tuyeres. Consistent with the study of Fukutake, the effect of gas superficial velocity was examined by solving the above system of equations, with all other factors held constant. Similar calculations were performed with varying superficial liquid velocity and porosity. By trial and error, parameter α in the modified liquid droplet diameter (Eq. (21)) was estimated at 0.8.

Predicted holdup of metal and slag, gas pressure drop with and without liquid, and the ratio of gas drag to gravity forces for metal and slag as a function of superficial gas velocity are presented in Fig. 5. Gas-liquid and solid-liquid contact areas for metal and slag on a liquid volume basis are presented in Fig. 6. Effective solid diameter, ϕd_p , liquid diameter and modified liquid diameter for metal and slag, and packing diameter are presented in Fig. 7. Results in the

same form as Fig. 5 for variable liquid and solid properties are given in Figs. 8 and 9, respectively.

With increasing gas velocity, gas pressure drop in an irrigated bed was calculated to increase at a greater rate than the equivalent dry bed (Fig. 5). Meanwhile liquid holdup increased relatively slowly at first, before slag approached a

Fig. 6. Effect of superficial gas velocity on liquid area to volume ratios.

Fig. 7. Effect of superficial gas velocity on effective diameters.

Fig. 8. Effect of superficial liquid velocity on gas-liquid interaction.

Fig. 5. Effect of superficial gas velocity on gas-liquid interaction.

Fig. 9. Effect of porosity on gas-liquid interaction.

flooding condition at a gas velocity around $2.3 \text{ m} \cdot \text{s}^{-1}$. At this point, slag holdup was almost triple that calculated at zero gas velocity but still less than 10% of bed volume. Metal holdup was minimally affected. The ratio of gas drag to gravity forces for slag approached unity towards the onset of flooding.

On a liquid volume basis, gas-liquid contact areas initially increased with gas flow (Fig. 6). This implied that droplets/rivulets did not simply enlarge to accommodate increasing holdup, as the surface area to volume ratio would be expected to decrease with increasing volume. Instead the form of flow must have altered as gas disturbed the liquid flow field. Approaching flooding, the gas-slag area levelled off and then began to decrease, consistent with a further change in flow form as holdup increased and droplets/rivulets became larger and/or more connected. The solid-slag area appeared to be similarly affected, gradually decreasing with increasing holdup.

While the effective solid particle diameter is constant for a given packing, changes in effective liquid droplet diameter with gas flow resulted in a variable effective packing diameter and modified liquid droplet diameters for metal and slag (Fig. 7). The latter two variables were closely associated with contact area through Eq. (22), and were seen to follow the opposite trends to gas-liquid contact areas (Fig. 6). At low gas flow rate, decreasing liquid diameter combined with increasing holdup to decrease the effective packing diameter. However, as liquid diameter stabilised for metal and increased for slag, effective packing diameter levelled off, despite the increasing influence of liquid as holdup increased. Figure 7 illustrates an important advantage of the present model over some alternatives^{1,14,15)} in that the characteristic size of liquid is able to vary independently of the packing, and therefore capture changes in the liquid flow form.

The effect of superficial liquid velocity on gas pressure drop did not appear to be as strong as gas flow, with a gradual approach to flooding at over 300% of the rates used to generate Table 2 (Fig. 8). The increase in pressure drop was matched by a gradual increase in holdup and the ratio of gas drag to gravity forces for both metal and slag. These results were a reflection of the low dynamic holdup experienced in the blast furnace, since any change in this holdup to accommodate an increase in liquid flow did not represent a significant change in the pore space available for gas flow. The ratio of gas drag to gravity forces for slag was again seen to approach unity with the onset of flooding.

Liquid flow was found to be very sensitive to changes in porosity (Fig. 9). This result must be carefully considered, since lower porosities are quite plausible in the blast furnace. For the average superficial gas and liquid velocities employed in this analysis, gas pressure drop, holdup and the ratio of gas drag to gravity forces increased slowly as porosity decreased to around 0.45. However, once porosity had decreased to the assumed average of 0.4, pressure drop and parameters for slag were changing at a significant rate. Once porosity dropped below 0.38, the bed approached a flooding condition. For comparison, by decreasing the gas velocity from 1.9 to $1 \text{ m} \cdot \text{s}^{-1}$ (a decrease of $\sim 47\%$) the porosity at flooding was found to drop below 0.3 (a decrease of $\sim 20\%$). These results clearly illustrate the compe-

Fig. 10. New and existing correlations for liquid holdup with gasflow.

titution for pore space between gas and liquid.

Just as the radial distribution of properties (such as ore to coke ratio) in a blast furnace promote formation of a permeable cohesive zone, they might also be advantageous in securing stable liquid flow to the hearth. For example, the average gas velocity used to generate Fig. 9 does not account for gas flow redistributing away from the deadman to the dropping zone as the relative permeability of the former decreases. Furthermore, redistribution of gas implies radial flow. The effect of radial flow on holdup was not considered in the countercurrent analysis of Fukutake. Studies of cross-flow cascade packed columns indicate that these can be operated at gas velocities above the flooding points of countercurrent packed beds.¹⁶⁾

6.2. Non-countercurrent Gas-Liquid Flow

While the blast furnace may be macroscopically characterised as a countercurrent reactor, there are a number of operating features which locally promote significant radial gas flow. The most obvious of these are the tuyeres which inject blast laterally into the furnace near its base to form high-voidage regions known as raceways. Adjacent to the raceways, the deadman will influence the distribution of gas flow based on its permeability relative to the dropping zone. Passage through the dropping zone is further influenced by the permeability distribution of the cohesive zone. The cohesive zone itself will accommodate significant radial flow through coke-slits as the permeability of intermediate softening-melting ore layers decreases. Liquids generated in the cohesive zone are affected by all these factors during passage to the hearth.

In the countercurrent system considered by Fukutake,⁴⁾ upward flowing gas created drag on liquid to counteract the downward gravity force and therefore increased the capacity of the bed to hold liquid. If a system was considered where gas and liquid flow were co-current, the drag of gas on liquid would be expected to augment the gravity force and therefore decrease the capacity of the bed to hold liquid. An exponential relationship was chosen to replace Eq. (5) since it could represent the accelerating rate of change of holdup with increasing pressure given by this equation, whilst predicting decreasing holdup for a negative pressure gradient. Using data generated by Eq. (5) over a range sufficient to encompass the onset of flooding in the countercurrent analysis ($0 < X_p < 2$), Eq. (29) was established, as shown in Fig. 10.

$$\frac{h_i}{h_o} = \exp[0.513 \sinh(X_p) |X_p|^{1.321}] \dots\dots\dots (29)$$

Fig. 11. New and existing correlations for liquid droplet diameter.

Two competing influences have been identified for the liquid droplet diameter. Initially, as gas flow increases, the liquid flow is disturbed and the diameter decreases. However, as holdup increases, a point is reached where liquid accumulation affects the surface area and diameter increases. Clearly, the second factor would not be relevant in the co-current system discussed above. Therefore it was deemed necessary to separate the two influences on liquid droplet diameter. The former, $d_{l,g}$, was assumed to be governed by the magnitude of gas velocity (Eq. (30)), and the latter, $d_{l,h}$, by the ratio of total holdup with and without gas flow, allowing substitution of Eq. (29) into Eq. (8) to yield Eq. (31). The applicable liquid droplet diameter was then determined using Eq. (32), resulting in equivalent behaviour to Eq. (8) for countercurrent flow, but a steady minimum diameter for co-current flow, as shown in Fig. 11.

$$d_{l,g} = [\max\{-6.828 \operatorname{sign}(f_5) f_5^2, 0\} + 0.695] / \sqrt{\rho_l g / \sigma} \quad \text{....(30)}$$

where $d_{l,s}$ is the liquid droplet diameter as affected by gas-flow and $f_5 = \sqrt{|X_p|} - 0.891$.

$$d_{l,h} = [\max\{6.828 \operatorname{sign}(f_6) f_6^2, 0\} + 0.695] / \sqrt{\rho_l g / \sigma} \quad \text{.....(31)}$$

where $d_{l,h}$ is the droplet diameter as affected by holdup and $f_6 = (\max[\ln(h_t/h_w), 0] / 0.513)^{1/2.642} - 0.891$.

$$d_l = \max\{d_{l,g}, d_{l,h}\} \quad \text{.....(32)}$$

Like downward gas flow, any radial component of velocity will increase the sum of gas drag and gravity forces acting on liquid and therefore decrease holdup. To apply Eq. (29) in such a system, it was necessary to determine an effective axial pressure gradient, $\Delta P_{e, \Delta x}$, for use in Eq. (6):

$$X_p = \frac{\Delta P_e}{\Delta x \rho_l g} \left(\frac{\rho_l g \phi^2 d_p^2}{\sigma(1-\epsilon)^2} \right)^{0.3} (1 + \cos \theta)^{-0.5} \quad \text{.....(33)}$$

where x is a length measure in the axial direction.

For dynamic liquid holdup, the effective pressure gradient, $\Delta P_{ed}/\Delta x$, will depend on the direction of liquid flow relative to gas flow. Under countercurrent conditions, the angle between gas and liquid flow is 180° and $\Delta P_{ed}/\Delta x$ is equal to the pressure gradient. Under co-current conditions, the angle between gas and liquid flow is 0° and $\Delta P_{ed}/\Delta x$ is equal to the negative of the pressure gradient. At a relative angle of 90° , gas flow neither augments nor hinders liquid flow and $\Delta P_{ed}/\Delta x$ is equal to zero. For intermediate angles, referring to Fig. 12, $\Delta P_{ed}/\Delta x$ can be determined by geometrical considerations:

Fig. 12. Angle between gas and liquid flows.

Fig. 13. Force balance on static liquid holdup.

$$\frac{\Delta P_{ed}}{\Delta x} = -\cos \beta \frac{\Delta P_w}{\Delta L} \quad \text{.....(34)}$$

where β is the angle between gas and liquid flow and $\beta = [(F_l^s)^2 + (F_l^g)^2 - (F_l^{gravity})^2] / [2(F_l^s)(F_l^g)]$.

For static liquid holdup, a balance is achieved between the vector sum of gas drag and gravity forces, and surface forces. The effective pressure gradient, $\Delta P_{es}/\Delta x$, will depend on the magnitude of the former since the latter will oppose flow in any direction. Referring to Fig. 13, if the vector sum of gas drag and gravity forces equals the magnitude of the gravity force, then holdup will not change from the zero gas flow condition and $\Delta P_{es}/\Delta x$ will be equal to zero. Schematically, this is represented by any point equidistant from the origin of the force-balance. If the vector sum is greater, then $\Delta P_{es}/\Delta x$ will be negative (the effective gas drag positive) and holdup will decrease. Alternatively, if the vector sum is less, then $\Delta P_{es}/\Delta x$ will be positive and holdup will increase. Conceptually, the vector sum may be rotated about the origin to coincide with the vertical axis, whereby prior to the onset of flooding (ie. the magnitude of gas drag is less than the gravity force) a simple linear summation will provide the effective gas drag:

$$F_{l, effective}^g = -|\vec{F}_l^{gravity}| + |\vec{F}_l^g + \vec{F}_l^{gravity}| \quad \text{.....(35)}$$

Continuing the assumption that the effective liquid droplet diameter is similar for static and dynamic holdup, $\Delta P_{es}/\Delta x$ was determined from Eqs. (12), (25), (27) and (35):

$$\frac{\Delta P_{es}}{\Delta x} = \rho_l g h_s f_7 - \left[\left(\rho_l g h_s f_7 - \frac{\Delta P_w}{\Delta x} \right)^2 + \left(\frac{\Delta P_w}{\Delta r} \right)^2 \right]^{0.5} \quad \text{.....(36)}$$

where $\Delta P_w/\Delta x$ and $\Delta P_w/\Delta r$ are the gas pressure gradient in

Fig. 14. Schematic of experimental apparatus for the examination of liquid flow with gas cross-flow.¹⁷⁾

Fig. 15. Effect of liquid flowrate on liquid shift with gas cross-flow.

axial and radial directions, respectively, and .

$$f_7 = \left(\frac{1 - \varepsilon + h_{ts} + h_{tm}}{d_w} \right) / \left(\frac{h_s}{d_{lm}} + \frac{A_{sl}}{6} \right).$$

As expected, if the radial gas pressure gradient is zero, Eq. (36) reduces to the axial gradient. Alternatively, if the axial gradient is zero, then $\Delta P_{es}/\Delta x$ can only be less than or equal to zero and holdup less than or equal to the zero gas flow condition, according to Eqs. (29) and (33). An important feature of Eqs. (34) and (36) is that the effective gas pressure gradient will differ between metal and slag, whereas the true gas pressure gradient is common. It should also be recognised that, since holdup does not vary directly with the true gas pressure gradient, liquid droplet diameter may not increase as quickly (if at all) with gas flow as under countercurrent conditions according to Eq. (31), as illustrated by the 'radial flow' case in Fig. 11.

Gupta *et al.*¹⁷⁾ have performed cross-flow experiments with horizontal gas (air) flow in an experimental apparatus of depth 76 mm (Fig. 14). The analysis outlined above was applied to this system for an aqueous barium chloride solution and mercury. The calculation of the superficial liquid velocity was not straightforward due to spreading of liquid away from the point source. However, as indicated by Fig. 8, the present model is not strongly sensitive to this velocity except close to the onset of flooding. The effect of superficial liquid velocity (0.12, 0.24, and 0.48 l/min) on liquid shift was calculated for barium chloride solution at different gas flow rates and compared with experimental data (Fig. 15). The effect of liquid type was calculated for a fixed superficial liquid velocity (0.12 l/min) at different gas flow rates and also compared to experimental data (Fig. 16).

The effect of gas cross-flow on liquid shift increases gradually with liquid flowrate (Fig. 15). Consequently, the choice of flow area for calculating superficial liquid velocity

Fig. 16. Effect of liquid type on liquid shift with gas cross-flow at $U_{ol} = 0.12$ l/min.

Fig. 17. Effect of gas flow direction on slag holdup and gas pressure drop.

ty was not expected to significantly influence the present results. The calculated liquid shift compared favourably to that determined by experiment, though the trend in the latter with increasing superficial liquid velocity was not clear from the limited data. Compared to barium chloride solution, the calculated effect of gas cross-flow on mercury was relatively weak (Fig. 16). Despite the disparate physical properties of the latter, the model again matched experimental results reasonably well.

The effect of superficial gas velocity on liquid holdup and gas pressure drop in a metal/slag system was analysed with the gas velocity at 45 and 90° to the vertical direction. The pressure gradient and total holdup of slag were compared to the countercurrent condition discussed in Sec. 6.1 (Fig. 17). At low gas velocity, the slag holdup and gas pressure gradient were insignificantly affected by radial flow. As gas flow increased, the radial component of velocity resulted in lower holdups compared to the countercurrent case ('0°'). For gas flow at 45° ('45°'), the onset of flooding was delayed. For gas flow at 90° ('90°'), holdup decreased with increasing gas velocity and there was no indication of flooding at gas pressure gradients beyond the onset of flooding for the other cases.

7. Conclusions

Using a 'force-balance' approach, the flows of molten iron and slag in the blast furnace were characterised. A consistent set of equations describing liquid holdup, gas-liquid interaction and solid-liquid interaction was developed with reference to previous experimental studies and furnace conditions. The model accounts for the effect of gas, liquid and

packing properties on liquid flow, as well as the effect of liquid on gas flow. Interaction between metal and slag phases occurs *via* a shared gas flow field. Importantly, the model can be applied under both countercurrent and non-countercurrent conditions, where gas can either hinder or enhance liquid flow. Integration of these techniques into a comprehensive numerical model of the blast furnace is required to further understand both the effect of operating conditions on the liquid flow fields and the effect of liquids on these operating conditions. Further study into the effect of cross and co-current gas flow on liquid holdup, and the interaction of metal and slag static holdup are recommended.

Nomenclature

A_{ij}	Area of interaction of phase i and j (m^2/m^3)
A_s	Surface area of packing (m^2/m^3)
Cp_l	Liquid Capillary number (—)
Cp_m	Modified liquid Capillary No. (Cp_s/N_c) (—)
Cp_s	Effective Capillary No. ($\rho_l g \phi^2 d_p^2 / \sigma (1 - \varepsilon)^2$) (—)
d_l	Effective liquid droplet diameter (m)
d_{lm}	Modified effective liquid droplet diameter (m)
d_p	Particle diameter (m)
d_r	Rivulet diameter (m)
d_w	Effective packing diameter (m)
F_i^{gravity}	Interaction force of gravity on phase i ($\text{N} \cdot \text{m}^{-3}$)
F_j^i	Interaction force of phase j on phase i ($\text{N} \cdot \text{m}^{-3}$)
Ga_m	Modified Galileo No. ($\rho_l^2 g \phi^3 d_p^3 / \mu_l^2 (1 - \varepsilon)^3$) (—)
g	Gravitational acceleration ($\text{m} \cdot \text{s}^{-2}$)
h_{di}	Dynamic holdup of liquid phase i (—)
h_d^*	Modified dynamic liquid holdup (—)
h_f	Liquid holdup (see Eq. (9)) (—)
h_s	Static liquid holdup (—)
h_s^*	Modified static liquid holdup (—)
h_t	Total liquid holdup (—)
h_{to}	Total liquid holdup without gas flow (—)
k_1	Ergun coefficient typically 150 (—)
k_2	Ergun coefficient, typically 1.75 (—)
N_c	Dimensionless interfacial force ($1 + \cos \theta$) (—)
$\Delta P / \Delta L$	Pressure gradient ($\text{N} \cdot \text{m}^{-3}$)
$\Delta P_c / \Delta x$	Effective axial pressure gradient ($\text{N} \cdot \text{m}^{-3}$)
$\Delta P_w / \Delta L$	Wet bed pressure gradient ($\text{N} \cdot \text{m}^{-3}$)
$\Delta P_w / \Delta r$	Radial wet bed pressure gradient ($\text{N} \cdot \text{m}^{-3}$)
$\Delta P_w / \Delta x$	Axial wet bed pressure gradient ($\text{N} \cdot \text{m}^{-3}$)
Re_m	Modified Reynolds No. ($\rho_l U_{ol} \phi d_p / \mu_l (1 - \varepsilon)$) (—)

U_{oi}	Superficial velocity of phase i ($\text{m} \cdot \text{s}^{-1}$)
V_l	Liquid volume fraction (—)
X_p	Dimensionless gas pressure drop (—)

Greek symbols

α	Parameter for d_{lm} (—)
ε	Porosity (—)
ϕ	Shape factor (—)
μ_i	Viscosity of phase i ($\text{Pa} \cdot \text{s}$)
θ	contact angle ($^\circ$)
ρ_i	density of phase i ($\text{kg} \cdot \text{m}^{-3}$)
σ	surface tension ($\text{N} \cdot \text{m}^{-1}$)

Super/subscripts

g, l, s	Gas, liquid, solid
i, j	Gas (g), liquid (l) or solid (s) phase
m, s	Metal, slag
s, d	Static, dynamic

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