1. Introduction

Blast furnaces still represent the most efficient metallurgical process to rapidly reduce ore into iron. In this process, complex gas–liquid flow occurs in the lower part of blast furnace, which plays an important role in achieving stability of the blast furnace operation. The blast furnace process is currently undergoing significant changes due to environmental and economic imperatives. A typical example is the implementation of technologies such as pulverised coal injection (PCI). However, increased PCI results in the decreased combustibility of coal in the raceway, and hence the increased unburnt char in blast furnace. These char particles driven by the ascending gas and descending liquids (i.e. hot metal and slag) travel through the coke and/or ore packing in blast furnace, being consumed by reacting with liquid iron and slag or being deposited onto coke and ore particles. Experiments indicate that the existence of a liquid stream or liquid holdup has a strong effect on the char behaviour in blast furnace. Therefore, further experimental investigation on the gas–liquid flow in the lower part of blast furnace is important, providing better understanding of other in-furnace phenomena such as the char behaviour under the given gas–liquid flow condition.

A distinct feature of the gas–liquid flow in the blast furnace is the spatially non-uniform counter-current flow of gas and liquid in the lower part of blast furnace. Strong gas cross flow could occur within cohesive zone and raceway in blast furnace, which usually occurs around the cohesive zone and raceway in blast furnace. The localised liquid flow phenomenon in presence of gas cross flow, which was investigated in detail. Such liquid flow is characterised in terms of liquid shift distance or liquid shift angle that can effectively be measured by the experiments involved in the current study. It is found that liquid shift angle does not significantly increase or decrease with different packing depth. This finding supports the hypothesis of the force balance model where a vectorial relationship among acting forces, i.e. gas drag force, gravitational force and solid–liquid friction force, and liquid shift angle does exist. Liquid shift angle is inversely proportional to particle size and liquid density, and proportional to square of gas superficial velocity, but is almost independent on liquid flowrate and liquid viscosity. The gas–liquid drag coefficient, an important aspect for quantifying the interaction between gas and liquid flows, was conceptually modified based on the discrete feature of liquid flow through a packed bed and evaluated by the combined theoretical and experimental investigation. Experimental measurements suggest that the gas–liquid drag coefficient is approximately a constant ($C_{DG}=5.4\pm1.0$) and is independent on liquid properties, gas velocity and packing structure. The result shows a good agreement with previous experimental data and prediction of the existing liquid flow model.

KEY WORDS: gas–liquid two-phase flow; discrete liquid flow; gas–liquid drag coefficient; packed bed; blast furnace.

An experimental study has been carried out for the gas–liquid two-phase flow in a packed bed simulating conditions of the gas and liquid flows in the lower part of blast furnace. The localised liquid flow phenomenon, in presence of gas cross flow, which usually occurs around the cohesive zone and raceway in blast furnace, was investigated in detail. Such liquid flow is characterised in terms of liquid shift distance or liquid shift angle that can effectively be measured by the experiments involved in the current study. It is found that liquid shift angle does not significantly increase or decrease with different packing depth. This finding supports the hypothesis of the force balance model where a vectorial relationship among acting forces, i.e. gas drag force, gravitational force and solid–liquid friction force, and liquid shift angle does exist. Liquid shift angle is inversely proportional to particle size and liquid density, and proportional to square of gas superficial velocity, but is almost independent on liquid flowrate and liquid viscosity. The gas–liquid drag coefficient, an important aspect for quantifying the interaction between gas and liquid flows, was conceptually modified based on the discrete feature of liquid flow through a packed bed and evaluated by the combined theoretical and experimental investigation. Experimental measurements suggest that the gas–liquid drag coefficient is approximately a constant ($C_{DG}=5.4\pm1.0$) and is independent on liquid properties, gas velocity and packing structure. The result shows a good agreement with previous experimental data and prediction of the existing liquid flow model.

KEY WORDS: gas–liquid two-phase flow; discrete liquid flow; gas–liquid drag coefficient; packed bed; blast furnace.

Experimental Investigation of Liquid Flow Shift Due to Gas Cross Flow in Non-wetted Packed Beds

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creasing interest in numerical modelling of blast furnace including the gas–liquid flow model, which provides better simulation of the real process and improvement of the blast furnace control.12) For development of the gas–liquid model, the interaction between gas and liquid is a significant consideration. In previous liquid flow models, as classified by the continuum model13–16) and the probabilistic model,11,17,18) the gas–liquid interaction is determined by using the continuous approach, which is conceptually inconsistent with the discrete feature of liquid flow in blast furnace. An alternative approach, i.e. force balance to liquid flow modelling in the blast furnace was presented by Gupta et al.19) The force balance approach was consequently integrated with gas flow model in a packed bed and a stochastic model for liquid dispersion by Wang et al., leading to a gas–liquid two-phase model for general use.20,21) Some of concepts for determination of gas–liquid interaction under discrete liquid flow condition were proposed in this model. However the gas–liquid drag coefficient involved has not yet been evaluated rigorously because of the lack of detailed experimental investigation.

The purpose of this article is to present an experimental study for the gas–liquid two-phase flow in a packed bed simulating conditions of the gas and liquid flows in the lower part of blast furnace. An effective measurement method of liquid shift distance is developed to characterise the localised liquid flow in terms of liquid shift distance or liquid shift angle. Accordingly the gas–liquid drag coefficient will be conceptually modified based on the discrete feature of liquid flow through a packed bed and quantitatively evaluated by the combined theoretical and experimental investigation.

2. Experimental

2.1. Liquids and Packing Materials

A variety of liquids listed in Table 1 were chosen to take into account various liquid properties for simulating the liquid flow condition in blast furnace. The particles used in all experiments were made from wax to simulate the non-wettability of coke bed with liquid in a blast furnace. Three sizes of wax particles were used: 9 mm, 15 mm and 19.4 mm.

2.2. Experimental Apparatus

Figure 1 gives a schematic diagram of experimental apparatus for the current study. The selected liquid was stored in two 0.162 m³ feeding tanks with an approximately constant head. During the experiment, the liquid entered the packed bed vertically through a 10 mm diameter stainless steel tube. The flowrate was controlled using two rotameters that were connected to a three-way valve to switch liquid flow between the two feeding tanks. Compressed air was introduced horizontally into the packed bed, via a variable area flowmeter and pressure gauge used to correct the measured flowrate.

Two scales of physical apparatus were separately designed to reduce wall effect of the models due to using different sizes of particles in the packed bed. The small-scale apparatus that has a measuring section of 485 x 229 x 76 mm (refer to Fig. 2(a)) was used only for filling with 9 mm wax spheres. The large-scale apparatus had inner dimensions of 655 x 460 x 240 mm as shown in Fig. 2(b), and was

![Fig. 1. A schematic diagram of experimental rig.](image-url)
used for the relatively large particles, *i.e.* 15 mm and 19.4 mm wax spheres. Both apparatuses were made from perspex sheet supported by steel bar reinforcement. Liquid was collected in separate collecting boxes at the bottom of the models to measure the liquid distributions. There were 11 collection boxes for large-scale and 13 collection boxes for small-scale one. In addition, an entrance section packed with 3 mm glass beads was used in both apparatuses to distribute the gas flow uniformly across the bed cross section.

### 2.3. Experimental Procedure and Measurements

Before running an experiment, the measuring section of the apparatus was completely filled with the selected wax spheres. To ensure accuracy and reproducibility in collecting the experimental data, the rotameter was calibrated before each experimental run for each kind of liquid. For the large-scale apparatus, gas pressure distributions on the back of apparatus (measuring section) and gas velocities on the exit-cross section from the measuring section were monitored for controlling a uniform and horizontal gas cross flow.

Each liquid flow run was started with a dry packed bed. In a typical experiment, the liquid flowrate in the packed bed was initially set at 10% higher than the desired value until liquid steady state flow field was reached (approximately 15 min). After a steady flow condition was established, the liquid flowrate was reduced to the desired value by switching the three-way valve that connects a specified rotameter with liquid inlet into the packed bed.

Once the steady liquid flow was achieved, a uniform gas cross flow was introduced at the desired flowrate. The gas flowrate was measured by a variable area flowmeter corrected by an upstream pressure. The apparatus was operated for 15 min to establish a stable flow condition of gas–liquid two-phase flow. The maximum of gas superficial velocity, which was calculated from the gas flowrate, was 1.18 m/s (100% scale in the variable area flowmeter) for all gas cross flow experiments. For large-scale apparatus, the upstream pressure was 551 kPa, while this upstream pressure was 303 kPa for small-scale apparatus.

Liquid was collected in several collection boxes and the exit valves on all collection boxes with no liquid flow were then shut. Finally, for the remaining liquid collection boxes, the exit tubes were carefully controlled to allow liquid flow without gas leakage.

It has been noted from previous theoretical and experimental investigations that, in absence of gas flow, liquid added into a packed bed as a point source percolates vertically through the packed bed. Such liquid flow usually disperses within the packed bed, featuring a main stream of liquid that locates at the centreline, corresponding to the liquid inlet, with a normal distribution of flow, as shown in Fig. 3(a). Introducing a gas cross flow gives an increased gas drag force on liquid flow in the horizontal direction. It is possible for the force to horizontally shift the liquid distribution and the liquid main stream a distance from liquid inlet, as shown in Fig. 3(b). In order to experimentally investigate such characteristics of liquid flow in the packed bed with gas cross flow, it is convenient to measure the liquid shift distance or the shift angle, as defined in Fig. 3(b), respectively.
To determine the liquid shift distance (the liquid shift angle can accordingly be calculated as desired), liquid distributions were measured by collecting liquid using the liquid collect boxes at the bottom of apparatus after a steady gas–liquid flow condition had been achieved. Then the mean liquid shift distance is calculated as the mean of the distribution of liquid flow on the bottom cross-section:

\[ \bar{x} = \frac{\sum_{i=1}^{n} \bar{x}_i \cdot L_i}{\sum_{i=1}^{n} L_i} \] ........................(1)

\[ \sigma^2 = \frac{\sum_{i=1}^{n} (x_i - \bar{x})^2 \cdot L_i}{\sum_{i=1}^{n} L_i} \] and \[ \sigma = \pm \sqrt{\sigma^2} \] ........................(2)

where \( \bar{x}_i \) is the horizontal shift corresponding to centreline of the collection box \( i \) from the centreline of liquid inlet pipe, mm; \( L_i \) is the liquid mass flowrate through collection box \( i \), g/s, and \( \sigma \) is the standard deviation of the liquid distribution, %.

Figure 4 gives an example of the liquid distribution histogram for water experiment with the calculated mean liquid shift. This sample experiment was performed in the 9 mm wax sphere packed bed in which the liquid flowrate was 200 ml/min and the superficial velocity of gas cross flow was 1.18 m/s. The packed bed was 200 mm height. In general, the liquid was collected in 3 to 5 collection boxes and spread over 30 to 200 mm depending on gas velocity, particle size and physical properties of liquid. The standard deviation of the liquid distribution was typically ±30 mm and the standard error in mean liquid shift was also of order ±30 mm. The measured error of duplicated experimental runs in mean liquid shift was less than 1%.

Each type of experimental run was duplicated once to measure liquid shift distance or shift angle and to justify the reproducibility of the result. An average value of liquid shift distance or angle was used to present the data. Generally, the relative error of the measurements of liquid shift distance or angle is less than 5% based duplicate experimental result for each case study.

3. Results and Discussion

3.1. Effect of Liquid Flowrate and Liquid Properties

Figure 5 gives the liquid shift distance as a function of the liquid flowrate for water percolation through 15 mm wax spheres under various gas cross flow conditions. It can be seen that the liquid shift distance increases clearly with the superficial velocity of gas cross flow. However, at a given gas cross flow, the liquid shift distance does not vary with liquid flowrate from 100 ml/min to 600 ml/min within experimental error. Meanwhile it was experimentally observed that the dispersion of the liquid distribution also slightly increases with increasing liquid flowrate. The results suggest that the local liquid hold up has a limited change with increasing liquid flowrate, and hence the liquid flowrate only offers very little effect on the liquid shift distance.

In order to investigate the effect of liquid properties such as viscosity and density, a packed bed with a 200 mm height was used, consisting of 9 mm wax spheres. Liquid viscosity was controlled by adding glycerol of 50%, 80% and 85% into water, forming various glycerol–water mixture solutions. Four liquids, i.e., water, 80% glycerol–water mixture solution, CaCl\textsubscript{2} and ZnCl\textsubscript{2} solutions, were used to...
simulate different liquid densities. The detailed physical properties of these liquids were described in Table 1.

In the current study, liquid shift angle, defined in Fig. 3(b), was used to quantitatively describe the effect of various variables on liquid shift movement, for consistency with theoretical analysis on gas–liquid interaction which will be discussed later in detail. The liquid shift angle was calculated in terms of $\tan \alpha$, based on the packing depth and the measured liquid shift distance under a given experimental condition.

Figure 6(a) shows the measurements of liquid shift angle $\tan \alpha$ for three glycerol–water mixture solutions with different viscosity. The gas superficial velocity used in these experiments was 1.18 m/s. As shown in Fig. 6(a), liquid shift angle does not significantly decrease or increase with increasing liquid viscosity. Within experimental error, liquid shift angle is almost independent on the liquid viscosity. This experimental result is consistent with the prediction by a theoretical model for modelling the gas–liquid flow in blast furnace.\(^{22}\)

Figure 6(b) gives the effect of liquid density on the liquid shift angle $\tan \alpha$ under the gas cross flow condition as same as that used in Fig. 6(a). At the given gas cross flow, liquid shift angle $\tan \alpha$ is inversely proportional to liquid density. As a matter of fact, liquid density represents the gravitational force acting on liquid flow that is one of the most important forces to control the main stream of liquid flow through a packed bed. The higher the density of liquid is, the more difficult the dispersion and the shift movement of liquid flow through a packed bed become. This result implies that the primary slag generated from the cohesive layers in blast furnace is more easy to be shifted by the gas cross flow through the cohesive zone, compared with hot metal gradually formed in the dropping zone due to the relatively high density of the hot metal.

### 3.2. Impact of Gas Cross Flow

It has been experimentally revealed that there exists strong gas cross flow in the cohesive zone and in front of the raceway of blast furnace due to formation of impermeable cohesive layers in cohesive zone and horizontal entrance of blast air from tuyers, respectively. Liquid (melting slag and iron) generates from the cohesive layers and flows through the coke bed in the lower part of blast furnace. As observed in previous studies, strong localised liquid flow usually occurs around cohesive zone in this case. This is because increasing gas cross flow forms increasing gas drag force on the liquid flow, resulting in a significant shift movement of liquid flow along with the direction of gas flow.

Experiments under various conditions of gas cross flow were carried out using the packing depth of 200 mm and 410 mm, respectively. Figure 7 indicates the correlation of liquid shift angle $\tan \alpha$ against the squared gas superficial velocity. The results show that liquid shift angle does not change too much with different depth of packed bed, but linearly increases as the squared gas superficial velocity increases.

The result shown in Fig. 7 also suggests a vectorial relationship among liquid shift angle $\alpha$, gas drag force $F_g^d$ and gravitational force $F_g^g$ for the gas–liquid two-phase flow in blast furnace due to gas cross flow. Such a relationship can conceptually be described by Fig. 8, in which three forces acting on the liquid flow are assumed to reach a state of force balance. In Fig. 8, $F_i$ represents the friction force between solid (packing particles) and liquid, and its negative direction determines the direction along which the main stream of liquid flow is shifted. Therefore the vectorial relationship can be expressed in the form of

$$\tan \alpha = \frac{F_g^d}{F_g^g} \quad \text{(3)}$$

for horizontal gas flow. Eq. (3) is useful to quantify the interaction between gas and liquid flows, and will be dis-
of packing particles against liquid and interaction between gas and liquid are not yet fully understood so that a quantitative description for such interactions has not yet been established.

As mentioned above, a reasonable treatment that has been employed for the gas–liquid numerical modelling in a blast furnace is so-called force balance approach.\(^{19–21}\) According to the force balance approach, the liquid flow is controlled by gas drag force \(\vec{F}_g\), gravitational force \(\vec{F}_g\), and solid–liquid friction force \(\vec{F}_s\), as shown in Fig. 8. Under the steady-state condition, the force balance can mathematically be expressed as:

\[
\vec{F}_g + \vec{F}_l + \vec{F}_s = 0 \quad \text{..............(4)}
\]

where

\[
\vec{F}_g = \frac{C_{DG}d_{g,l}}{2\varepsilon} \rho_g |\vec{u}_g| |\vec{u}_l| \quad \text{..............(5)}
\]

\[
\vec{F}_l = \frac{C_{DG}d_{g,l}}{2\varepsilon} \rho_l |\vec{u}_g| |\vec{u}_l| \quad \text{..............(6)}
\]

\[
\vec{F}_s = \varepsilon \rho_l g \quad \text{..............(7)}
\]

where \(\rho_g\) and \(\rho_l\) are densities of gas and liquid, kg/m\(^3\), respectively; \(\vec{u}_g\) and \(\vec{u}_l\) are interstitial velocity vectors of gas and liquid in packed beds, m/s, respectively; \(g\) is gravity vector, m/s\(^2\); \(d_{g,l}\) and \(d_{s,l}\) represent the specific gas–liquid and solid–liquid contact areas, 1/m, respectively; and \(C_{DG}\) and \(C_{DS}\) are defined as gas–liquid drag coefficient and solid–liquid friction factor, (−), respectively.

As indicated in Eqs. (5) and (6), one must know the gas–liquid drag coefficient and solid–liquid friction factor, and the specific gas–liquid and solid–liquid contact areas for determination of forces \(\vec{F}_g\) and \(\vec{F}_l\). Formulations to estimate such parameters for a continuous and saturated liquid flow in packed beds have been reported in open literature.\(^{23,24}\) In the current case where the discrete and unsaturated liquid flow exists, however, it is extremely difficult to calculate the specific gas–liquid and solid–liquid contact areas due to the undetermined liquid diameters (size) for the liquid droplets or rivulets. For simplification, an average effective liquid diameter is conceptually used in the current study and is assumed to be proportional to the effective capillary size in the packed bed, giving by

\[
d_i \propto \frac{\varepsilon}{1 - \varepsilon} \phi d_p \quad \text{or} \quad d_i \propto \frac{\varepsilon}{1 - \varepsilon} \phi d_p \quad \text{..............(8)}
\]

Additionally, assuming that there are many individual liquid droplets or rivulets in the packed bed, the specific gas–liquid and solid–liquid contact areas should be proportional to the total surface areas of the liquid. This implies the following correlation, i.e.:

\[
a_{g,l} \propto \frac{1}{d_i} \quad \text{or} \quad a_{g,l} = \frac{1}{d_i} \frac{1}{d_i} \quad \text{..............(9a)}
\]

\[
a_{s,l} \propto \frac{1}{d_i} \quad \text{or} \quad a_{s,l} = \frac{1}{d_i} \frac{1}{d_i} \quad \text{..............(9b)}
\]

where \(\varepsilon\) and \(\varepsilon_i\) is porosity of the packed bed and volume fraction of liquid in the packed bed, (−), respectively; \(d_p\) and \(d_{g,l}\) are capillary size in the packed bed, giving by

\[
d_i \propto \frac{\varepsilon}{1 - \varepsilon} \phi d_p \quad \text{or} \quad d_i \propto \frac{\varepsilon}{1 - \varepsilon} \phi d_p \quad \text{..............(8)}
\]

Additionally, assuming that there are many individual liquid droplets or rivulets in the packed bed, the specific gas–liquid and solid–liquid contact areas should be proportional to the total surface areas of the liquid. This implies the following correlation, i.e.:

\[
a_{g,l} \propto \frac{1}{d_i} \quad \text{or} \quad a_{g,l} = \frac{1}{d_i} \frac{1}{d_i} \quad \text{..............(9a)}
\]

\[
a_{s,l} \propto \frac{1}{d_i} \quad \text{or} \quad a_{s,l} = \frac{1}{d_i} \frac{1}{d_i} \quad \text{..............(9b)}
\]

where \(\varepsilon\) and \(\varepsilon_i\) is porosity of the packed bed and volume fraction of liquid in the packed bed, (−), respectively; \(d_p\) and \(d_{g,l}\) are capillary size in the packed bed, giving by

\[
d_i \propto \frac{\varepsilon}{1 - \varepsilon} \phi d_p \quad \text{or} \quad d_i \propto \frac{\varepsilon}{1 - \varepsilon} \phi d_p \quad \text{..............(8)}
\]

Additionally, assuming that there are many individual liquid droplets or rivulets in the packed bed, the specific gas–liquid and solid–liquid contact areas should be proportional to the total surface areas of the liquid. This implies the following correlation, i.e.:

\[
a_{g,l} \propto \frac{1}{d_i} \quad \text{or} \quad a_{g,l} = \frac{1}{d_i} \frac{1}{d_i} \quad \text{..............(9a)}
\]

\[
a_{s,l} \propto \frac{1}{d_i} \quad \text{or} \quad a_{s,l} = \frac{1}{d_i} \frac{1}{d_i} \quad \text{..............(9b)}
\]

where \(\varepsilon\) and \(\varepsilon_i\) is porosity of the packed bed and volume fraction of liquid in the packed bed, (−), respectively; \(d_p\) and \(d_{g,l}\) are capillary size in the packed bed, giving by

\[
d_i \propto \frac{\varepsilon}{1 - \varepsilon} \phi d_p \quad \text{or} \quad d_i \propto \frac{\varepsilon}{1 - \varepsilon} \phi d_p \quad \text{..............(8)}
\]

Additionally, assuming that there are many individual liquid droplets or rivulets in the packed bed, the specific gas–liquid and solid–liquid contact areas should be proportional to the total surface areas of the liquid. This implies the following correlation, i.e.:

\[
a_{g,l} \propto \frac{1}{d_i} \quad \text{or} \quad a_{g,l} = \frac{1}{d_i} \frac{1}{d_i} \quad \text{..............(9a)}
\]

\[
a_{s,l} \propto \frac{1}{d_i} \quad \text{or} \quad a_{s,l} = \frac{1}{d_i} \frac{1}{d_i} \quad \text{..............(9b)}
\]
and \( \phi \) are diameter of packing particles, mm, and particle shape factor, \((-\)) respectively; and \( \xi_1 \), \( \xi_2 \) and \( \xi_3 \) are proportional coefficients, \((-\)).

Substituting the Eqs. (5) to (9b) into Eq. (4) and then letting \( C'_{DG} = (\xi_2 / \xi_1) C_{DG} \) and \( C'_{DS} = (\xi_3 / \xi_1) C_{DS} \) give:

\[
\frac{C'_{CD}}{2} \frac{1 - \epsilon}{\epsilon \phi d_p} \rho_g |u|_g + \frac{C'_{CD}}{2} \frac{1 - \epsilon}{\epsilon \phi d_p} \rho_l |u|_l + \rho_d g = 0
\]

\[\text{Eq. (10)}\]

where \( C'_{DG} \) and \( C'_{DS} \) are defined as the modified gas drag coefficient and the modified solid–liquid friction factor, respectively. The \( C'_{DG} \) and \( C'_{DS} \) reflect the interactions between gas and liquid, and liquid and packing particles under unsaturated condition of the discrete liquid through packed beds. \( C'_{DS} \) has been studied in detail elsewhere.\(^{25}\) The experiments in the current study are specially designed for investigation of gas–liquid interaction. Therefore the below discussion only focuses on how to determine the modified gas–liquid drag coefficient that provides better understanding of gas–liquid interaction under the condition of discrete liquid flow in the lower part of blast furnace.

As indicated in Eq. (3), the liquid shift angle \( \alpha \) can be evaluated using magnitudes of gas drag force and gravitational force. Thus combining Eq. (3) with Eqs. (5) to (10) gives:

\[
\tan \alpha = \frac{C'_{CD}}{2} \frac{1 - \epsilon}{\epsilon \phi d_p} \left( \frac{\rho_g}{\rho_l} \right) \left( \frac{u^2_g}{(\phi d_p) g} \right)
\]

\[\text{Eq. (11)}\]

Note that the interstitial velocity \( u_g \) is converted from the superficial gas velocity dividing by porosity of the packed bed.

Eq. (11) implies a plot of \( \tan \alpha \) against the dimensionless number group \( \frac{1}{2} \left( \frac{1 - \epsilon}{\epsilon} \right) \left( \frac{\rho_g}{\rho_l} \right) \left( \frac{u^2_g}{(\phi d_p) g} \right) \) should yield a straight line with a slope \( C'_{DG} \) of which passes through the origin. This means that the modified gas drag coefficient is mainly related to the contributions of gas flow, density of liquid and gas, packing structure and particle size.

Figure 10 shows all the liquid shift data versus the dimensionless number group \( \frac{1}{2} \left( \frac{1 - \epsilon}{\epsilon} \right) \left( \frac{\rho_g}{\rho_l} \right) \left( \frac{u^2_g}{(\phi d_p) g} \right) \) for gas–liquid two-phase flow. The experiments that were carried out to create these data included the packed beds consisting of 9 mm, 15 mm and 19.4 mm wax spheres, using small-scale apparatus with the bed height of 200 mm and large-scale apparatus with the bed height of 410 mm, respectively. The liquid flowrate was 200 mm/min and the gas superficial velocity was varied from 0.472 m/s to 1.18 m/s. Results are shown for all six liquids in Table 1. The results show that most of the data falls on a single straight line, indicating a single, constant value for the modified gas drag coefficient as mentioned above. A linear least squares fit to all data, constrained through the origin gives \( C'_{DG} = 5.4 \pm 1.0 \). Noticeably, the gas–liquid interaction also does not vary with gas flowrate only, but is a complex function of the gas and liquid properties, the packing structure and the gas flow condition. Current experiments cover the fully turbulent region for the gas flow and quite a range of liquid physical properties and packing structure. The experimental data have been tested only against the low surface tension and relatively lower density liquids when compared to the hot metal in blast furnace. However, previous experimental measurements have also identified that there is an approximate estimation value for the gas–liquid drag coefficient in the event of using the high-density liquids such as mercury.\(^{9}\) A number of previous experimental data obtained using an X-ray photograph technique are also included in Fig. 10 for comparison. The data cover three liquids including mercury and two BaCl\(_2\) solutions with a 200 mm height packed bed using two particle sizes of 5.1 mm and 4 mm, respectively. It can be seen from Fig. 10 that the data also fall to the single straight line determined by the current experimental measurements. These results suggest therefore that a reasonable accuracy can be achieved by extrapolating current gas–liquid drag coefficient into an actual blast furnace model to simulate the gas–liquid two-phase interaction.

4. Conclusions

The study has examined the horizontal shift of a percolating liquid flow due to interaction with gas cross flow. It is found that liquid shift angle does not significantly increase or decrease with different packing depth. This finding supports the hypothesis of the force balance model where a vectorial relationship among acting forces, i.e. gas drag force, gravitational force and solid–liquid friction force, and liquid shift angle does exist. Liquid shift angle is inversely proportional to particle size and liquid density, and proportional to square of gas superficial velocity, but is almost independent on liquid flowrate and liquid viscosity. Experimental measurements suggest that the gas–liquid drag coefficient is approximately a constant, i.e. \( C'_{DG} = 5.4 \pm 1.0 \), and is independent on liquid properties, gas velocity and packing structure. The result shows a good agree-
ment with previous experimental data and prediction of the existing liquid flow model.

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Nomenclature

- $a_g$: Specific gas–particles contact area (m$^{-1}$)
- $a_s$: Specific liquid-particles contact area (m$^{-1}$)
- $C_{DG}$, $C_{DS}$: Gas–liquid drag coefficient and modified gas–liquid drag coefficient
- $d_p$, $d_l$: Diameters of particles and liquid droplets or rivulets, respectively (m)
- $F_g$: Gas drag force (N/m$^3$)
- $F_s$: Frictional force between liquid and packed bed (N/m$^3$)
- $g$, $g^0$: Gravity vector and gravitational acceleration (m/s$^2$)
- $L_i$: Collected liquid flowrate via the $i$-th liquid collection box (see Fig. 1) (g/s)
- $u_g$, $u_l$: Interstitial velocity vectors of gas and liquid, respectively (m/s)
- $\vec{u}_g$, $\vec{u}_l$: Interstitial velocity of gas and liquid, respectively (m/s)
- $\bar{x}$: Mean horizontal shift defined by Eq. (1) (mm)
- $\bar{x}_i$: Horizontal shift in Eq. (1), associated with the $i$-th liquid collection box (mm)

Greek letters

- $\alpha$: Liquid shift-angle defined by Fig. 3 and Eq. (3) (deg.)
- $\varepsilon$: Porosity of the packed bed (—)
- $\varepsilon_i$: Volume fraction of liquid in the packed bed (—)
- $\phi$: Shape factor of particles (—)
- $\rho_g$: Density of gas (kg/m$^3$)
- $\rho_l$: Density of liquid (kg/m$^3$)
- $\sigma$: Standard deviation of the measured liquid distribution shown in Eq. (2) (mm)

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