Mass transfer characteristics in double-contact-flow absorber with liquid column/screen flow type: Modeling and experiment

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Abstract

Double-contact-flow absorbers with liquid column/screen flow type are investigated experimentally using systems of different length scales, where SO2 is the absorbate and limestone slurry is the sorbent. The study of the flow characteristics shows that the practical bed height coefficient can be expressed as an exponential function of the Froude number and essentially insensitive to the spray pipe pattern and the scale of system. The main parameters influencing the desulfurization rate include the liquid flow rate, flue gas flow rate, and different spray pipe patterns; their effects are tested on systems with two different scales. Under moderate working conditions, the desulfurization efficiency can reach a high level of above 95%. A method is established to describe the detailed mass transfer process in a double-contact-flow reactor, which may be applied to absorbers of different scales. An empirical formula is derived to predict the desulfurization rate which is in good agreement with experiment data, making it a potential tool for estimating the desulfurization performance in practical applications.

1. Introduction

Reducing SO2 emission from power plants is an outstanding issue for environmental sustainability. Many flue gas desulfurization (FGD) processes are available for the reduction of SO2 emission, which may be categorized as dry-, semidry-, and wet-processes, respectively. Among these processes, the wet-process has a lower operating cost and its operation is more stable, thus more promising and widely employed (Srivastava and Jozewicz, 2001; Srivastava et al., 2001). In a typical setup of wet scrubber, SO2 is removed from the flue gas by reacting with the absorbent containing limestone and gypsum. In order to be effectively employed in large scale power plants, the working mechanism of wet FGD needs to be kept simple, with convenient operation, low cost, and high efficiency. Several possible designs, including the packed column (Kiil et al., 1998; Kiil et al., 2002), spray tower (Eden and Luckas, 1998; Michalski, 1997), bubbling bed reactor (Adewuyi et al., 1999; Lancia et al., 1994), and impinging stream reactor (Krasinski and Warych, 2008; Yuan et al., 2007), have been proposed and tested. The corresponding mathematical models of these devices were established during the last decade (Akbar and Ghaasiaan, 2004; Berman et al., 2000a, b; Frandsen et al., 2001; Liu and Xiao, 2006).

Another promising design for wet FGD is the double-contact-flow absorber (Cheng et al., 2005; Fujimori and Nakashoji, 2003; Ukawa et al., 1996). A schematic is shown in Fig. 1, which consists of the following main elements: (1) gas generator; (2) double-contact-flow absorber; (3) slurry tank. Comparing with other wet FGD equipments, the advantages of the double-contact-flow absorber include:

1. The absorption setup is simple: there is essentially no internal component except for the spray pipes with nozzles at the bottom, hence there is little risk of fouling.
2. The liquid is jetted from the bottom to the top with a controlled practical bed height, and then the liquid falls back to the tank; therefore, the contact time between the gas and the liquid is increased.
3. A high desulfurization rate can be obtained by increasing the liquid–gas ratio.

In our previous study (Hou, 2005; Hou et al., 2006), inside the double-contact-flow absorber, four different types of flow were identified experimentally with increase of the liquid gas ratio, i.e., the liquid-column, liquid-screen, surging, and quasi-bubble (the latter three types of flow were reported for the first time in Hou (2005)). Schematics of the four flow patterns are shown in Fig. 2(a), and the flow pattern maps are shown in Fig. 2(b) and (c). Conventionally, the co-current flow indicates the flow direction of gas is the same as the falling liquid, and the counter-current flow denotes that the flow direction of gas is opposite to the falling
The quasi-bubble flow and surging flow are unstable, and their operating parameters are not suitable for the desulfurization system, and thus these two types should be avoided. Nevertheless, the other two types, namely the liquid-column and liquid screen flows, are more welcome for practical application, and thus they are studied in this paper.

The purpose of this work is to further investigate the flow characteristics and absorption processes of SO2 in a double-contact-flow device developed for the industrial application of wet FGD. This experimental study focuses on the mechanism of mass transfer, and proposes a method which could effectively describe and predict the mass transfer process.

2. Experimental system

The schematic of experimental system is shown in Fig. 1 with the following details:

1. Gas generator: the flue gas simulant is produced by oil combustion with CS2 or pulverized sulfur as an additive, and the SO2 load can easily be adjusted.
2. Double-flow-contact absorber: the sulfur dioxide phase is removed by the reaction between the gas and the liquid in the absorption reactor. The liquid-column flow or the liquid-screen flow is formed with the slurry jetted into the reactor, via the spray nozzles shown in Fig. 3. The two phases contact by the co-current manner first and then by the counter-current manner. Since atomization is not a main function of the nozzle, the structure/design of each nozzle can be made simple with low energy consumption and easy replacement (in case of wear or erosion).
3. Slurry tank: its main function is to circulate the absorbent for the absorber and supply fresh air for oxidation. The flue gas mixed with the oxidation air passes through the top of the tank and enters the absorber. The absorbent is a suspension mixture of finer CaCO3 particles (with the average particle diameter of 21.4 μm); these parameters are chosen for improved dispersion of particles in liquid, which may lead to better reaction characteristics (Shih et al., 2000; Sun et al., 2007).

In this study, the flow characteristics of absorber are investigated using three devices with different scales, with spray nozzle diameters of 1, 9, and 20 mm, respectively. The SO2 absorption characteristics are tested on two systems with different scales, where the maximum practical bed heights are 8 and 1 m (Fig. 3(a) and (b)) with nozzle diameters of 1 mm and 9 mm, respectively. The experimental study with a 20 mm nozzle is carried out in open air. The spray pipe pattern is defined by the parameter $s/d_n$, where $s$ is the space between two main pipes and $d_n$ is the nozzle diameter. In this study, $s/d_n$ varies from 20 to 12 for the small scale device with $d_n=1$ mm, where $s/d_n=20, 15, and 12$ means 9, 16, and 25 jets in the absorber, respectively. The $s/d_n$ value is fixed at 13.6 on the large scale device with $d_n=9$ mm since it is difficult and inconvenient to change the nozzle pattern. Fig. 3(c) shows an example with the spray pipe comprising 16 nozzles and with $s/d_n=15$. In the mass transfer experiment, the concentration of SO2 is measured by a Fourier Infrared gas analyzer (GASMET FTIR Dx4000). The tank's pH is kept constant at about 5.5 by adding fresh limestone slurry.

3. Results and discussion

3.1. Flow characteristic in the absorption tower

The practical bed height formed by the liquid jet is defined by the follow equation:

$$h_l = K_t \frac{u^2}{2g}$$  \hspace{1cm} (1)
where \( K_h \), which must be < 1, is the practical bed height coefficient (PBHC) to be determined in experiment. The jet velocity can be calculated from the liquid flow rate \( Q_l \) as

\[
v_j = \frac{4Q_l}{\pi d_n^4}
\]  

(2)

If there is no resistance, the vertical liquid jet with an initial jet velocity \( v_j \) should reach the theoretical bed height of \( v_j^2 / 2g \) (where \( K_h = 1.0 \)).

In the absorber, the liquid flow is mainly influenced by the gravity and jet inertia, and a dimensionless factor (the Froude number, referred to as Fr below) is chosen as a variable to describe the jet flow property:

\[
Fr = \frac{v_j}{\sqrt{gd_n}}
\]  

(3)

Fig. 4 shows the relationship between the PBHC (\( K_h \)) and Fr number measured from experiments using different absorber scales. When Fr is very small, i.e., when the jetting velocity is low or the nozzle diameter is large, our experimental results show that the energy conversion ratio (from kinetic energy to potential energy) is high and \( K_h \) is near 1.0. Meanwhile, when Fr increases (higher jet velocity and/or smaller nozzle diameter), \( K_h \) decreases rapidly. Indeed, when Fr is high, the liquid is observed to break up into many small drops—since part of the kinetic energy transfers to the surface energy of the small drops, the potential energy of the liquid is lower; moreover, the movement of small liquid drops can be retarded by gas. Subsequently, the PBHC is reduced when Fr increases. For the three nozzle diameters used in different designs, the experimental data conform roughly to a similar trend, which can be fitted as an exponential form:

\[
K_h = e^{-0.0246Fr}
\]  

(4)

with an average relative error of 28%. This equation may be effectively employed to describe the practical bed height of a double-contact-flow absorber. It has been demonstrated that Eq. (4) can be used whether the liquid is pure water or the slurry (with a solid mass concentration below 30%).

Fig. 2. (a) Four types of flow patterns in double-contact-flow absorber; (b) the flow pattern map with \( R_{e_g} \) when the co-current flow is used in the absorber; (c) the flow pattern map with \( R_{e_g} \) when the counter-current flow is used in the absorber.
3.2. Effects of parameters on desulfurization efficiency

Table 1 gives the range of operating parameters and experimental conditions in the two systems with different scales (i.e., with nozzle diameter 1 and 9 mm). The properties of the limestone are presented in Table 2. The reaction of absorption of all SO2 takes place in a wet FGD system and can be summarized in the overall reaction:

\[
\text{CaCO}_3(s) + \text{SO}_2(g) + \frac{1}{2} \text{O}_2 + \text{H}_2\text{O} \rightarrow \text{CaSO}_4 \cdot \text{2H}_2\text{O} + \text{CO}_2(g)
\] (5)

Detailed descriptions of the reactions and other phenomena have been extensively reported, e.g., in Jiang et al. (2006), Lancia et al. (1995), Pasiukbronikowska and Rudzinski (1991). On the basis of the measurements made in this study, the desulfurization efficiency of the system (for absorbing SO2) is calculated as

\[
\eta\% = \frac{C_{\text{SO}_2,\text{in}} - C_{\text{SO}_2,\text{out}}}{C_{\text{SO}_2,\text{in}}} \times 100\% \tag{6}
\]

where \(C_{\text{SO}_2,\text{in}}\) is the SO2 ratio in the inlet gas, ppmv; and \(C_{\text{SO}_2,\text{out}}\) is the SO2 concentration at the exit of the absorber, ppmv.

The main results of the experiments are presented in Figs. 5–8. Fig. 5 shows the influence of slurry concentration (mass fraction of limestone) on the absorption efficiency of SO2. The data show that the solid concentration does not significantly influence the desulfurization efficiency in the absorber (although it affects the retention time in the tank, power wastage, and reliability of wet FGD operation, Srivastava et al., 2001). The effect of the variation of the inlet SO2 concentration in flue gas is shown in Fig. 6, which indicates that the desulfurization efficiency is somewhat lower when the inlet SO2 concentration increases. According to Ruhland et al. (1991), the reaction process is limited by the gas diffusion stages and the characteristic of slurry (sorbent dissolution). In our experiment, the pH is kept between 5 and 6, same as that in engineering application, and the desulfurization efficiency is dependent on the SO2 diffusion through the flue gas. Meanwhile, if the liquid flow rate \(Q\) increases, the curves become closer and thus the desulfurization efficiency becomes less sensitive to the SO2 concentration.

The main parameters influencing the desulfurization efficiency are the flue gas flow rate, slurry flow rate, and spray pipe patterns. Note that due to operational and economical concerns, in this study, the effects of slurry concentration, spray pipes, and SO2 inlet concentration are investigated only in the small scale system. The influence of gas flow rate is given in Fig. 7: when the gas flow rate is fixed, the desulfurization efficiency increases with the increasing of liquid flow rate, because there is more contact area and contact time for absorption. On the other hand, when the flow rate of the circulating slurry is fixed, the desulfurization efficiency decreases with the increase of gas flow rate due to the reduced diffusion time between gas and liquid.

In engineering application, the liquid–gas ratio (L/G, L/m³) of slurry flow rate to flue gas volume flow rate is often employed as a
A higher liquid density or a higher bed height makes the absorption efficiency higher. Under the fixed \( s/d_n \) value, a higher \( L/G \) value means that more liquid is jetted into the absorber when the gas flow rate is fixed; in this case, both the liquid density and the bed height are higher which could make the absorption efficiency higher. However, if the value of \( L/G \) is fixed (i.e., with fixed gas and liquid flow rates), when \( s/d_n \) is small, the absorption efficiency will be lower.

### Table 1
Range of the operating parameters and experimental condition in small system.

<table>
<thead>
<tr>
<th>System</th>
<th>Flow rate of gas ( Q_g ) (m(^3)_s(^{-1}))</th>
<th>Temperature of gas ( T_{in} ) (°C)</th>
<th>Concentration of SO(<em>2) in flue ( C</em>{SO2,in} ) (ppmv)</th>
<th>Flow rate of circulating slurry ( Q_l ) (m(^3)_s(^{-1}))</th>
<th>Density of slurry ( \rho ) (kg m(^{-3})) (×10(^3))</th>
<th>pH</th>
</tr>
</thead>
<tbody>
<tr>
<td>Small scale</td>
<td>0.00472–0.01158</td>
<td>60–80</td>
<td>1500–3000</td>
<td>0–1.95 × 10(^{-4})</td>
<td>≈ 1.1</td>
<td>5–6</td>
</tr>
<tr>
<td>Large scale</td>
<td>0.4178–0.9724</td>
<td>80–100</td>
<td>≈ 5000</td>
<td>0.005–0.0165</td>
<td>≈ 1.1</td>
<td>5–6</td>
</tr>
</tbody>
</table>

### Table 2
Properties of limestone used as absorbent.

<table>
<thead>
<tr>
<th>Components</th>
<th>CaO</th>
<th>MgO</th>
<th>Al(_2)O(_3)</th>
<th>Fe(_2)O(_3)</th>
<th>Others</th>
<th>Mn(^{2+})</th>
<th>Cd(^{2+})</th>
<th>Hg(^{2+})</th>
<th>Pb(^{2+})</th>
</tr>
</thead>
<tbody>
<tr>
<td>Contain (wt%)</td>
<td>51.30</td>
<td>2.47</td>
<td>0.13</td>
<td>0.36</td>
<td>5.72</td>
<td>1.5 × 10(^{-5})</td>
<td>&lt; 10(^{-5})</td>
<td>&lt; 10(^{-5})</td>
<td>&lt; 10(^{-5})</td>
</tr>
</tbody>
</table>

### Fig. 5
Influence of the absorbent concentration on the desulfurization efficiency \((s/d_n=15, Q_g=0.01158\, m^3\, s^{-1}, C_{SO2,in}=3000\, ppmv)\).

### Fig. 6
The desulfurization efficiency varying with different inlet SO\(_2\) concentrations \((s/d_n=12, C_{CaCO3}=20\%, C_{SO2,in}=3000\, ppmv)\).

### Fig. 7
The desulfurization efficiency varying with different gas flow rates \((s/d_n=12, C_{CaCO3}=20\%, C_{SO2,in}=3000\, ppmv)\).

### Fig. 8
Influence of the spray pipe pattern on the desulfurization efficiency, with the value of \( L/G \) as the x axis in two system scales: (a) \( d_n=1\, mm, Q_g=0.01158\, m^3\, s^{-1}, C_{CaCO3}=20\%, C_{SO2,in}=3000\, ppmv\); (b) \( d_n=9\, mm, C_{CaCO3}=20\%, C_{SO2,in}=5000\, ppmv\).
the number of nozzles increases; in this scenario, the total area of nozzles increases and the liquid jet velocity decreases; in other words, despite the higher liquid density in the cross section, the bed height is in fact lower. Considering these factors, if the L/G value is fixed, there should be an optimum value of \( s/d_n \) so that the best absorption efficiency can be obtained. Meanwhile, during the experiment, the desulfurization efficiency can reach 95\% when the L/G value is about 20L/m\(^3\), and we have further shown that for a large range of variation of the liquid–gas ratio, the absorption efficiency is very good with a proper value of bed height at \( s/d_n=15 \) (Fig. 8).

3.3. Resistance of the liquid-screen reactor

An important characteristic of the absorber is the hydraulic resistance. In a double-contact-flow absorber, the liquid has two flow directions, up and down. And the liquid density in the cross section will increase with the increase of nozzle numbers. Therefore, the resistance will increase with the liquid flow rate. The resistance in different scales of absorbers can be obtained by measuring the pressure drop \( \Delta p \), which is caused by the absorber itself and the liquid jetted into the absorber:

\[
\Delta p = (\xi_t + \xi_l) \rho g v^2
\]

where \( v \) is the average gas velocity in the absorber, \( \xi_t \) and \( \xi_l \) are the resistant coefficients of the absorber and the liquid, respectively. The pressure drop is measured by a silicon piezoresistive pressure sensor. It should be noted that the main resistance was caused by the liquid in the present study. In Eq. (7), the absorber structure resistance coefficient \( \xi_t \) can be measured in an empty absorber (without liquid), subsequently, the liquid-column/screen resistance coefficient \( \xi_l \) can be separated from the total resistance. The values of \( \xi_l \) with different spray pipes and different scale equipments are given in Fig. 9. With the same gas flow rate and liquid flow rate, more liquid presents in a cross section of the absorber when \( s/d_n \) is small; in this scenario, the contact area between the gas and the liquid increases, and the interaction between the gas and liquid becomes stronger, and thus the liquid-screen resistance coefficient is larger. If the liquid–gas ratio L/G is more than 40L/m\(^3\), the liquid-screen resistance coefficient can reach 300, but it is still smaller than the results of impinging liquid column tower (Cheng et al., 2005) where \( \xi_l \) may reach 400 according to Eq. (7). In an atomization spray scrubber, the resistance could be much higher (Berman et al., 2000a). For different spray pipes and \( s/d_n \) values studied in the current experiment, an empirical equation for the liquid-screen resistance coefficient in the double-contact-flow reactor can be fitted as

\[
\xi_l = 2.356 \left( \frac{L}{C} \right)^{1.31}
\]

In the above equation, the liquid–gas ratio \( L/G = 0–40L/m^3 \).

4. Modeling and results of mass transfer

The results presented above demonstrate that the desulfurization efficiency is influenced by many factors. Among them, the gas flow rate, liquid flow rate and spray pipe patterns are most important, whereas the slurry concentration and the SO\(_2\) inlet concentration are relatively less important. In order to clarify the mechanism of desulfurization and predict its efficiency in a more quantitative manner, we will now develop a model to describe the mass transfer process in a double-contact-flow absorber.

A mass transfer coefficient of the system is defined by the following equation:

\[
k = \frac{M}{A(C_{SO_2,g} - C_{SO_2,l})}
\]

where \( M \) is the total mass transfer rate which can be calculated from the flow rates of SO\(_2\) through the inlet and the outlet of the absorber, \( A \) is the equivalent mass transfer area, and \( C_{SO_2,g} \) is the mole concentration in the gas phase. In the liquid, where SO\(_2\) transforms to SO\(_2^2\), \( C_{SO_2,l} \) can be considered to be zero.

In the nozzle spray systems, there are various flow patterns that may affect the contact between gas and liquid, for example, the liquid may appear as a column, single droplets, and liquid blocks. And since the interface constantly changes with time, it is very difficult to measure or calculate the accurate contact area between gas and liquid. Conceptually, several parameters of the absorber structure and liquid jets should contribute to the mass transfer area: First, \( A \) should depend on the number of liquid columns (\( n \)), so that there is more contact area when more nozzles are employed. Secondly, \( A \) should depend on the diameter of the column (\( d_n \)). Thirdly, \( A \) should become larger when the jetting velocity of liquid (\( v_j \)) is higher, due to the higher column. Therefore, we propose two different approaches to estimate \( A \) in Eq. (9) (both of which meet the above criteria):

\[
A = n \pi d_n \frac{v_j^2}{2g}
\]

Or alternatively,

\[
A = n \pi d_n h_j
\]

The physical meaning of Eq. (10) is that \( A \) equals to the total lateral area of the cylinders whose baseline are the circumferences of the nozzle, \( n \cdot \pi d_n \), and whose height is the theoretical bed height formed by the liquid jet, \( \frac{v_j^2}{2g} \). Although this definition may not represent the actual contact area between the liquid and the gas, it satisfies all three criteria depicted in the last paragraph. In Eq. (11), the height is taken to be the practical bed height formed by the liquid jets, \( h_j \), which is measured from experiment and that is the only difference between Eqs. (10) and (11). Again, the definition in Eq. (11) also satisfies the three characters mentioned above.

Furthermore, according to Eq. (9), a dimensionless mass transfer Sherwood number can be calculated as

\[
Sh = \frac{kd_j}{D_{SO_2}}
\]
where $d_n$ is the characteristic length (the wetted perimeter diameter) of the absorber, and the SO$_2$ mass diffusion coefficient in air is $D_{SO_2} = 1.2 \times 10^{-5}$ m$^2$/s (Guan and Zhao, 2003).

The Reynolds numbers are used to describe the flow characteristics of gas and liquid. In the above equation, the wetted perimeter diameter is taken to be the characteristic length for the gas phase’s Reynolds number ($Re_g$), and the nozzle diameter is chosen as the characteristic length for the liquid phase’s Reynolds number ($Re_l$), that is

$$ Re_g = \frac{v_g d_n}{d_n} \quad Re_l = \frac{v_l d_n}{d_n} \quad (13) $$

After the mass transfer rate $M$, and the concentration of SO$_2$, are obtained from the experiment, the Sh numbers can be calculated according to the above equations, shown in Figs. 10–13. Fig. 10 shows the influence of the liquid Reynolds number on the mass transfer process with different gas $Re_g$ numbers (in the small scale system). If the Sh number is calculated according to the contact area defined by Eq. (10), the curves of $Re_g=4823$, 8498, 12059, 15964, are ordered according to $Re_g$ in Fig. 10(a). However, if the Sh number is calculated from the contact area defined in Eq. (11), some of the curves intersect in Fig. 10(b), indicating a problem with this particular definition. In other words, the results in Fig. 10(a) show better regularity than that in Fig. 10(b), which implies the calculation of $A$ based on Eq. (10) is more reasonable.

In fact, the practical bed height embeds the influence of the contact area of mass transfer. A higher liquid bed height means more liquid drops (or blocks) within the bed, and thus more contact area between liquid and gas. However, another factor governing the contact area, the surface area of liquid drops/blocks, is not directly related to the practical bed height. From an energy point of view, the practical bed height corresponds to the gravitational potential energy, and the surface area of liquid drops/blocks is related to the surface energy. Both energy terms are converted from the kinetic energy of the liquid jets, and the contact area defined by the theoretical bed height in Eq. (10) in fact includes contributions of both. Therefore, in what follows, we will use the contact area calculated by Eq. (10) to derive the mass transfer criterion, the Sh number.

In the large scale system, the variation of Sh number with gas $Re_g$ number is shown in Fig. 11. When $Re_g$ increases, the Sh number increases, and the mass transfer process is improved. Note that for the two different scale systems, dimensionless numbers such as $Re_l$, $Re_g$ and Sh lie in different ranges (Figs. 10 and 11). When the flue gas flow rate is kept at constant, the Sh number strongly depends on the spray pipe pattern. The influence of sorbent concentration and spray pipe pattern is investigated in

![Fig. 10. Mass transfer Sh number is influenced by the gas Reynolds number ($s/d_s = 12$, $C_{CaCO_3} = 20\%$, $C_{SO_2,in} = 3000$ ppmv); (a) the results calculated by Eq. (10); and (b) results calculated by Eq. (11).](image)

![Fig. 11. The mass transfer Sh number is influenced by the gas Reynolds number in a large scale system ($C_{CaCO_3} = 20\%$, $C_{SO_2,in} = 5000$ ppmv).](image)

![Fig. 12. The Sh number is influenced by different spray pipe patterns and absorbent concentration with the increase of liquid $Re_l$ ($Re_g = 9498$, $C_{CaCO_3} = 3000$ ppmv).](image)
the small device, as shown in Fig. 12. With the Re held fixed, there is more liquid jetted into the absorber with the pattern of $s/d_n = 12$. According to Eq. (10) and Fig. 12, the contact area of $s/d_n = 12$ is larger than that of $s/d_n = 15$ and 20. However, the mass transfer rate ($M$ in Eq. (9)) does not increase as quickly as the contact area. Consequently, when $s/d_n$ is smaller, the mass transfer coefficient $k$ in Eq. (9) is smaller, and the curve of $s/d_n = 12$ falls below other curves in Fig. 12. And the Sh number is less insensitive to the sorbent concentration.

The liquid and flue gas flow rates are the two primary parameters which influence the mass transfer process. With the change of the value of L/G, the variations of the Sh numbers from the two systems with different scales are shown in Fig. 13(a) and (b), respectively. The Sh number is insensitive to the inlet flue gas concentration as shown in Fig. 13(a). The curves with different Re values get close to each other, implying that the liquid gas ratio is the dominant parameter in the mass transfer process.

It has been previously demonstrated that the Sh number of a double-contact-flow absorber is relatively insensitive to the sorbent concentration and the inlet flue gas concentration; however, the Sh number depends on the gas and liquid Re number (or liquid–gas ratio), as well as the spray pipe patterns. When the bed expansion coefficient $z_{bed} = (\gamma_l^2/\gamma_g)/d_p$ is incorporated, an empirical relationship of the Sh number can be fitted to the experimental results of both the large and small scale systems

$$Sh = 0.01387z_{bed}^{-1.0704}Re_0^{0.8385}Re_1^{-0.2874} \left(\frac{S}{d_n}\right)^{1.6158}$$

This equation can be used to predict the Sh number or mass transfer coefficient in a double-contact-flow absorber.

### 5. Application

The outlet concentration of SO$_2$ can be calculated for given operation parameters using the following approach. The mass transfer rate is essentially the change of content of SO$_2$ at the inlet/outlet, therefore,

$$M = Q_0(C_{SO_2, in} - C_{SO_2, out}) = kA \cdot C_{SO_2, g}$$

In the above equation, the SO$_2$ concentration in the gas phase is assumed to be

$$C_{SO_2, g} = (C_{SO_2, in} + C_{SO_2, out})/2$$

Substituting Eqs. (10) and (12) into (17), the SO$_2$ concentration in the outlet is

$$C_{SO_2, out} = \frac{1}{d_g} \cdot \frac{ShD_{SO_2}}{d_g} \cdot \frac{4Q_l^2}{\gamma_g \gamma_d Q_g} \cdot C_{SO_2, in}$$

The efficiency of SO$_2$ absorption can be calculated as

$$\eta = \frac{C_{SO_2, in} - C_{SO_2, out}}{C_{SO_2, in}} \times 100 = \frac{2}{1 + \frac{ShD_{SO_2}}{d_g} \cdot \frac{4Q_l^2}{\gamma_g \gamma_d Q_g}} \times 100\%$$

where Sh is obtained from Eq. (14). The results calculated from the above equation are shown in Fig. 14 (y-axis) and compared...
with the experimental data (x-axis). As observed, the empirical model is applicable for various spray pipe patterns, with an error of \( \pm 10\% \). Thus the proposed model can be used to estimate the working conditions (or evaluate the absorption performance) in practical applications.

### 6. Conclusions

We carried out an experimental study to evaluate the desulfurization process in a double-contact-flow absorber with the liquid column/screen flow type. The flow characteristic and mass transfer process are investigated on systems of different scales. The main conclusions include:

1. The practical bed height coefficient is an important parameter. Based on the experimental data from systems with different scales, the coefficient can be expressed as an exponential function (Eq. (4)) of the dimensionless Fr number (which characterizes the jet flow, Eq. (3)). The dimensionless Fr number is determined by the liquid jet velocity computed from the liquid flow rate and nozzle diameter. From our experiments, the resistance coefficient of the liquid follows a power-law function of the liquid–gas ratio (Eq. (8)).

2. The double-contact-flow absorber with liquid-column/screen flow type demonstrates good performance for wet FGD. Under the conditions employed, the desulfurization efficiency reached as high as 95%. Two important parameters can be recommended for optimal operation with the double-contact-flow absorber:

   1. spray pipe pattern
   2. liquid gas ratio L/G

   In addition, based on the practical height limit for a given absorber, the optimum diameter of nozzles can be selected based on Eqs. (1) and (2).

3. A method for estimating the contact area between the gas and the liquid is established (based on the theoretical bed height, Eq. (10)); the Sherwood number Sh is used to describe the mass transfer process for the absorber (Eq. (12)). In the context of Sh, the main parameters affecting the mass transfer are \( \text{Re}_g \), \( \text{Re}_b \) (or liquid gas ratio), and the spray pipe pattern. Based on experimental data of systems with two different scales, an empirical relation (Eq. (14)) is obtained for the Sh number and its variation with the gas and liquid Re number and spray pipe pattern. With the above model, a predictive equation of the desulfurization rate is derived (Eq. (19)) which agrees within \( \pm 10\% \) of experimental data.

It is clearly demonstrated that the double-contact-flow absorber with different spray pipe patterns has a number of advantages. Despite the progresses in this paper, several questions still remain unclear, such as the relationship between the flow fluctuations and mass transfer in the absorber, which require future investigations.

### Notation

- \( A \) contact area between gas and liquid in the absorber, m\(^2\)
- \( C_{\text{CaCO}_3} \) mass concentration of limestone in absorber, %
- \( C_{\text{SO}_2,g} \) average concentration of \( \text{SO}_2 \) in gas phase, mol/m\(^3\)
- \( C_{\text{SO}_2,n} \) \( \text{SO}_2 \) ratio in inlet gas, ppmv
- \( C_{\text{SO}_2,l} \) average concentration of \( \text{SO}_2 \) in liquid phase, mol/m\(^3\)
- \( C_{\text{SO}_2,\text{out}} \) \( \text{SO}_2 \) ratio in the exit of the absorber, ppmv
- \( d_b \) wetted perimeter diameter for absorber, m
- \( d_n \) diameter of nozzle, m
- \( D_{\text{SO}_2} \) mass diffusion coefficient of \( \text{SO}_2 \), m\(^2\)/s
- \( \text{Fr} \) Froude number of single nozzle
- \( g \) acceleration of gravity, m/s\(^2\)
- \( h_j \) practical bed height formed by liquid jets, m
- \( k \) mass transfer coefficient, m/s
- \( k_h \) practical bed height coefficient
- \( L/G \) liquid gas flow rate ratio L/m\(^3\)
- \( n \) number of all the nozzles in use
- \( \Delta p \) pressure drop in the absorber, Pa
- \( Q_g \) volumetric flow rate of the flue gas, m\(^3\)/s
- \( Q_l \) volumetric flow rate of the slurry, m\(^3\)/s
- \( \text{Re}_g \) Reynolds number of flue gas in the absorber defined in Eq. (13)
- \( \text{Re}_b \) Reynolds number of single jet in the absorber defined in Eq. (13)
- \( s \) distance between two main pipes, m
- \( Sh \) Sherwood number for mass transfer process defined in Eq. (12)
- \( v_g \) average flue gas velocity in the empty absorber, m/s
- \( v_j \) liquid jet velocity of single nozzle calculated from the liquid flow rate, m/s

### Greek letters

- \( \eta \) desulfurization efficiency, %
- \( v \) kinematic viscosity of gas phase, m\(^2\)/s
- \( \eta \) kinematic viscosity of liquid phase, m\(^2\)/s
- \( \zeta \) bed expansion coefficient
- \( \zeta_b \) resistance coefficient caused by the liquid
- \( \zeta_t \) resistance coefficient of the absorber

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### References


