Hydrodynamic Characterization and Mass Transfer Analysis of an In-Line Multi-Jets Ozone Contactor

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This research study investigates mixing and ozone mass transfer characteristics of a pilot-scale in-line multi-jets ozone contacting system. The hydrodynamic characteristics of the contactor were studied using a two-dimensional laser flow map particle image velocimetry coupled with planar laser induced fluorescence (PIV/PLIF). The PIV/PLIF system provided a combination of simultaneous whole-field velocity and concentration data in two-phase flows for different operating conditions. All measurements were conducted under a total liquid flow rate of about 10 L/s with gas flow rate ranging from 0.05 to 0.4 L/s. The gas was introduced to the system through a series of side stream injectors. The side injectors were tested under opposing and alternating modes. A mass transfer study was also conducted to estimate the overall mass transfer coefficient under the same operational conditions used for the hydrodynamics study. It was found that for the same number of jets (i.e., same gas flow rate) the liquid dispersion (DL) was higher when alternating jets were used. Higher ozone mass transfer rates were observed when using opposing jet compared to the same number of alternating jets.

Keywords Ozone, Hydrodynamics, Mixing, Mass Transfer, Dispersion Coefficient, Laser, Particle Image Velocimetry, Planar Laser Induced Fluorescence

INTRODUCTION

Designing an effective reactor requires information, knowledge, and experience from different areas including chemical kinetics, fluid mechanics, and mass transfer (Levenspiel, 1999). However, due to the difficulty of accurately estimating the parameters needed for these areas, empirical and semi-empirical relationships are used in designing such reactors. These relationships are not reliable and/or valid over the reactors’ wide range of practical applications (Joshi, 2002). Furthermore, recent challenges resulting from the existence of micro-pollutants in drinking water supplies (Petrovic et al., 2003) have increased the need for accurately designed advanced water and wastewater treatment reactors.

Hence, a proper design procedure has to be followed in order to reduce the prevailing empiricism associated with conventional water treatment reactors. Joshi and Ranade (2003) suggested the following procedure: (1) identify the desired fluid dynamic characteristics by understanding the process requirements; (2) develop possible reactor configurations/operating protocols to achieve the desired fluid dynamic characteristics; (3) develop quantitative relationships between the reactor configuration and performance; and, (4) optimize and fine-tune the final reactor design.

The use of non-intrusive measurement techniques (i.e., techniques resulting in no direct interaction with the flow field) such as the use of laser measurement systems can provide accurate hydrodynamic characteristics for a reactor. Laser measuring techniques are non-intrusive and directional-sensitive and provide accurate and high-resolution measurements (Albrecht et al., 2002). The laser systems used for characterizing reactors’ hydrodynamics include the laser Doppler anemometer (LDA), the particle dynamics analyzer (PDA), the particle image velocimetry (PIV), and the planar laser-induced fluorescence (PLIF) (Atkinson et al., 2000; Albrecht et al., 2002).

The LDA and PDA systems are commonly used to provide the flow velocity at one point in a flow field (the PDA can also provide a simultaneous measurement of the particles’ sizes) (Durst et al., 1997). The PIV system provides simultaneous planar measurements of the flow velocity by measuring the displacement of seeded particles over a relatively short time interval (Raffel et al., 1998; Atkinson et al., 2000; Bernard
The PLIF system is used to obtain a scalar concentration field in water by introducing a fluorescent dye as a passive scalar in the flow. The dye absorbs incident light, during the illuminating process of the laser, at one wavelength and re-emits it at a different wavelength with an intensity proportional to the dye concentration at the measuring point (Bernard and Wallace, 2002).

The objectives of this study are to: (1) characterize the hydrodynamics of an in-line multi-jets ozone contactor by utilizing PLIF laser measurement techniques, (2) evaluate the flow patterns at different locations along the contactor by using PIV laser measurement techniques (3) analyze the mass transfer efficiency of the contactor, and (4) optimize the system’s operating conditions.

EXPERIMENTAL SETUP

Flow Facility

A pilot-scale ozone contactor is shown in Figure 1. The pipe was made of a round clear acrylic glass tube 0.1 m in diameter \((d)\) and 2 m in length \((L)\). The water flow was driven through the system from a 2 m\(^3\) tank via two lines, main stream and side stream, by utilizing centrifugal pumps. The designed water flow rate \((Q_L)\) for this contactor was 0.01 m\(^3\)/s with a side stream to main stream (liquid to liquid) ratio ranging from 2.5 to 20%. However, other flow rates were also tested to cover a range of Reynolds numbers, \(Re\) (based on the pipe diameter) of 45,000 to 100,000.

The side-stream line had multi jets that employed Mazzei® venturi injectors (model 584, Bakersfield, CA, USA) with N-8 Mazzei® nozzles as shown in Figure 2. The jets were aligned along the axial direction of the contactor in groups of two opposing jets. The axial distance between any two consecutive jets along the axial direction of the contactor was fixed at 0.15 m. These jets allowed for the gas to be introduced into the system in either injection or suction modes. The end of the pipe had two outlets to allow for recycling or draining of the water. An outer square jacket of 0.15 m sides and 1.5 m length was installed around the pipe to reduce optical distortion. A diffuser with a diameter of 0.003 m was installed at the entry region of the pipe at a position 0.3 m downstream from the first side-stream injector. This diffuser was used to inject the tracer for the PLIF runs and the seeding particles for the PIV runs.

When a side jet enters a pipe flow (i.e., a cross-flow), the jet can be characterized by using four geometric/flow parameters: the jet diameter \(D_j\) (m), the pipe diameter \(d\) (m), the jet to pipe velocity ratio \(r\) (\(u_j/u_p\)), and the pipe flow Reynolds number \(Re\) (Forney et al., 1999; Pan and Meng, 2001). The combination of \(rD_j\), known as the jet momentum length \((lm = D_ju_j/u_p, m)\), represents the distance over which the jet travels before it bends over in the cross-flow. In the case of a turbulent pipe flow, it is desired that the jet penetrates into the potential

![Diagram of experimental setup](image)
core of the pipe flow before the jet turns and aligns with the pipe flow.

This regime is referred to as the jet-mixing regime and can be achieved when the dimensionless parameter $r_{Dj}/d$ is within a range from 0.07 to 1.0 (Sroka and Forney, 1989; Pan and Meng, 2001). In the jet-mixing regime, the jet expands quickly in the pipe due to turbulent entrainment and, therefore, creates efficient macromixing (Cozewith and Busko, 1989; Pan and Meng, 2001). If the parameter $r_{Dj}/d$ is greater than 1.0, the jet hits the opposite wall of the pipe, and the impingement helps in creating more efficient micromixing (Tosun, 1987). However, a significant stress on the pipe wall is exerted due to the jet impingement and, thus, such designs are not desirable from a practical point of view (Pan and Meng, 2001). When $r_{Dj}/d$ is smaller than 0.07, the jet attaches to the pipe wall and grows slowly without significant penetration into the pipe flow. This result is undesirable, as the resultant mixing will be inefficient.

Two side-jet types were used in this study: the tracer/seeding particle injection port (0.003 m diameter) and the 1-phase/2-phase side stream nozzles (N-8 Mazzei® nozzles, 0.008 m diameter). The flow rates through these jets were $7.6 \times 10^{-5}$ m$^3$/s and $2.5 \times 10^{-4}$ m$^3$/s (per one nozzle), respectively. The corresponding $r_{Dj}/d$ values for the injection port and the side stream nozzles were 0.24 and 0.35, respectively, which were within the desirable jet-mixing regime.

**PIV/PLIF Setup**

The PIV/PLIF setup used in the hydrodynamic analysis of the contactor consisted of a laser source, charge coupled device (CCD) cameras, and processing units (Figure 1). An Nd:Yag dual cavity laser was utilized in this study for both the PIV and PLIF experiments. The emitted wavelength of the utilized Nd:Yag laser was 532 nm with a pulse duration of 10 ns. The period between pulses was set to 1000 µs during PIV measurements and 100 µs during PLIF measurements with a maximum repetition rate of 8.0 Hz. The measurements were obtained at a time interval of 1000 ms during PIV measurements and 125 ms during PLIF measurements. The CCD cameras were configured to use double frames for PIV measurements (velocity measurements) and a single frame for PLIF measurements (concentration measurements). A FlowMap System Hub® produced by Dantec Dynamics (Skovlunde, Denmark) was used to transfer the data to a PC where FlowMap Software® was used for sequential analysis of the collected data.

The PLIF system was first calibrated by measuring the intensity of 5 different concentrations of Rhodamine 6G (Rh6G) solutions ranging from zero to 500 µg/L at a power level ranging from 50 to 150 mJ. The concentration versus the intensity was plotted to determine the most appropriate calibration curve for this study. The calibration curve obtained for the 150 mJ power gave the highest correlation coefficient.
Therefore, this power level was used for the PLIF measurements.

The PIV measurements were taken by using two CCD cameras with a double-frame mode for measuring the velocity of both phases (liquid and gas) simultaneously by utilizing special filters. Melamine-formaldehyde (MF) spheres, coated with Rhodamine B (RhB), were used as seeding particles to obtain the liquid velocity measurements while gas bubbles represented the seeding particles for the gas velocity measurements. The 2-D velocity vectors were then obtained by employing an interrogation cell of 64 × 64 pixels that was a subdomain of a 1344 × 1024 pixel viewing area. The interrogation cell was then shifted with 25% overlap, and, thus, 21 × 16 velocity vectors were obtained in each instantaneous PIV sample. However, for illustration purposes, 13 × 10 velocity vectors maps were produced.

Gas Mass Transfer Setup

The experimental setup for the mass transfer of the gas in the pilot-scale multi-jets ozone contactor shown in Figure 1 consisted of two dissolved oxygen (DO) probes (model YSI 5750, Yellow Springs, OH, USA) that were placed 1.37 m apart in the contactor (one upstream of the jets and one downstream of the two-phase jets), two DO meters (YSI 50B model) to obtain DO measurements directly, air and nitrogen gas cylinders. The system was arranged to operate under the suction or injection modes of gas through the Mazzei injectors (Figure 2). The gas flow rate during the suction mode was kept constant through each injector by keeping the side stream pressure constant when using any number of injectors and monitoring the flow rates by using gas flow meters. During the injection mode, gas cylinders were used to supply the required amounts of gas by using pressure regulators and gas flow meters.

MEASUREMENT DESCRIPTION

The PLIF and PIV experimental conditions conducted in this study are summarized in Tables 1 and 2, respectively. The pressure in the main flow and side stream was kept constant at 69 kPa (10 psi) and 138 kPa (20 psi), respectively, under the operating conditions shown in Tables 1 and 2. After reaching a steady-state flow condition, a continuous injection of a 12.5 mg/L of the Rh6G tracer at about 7.56 × 10⁻⁵ m³/s (yielding about 95 µg/L average concentration when fully mixed with the total liquid flow rate (QL) of 0.01 m³/s) was introduced into the system through the injection point at the entrance of the contactor, as shown in Figure 1.

<table>
<thead>
<tr>
<th>Experiment number</th>
<th># of jets</th>
<th>Jet alignment*</th>
<th>Total Qₐ (× 10⁻³ m³/s)</th>
<th>Side stream Qₐ (× 10⁻³ m³/s)</th>
<th>Side stream Q₉ (× 10⁻³ m³/s)</th>
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<td>–</td>
<td>8.19</td>
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<tr>
<td>3</td>
<td>–</td>
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<td>8.82</td>
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<tr>
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<tr>
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<td>6</td>
<td>A</td>
<td>1.42</td>
<td>0.30</td>
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<tr>
<td>24</td>
<td>8</td>
<td>O</td>
<td>1.76</td>
<td>0.40</td>
<td>–</td>
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</tbody>
</table>

* O = opposing; A = alternating.
### TABLE 2. Summary of the Operating Conditions During PIV Experiments

<table>
<thead>
<tr>
<th>Experiment number</th>
<th># of jets</th>
<th>Measurement location(^*)</th>
<th>Total (Q_L) ((\times 10^{-3} \text{ m}^3/\text{s}))</th>
<th>Side stream (Q_L) ((\times 10^{-3} \text{ m}^3/\text{s}))</th>
<th>Side stream (Q_G) ((\times 10^{-3} \text{ m}^3/\text{s}))</th>
</tr>
</thead>
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<td>–</td>
<td>E</td>
<td>8.19</td>
<td>–</td>
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<td>3</td>
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<td>E</td>
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<tr>
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<tr>
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<td>E/A</td>
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<td>0.91</td>
<td>–</td>
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<tr>
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<td>E/O</td>
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<td>1.21</td>
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<tr>
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<td>2</td>
<td>M/A</td>
<td>10</td>
<td>0.47</td>
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</table>

\(^*\)E = pipe end; M = mixing zone; O = opposing jets; A = alternating jets.

The continuous (step) input of the tracer was chosen over the slug input for this contactor due to the relatively short theoretical residence time (1.0 to 2.0 s) of the contactor. The effect of increasing Re was studied first by varying \(Q_L\) from \(7.56 \times 10^{-3}\) to \(0.01\) \(\text{m}^3/\text{s}\) without side-stream injection. The effect of 1-phase (liquid phase) side stream injection was investigated by keeping \(Q_L\) at the designed value (\(0.01\) \(\text{m}^3/\text{s}\)) and varying the liquid side stream flow from \(2.5 \times 10^{-4}\) \(\text{m}^3/\text{s}\) (for one jet) to about \(1.8 \times 10^{-3}\) \(\text{m}^3/\text{s}\) (for eight jets).

Then, the system was studied under 2-phase side stream injection by varying the gas flow rate \((Q_G)\) from \(5.0 \times 10^{-5}\) to \(4.0 \times 10^{-4}\) \(\text{m}^3/\text{s}\) (for one jet) to \(4.0 \times 10^{-4}\) \(\text{m}^3/\text{s}\) (for eight jets). Both the 1-phase and the 2-phase, side jets were tested under opposing (jets released to the contactor facing each other) and alternating (jets released to the contactor from the opposite side but apart with a distance of 0.15 m) alignments. The PLIF process of image capturing covered about 5 times the detention time required for the tracer to pass through the system under each operating condition (duplicate measurements were applied to reduce the uncertainty associated with the measurements).

During the PIV measurements, the flow patterns of the system were studied for 1-phase and 2-phase flow conditions as shown in Table 2. The value of \(Q_L\) was varied from \(7.56 \times 10^{-3}\) to \(0.01\) \(\text{m}^3/\text{s}\) while \(Q_G\) ranged from \(5.0 \times 10^{-5}\) \(\text{m}^3/\text{s}\) to \(4.0 \times 10^{-4}\) \(\text{m}^3/\text{s}\). The measurements were taken at several locations along the system (at the pipe end where the PLIF measurement were taken, at the mixing zone of the alternating jets, and at the mixing zone of the opposing jets) in order to evaluate the system’s flow patterns. The PIV process of image capturing was taken in duplicate measurements. To avoid contaminating of the water tank with the tracer dye and the seeding particles, the water was not recycled during the PLIF and PIV measurements.

The gas mass transfer measurements were conducted for the wide range of operating conditions as shown in Table 3. The DO was first stripped out of the water in the tank by the injection of nitrogen gas until the DO level fell below 0.5 mg/L. The water was then recycled through the system for a few minutes by using both the main and side streams, with no injection of gas, to ensure that a steady state was reached and that the DO level was still below 0.5 mg/L. The gas was then introduced into the system through the Mazzei injectors (Figure 2), and the change in DO concentrations was monitored under each operating condition at the two locations. As the flow was recycled to the tank, a pressure drop at the main line to about 41 kPa (6 psi) was observed when the total \(Q_L\) was 0.01 \(\text{m}^3/\text{s}\) and to about 36 kPa (5.3 psi) when \(Q_L\) was 7.6 \(\times 10^{-3}\) \(\text{m}^3/\text{s}\).
RESULTS AND DISCUSSION

Reactor Hydrodynamics

All images captured during the PLIF experiments were converted to 2D concentration fields through the obtained calibration relation by using the FlowMap Software®. A re-sampling of these concentration fields yielded colored contour maps that show the concentration distribution of the tracer along the cross-section of the reactor parallel to the flow direction. A colored contour map representing the concentration distribution at different sampling (image capturing) times for two different liquid flow rates ($Q_L$) of 7.56 × 10^{-3} m^3/s and 0.01 m^3/s without side stream injection is shown in Figure 3. Figure 3a ($Q_L = 7.56 \times 10^{-3}$ m^3/s) shows that segments of the tracer reached the measurement point earlier than those shown in Figure 3b ($Q_L = 0.01$ m^3/s).

Furthermore, a higher cross-sectional (radial) mixing can be observed under the higher liquid flow rate as a relatively uniform concentration of the Rh6G dye can be perceived. The concentration distribution for 1-phase (liquid phase) side stream jets at $Q_L = 0.01$ m^3/s is shown in Figure 4. This figure shows color contour maps for 4 opposing and 4 alternating jets. Some tracer segments appear to have passed through the reactor with the opposing 1-phase jets’ condition slightly faster than the alternating 1-phase jets’ condition. However, a relatively better radial mixing was observed under the opposing jets’ condition. Therefore, further analysis is needed to explain the dispersion under these conditions (see next section).

The concentration distribution of 2-phase (liquid and gas phases) side stream jets at $Q_L = 0.01$ m^3/s and $Q_G$ of 2.0 × 10^{-4} m^3/s is shown in Figure 5 for 4 opposing and alternating jets. This figure shows that segments of the tracer have passed through the contactor with the alternating 2-phase jets relatively faster than the opposing 2-phase jets. In contrast, a relatively higher cross-sectional mixing can be observed in the case of the opposing jets. Therefore, it is expected that the 2 opposing jets will exert a lower axial dispersion compared to that of the same number of alternating jets.

The mixing and the dispersion in the contactor could be further analyzed through velocity measurements obtained from the PIV system for the studied operating conditions. Figure 6 shows the averaged and radial velocity vectors for two different liquid flow rates ($Q_L$ of 7.56 × 10^{-3} m^3/s and 0.01 m^3/s) taken at the pipe’s end (the same location as that of the PLIF measurements). This figure shows that the contactor has not reached a fully developed flow at this stage as the distance from the tracer diffuser to the measurement point is 1.61 m.

The required length to reach a fully developed flow, under the given turbulent conditions ($l_d = 4.4dRe^{1/6}$, m), is in the order of 2.85 m and 3.0 m for $Q_L$ of 7.56 × 10^{-3} m^3/s and 0.01 m^3/s, respectively. Furthermore, Figure 6 also shows that the random radial movement of the flow velocity in the higher flow was more pronounced. This result explains the relative uniformity in the cross-sectional mixing of such a flow rate compared to that of the lower flow rate shown earlier in Figure 3.

The velocity vectors of the liquid-phase in-line multi jets (opposing and alternating), obtained at the mixing zone (the entrance zone of the jets), for $Q_L$ of 0.01 m^3/s are shown in Figure 7. A better cross-sectional (radial) mixing can clearly be observed when the opposing liquid-phase jets are used. Each jet penetrates to about 0.04 m ($l_m = rD_i$) from each side.
before the jets bend, merge into each other, and expand with the cross-flow. Similarly, the opposing gas-phase jets exert better cross-sectional mixing than the alternating ones, as Figure 8 shows. Furthermore, some backmixing was observed in the opposing jets, which was believed to favorably affect the gas transfer rate.

To evaluate the mixing in the contactor numerically, the following differential equation representing the dispersion of a conservative tracer \( C, \mu g/L \) was considered (Levenspiel, 1999):

\[
\frac{\partial C}{\partial \theta} = \left( \frac{D_L}{uL} \right) \frac{\partial^2 C}{\partial z^2} - \frac{\partial C}{\partial z} \tag{1}
\]

where \( \theta \) is a dimensionless time \( (\theta = t/\tau = tu/L, t \) is time (s)), \( \left( \frac{D_L}{uL} \right) \) is the dispersion number (the inverse of the Peclet number, Pe), \( D_L \) is the liquid axial dispersion coefficient \( (m^2/s) \), \( u \) is the pipe flow average velocity \( (m/s) \), \( L \) is the axial distance between the tracer input point and measurement point \( (m) \), and \( z \) is the dimensionless axial distance \( (z = (ut + x)/L, x \) is the axial distance along the pipe \( (m) ) \).

For a step input of a tracer, the shape of the tracer at the measurement point is S-shaped and referred to as “the F-curve.” The F-curves normally represent the dimensionless concentration \( F \), the ratio between the tracer concentration at the measurement point \( (C, \mu g/L) \) to the initial mixed tracer concentration \( (C_0, \mu g/L) \), as a function of time. The shape of the F-curve depends on the boundary conditions of the contactor and the dispersion number \( \left( \frac{D_L}{uL} \right) \). The analytical expressions of the F-curves are not available; however, their graphs can be constructed (Levenspiel, 1999). The value \( \left( \frac{D_L}{uL} \right) \) can be obtained directly by plotting the experimental data on a probability graph paper as explained below or by differentiating the S-shaped response curve and considering the boundary conditions (Levenspiel and Smith, 1957).

The numerical concentration of the Rh6G values obtained by the PLIF system were extracted for all images under each operating condition, and the step response curves (F-curves)
at any position of interest were plotted. An example of an F-curve obtained from the PLIF experiments for the tracer concentration at the center of the images captured for any position of interest were plotted. An example of an F-curve obtained from the PLIF experiments for the tracer concentration at the center of the images captured for each operating condition. To validate the findings from the PLIF experiments, the dispersion in the contactor without side stream injection (i.e., the dispersion in a turbulent pipe flow) was compared with the published data.

Therefore, the intensity of the dispersion (also referred to as the dispersion parameter, }D_L/ud\) was determined by multiplying the dispersion number by the geometric factor (L/d). The dispersion parameter is a function of Re (Re = ud/ν, ν is the kinematic viscosity (m²/s)) and Schmidt number (Sc = ν/D_m, D_m is the molecular diffusivity of the tracer) (Levenspiel, 1999). According to the operating conditions, Re varies from about 45,000 to 100,000, and Sc varies from 3500 to 5000 as D_m of Rh6G is 2.8 \times 10^{-10} m²/s (Hansen et al., 1998; Benes et al., 2001). The average water temperature during the PLIF experiments was 9 ± 1 °C. However, Levenspiel (1958, 1999) and Ekambara and Joshi (2003) showed that the effect of Sc (i.e., the effect of molecular diffusion, D_m) can be ignored for turbulent flow conditions (Re > 2300).

This phenomenon is illustrated in Figure 11, as the values of the dispersion number obtained by several researchers agree with each other at relatively high Re values (especially for Re > 10000). The absolute relative error (ARE) between the dispersion parameter models presented in several research studies (Taylor, 1954; Levenspiel, 1958; Sittel et al., 1968; Ekambara and Joshi, 2003) and the experimental data obtained by Sittel et al. (1968) ranged from 5% to 30% for Re ranging between 10,000 to 200,000. Furthermore, Figure 11 shows a very good agreement between the measured dispersion parameter of the studied contactor and the selected published data for turbulent pipe flow. The absolute relative error (ARE) was within a range of 5 to 20%, which indicates the validity of the obtained PLIF results.

Figure 12 shows a plot of D_L as a function of the number of jets, for both 1-phase and 2-phase side jets, under a total Q_L of 0.01 m³/s. The results show that the contactor with both 1- and 2-phase side injection yielded a higher dispersion coefficient compared to that of the condition without side injection. According to Pan and Meng (2001), the decay of a scalar concentration along the centerline of a jet with a downstream axial distance (x, m) for pipe flow can be written as follows:

\[ \xi \sim e^{-0.88 \left( \frac{x}{r_D} \right)^{-0.67}} \]  

where }C_i\) is the decay of the mean concentration along the jet centerline when }x > r_D\). It should be noted that }\xi\) in Equation 4 cannot be less than the ratio between the fully mixed concentration in the pipe (C_mixed) to the initial concentration of the jet inlet (C_i).

Hence, when }\xi\) is found to be equal to }C_mixed/C_i\), ideal cross-sectional mixing is achieved. According to Equation 4, }\xi\) is about 1% at }x = 1.61 m\) (i.e., at the PLIF measurement point) compared to a }C_mixed/C_i\) of about 0.7%. This result

\[ D_L = \frac{\sigma^2}{2} = \frac{\left( \sigma/\tau \right)^2}{2} \]  

where }\sigma\) is the dimensionless standard deviation, and }τ\) is the theoretical detention time of the contactor. To validate the findings from the PLIF experiments, the dispersion in the contactor without side stream injection (i.e., the dispersion in a turbulent pipe flow) was compared with the published data.

Figure 5 shows a very good agreement between the measured dispersion parameter of the studied contactor and the selected published data for turbulent pipe flow. The absolute relative error (ARE) was within a range of 5 to 20%, which indicates the validity of the obtained PLIF results.

Figure 12 shows a plot of D_L as a function of the number of jets, for both 1-phase and 2-phase side jets, under a total Q_L of 0.01 m³/s. The results show that the contactor with both 1- and 2-phase side injection yielded a higher dispersion coefficient compared to that of the condition without side injection. According to Pan and Meng (2001), the decay of a scalar concentration along the centerline of a jet with a downstream axial distance (x, m) for pipe flow can be written as follows:

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indicates a very good radial mixing. When the first two side jets are used, then $\xi$ at $x = 0.3$ m will be 3% (about 350 $\mu$g/L). However, the tracer-free jets will expand with the flow axial direction and, hence, contribute to the higher dispersion effect. Therefore, as the number of jets increases, the liquid dispersion coefficient increases.

As the opposing jets yielded a better mixing effect at the entrance zone (Figures 7 and 8), their dispersion effect was thus lower compared to that of alternating jet for both 1-phase and 2-phase jets. The dispersion effect of the side jets increases as the number of jets increases in the opposing jets (for both 1- and 2-phase jets). However, the value of $D_L$ approaches a plateau as the number of opposing jets reaches 8. The higher $D_L$ observed in the 2-phase jets compared to that in the 1-phase jets can be related to the axial dispersion of bubbles and, hence, to the axial carrying of some tracer segments.

The dispersion coefficient, $D_L$, was found to be affected by the superficial gas velocity, ($u_G$ (m/s), the ratio between $Q_G$ to the cross-sectional area of the contactor), at the fixed superficial liquid velocity ($u_L$ (1.21 m/s), the ratio between $Q_L$ to the cross-sectional area of the contactor) for both the opposing and alternating jets. Figure 13 shows the experimental results of $D_L$ for different $u_G$ values. The following empirical correlation was used to express $D_L$ in terms of $u_G$:

$$D_L = \alpha_1 u_G^{\beta_1}$$

[5]

where $\alpha_1$ and $\beta_1$ are empirical constants that can be obtained through a non-linear regression analysis.
The constants $\alpha_1$ and $\beta_1$ were found to be 0.40 and 0.43, respectively, for the opposing jets case with a coefficient of multiple determination ($R^2$) value of 0.99. In the case of alternating jets, $\alpha_1$ and $\beta_1$ were found to be 0.42 and 0.36, respectively, with $R^2$ value of 0.92. It can be seen from the Equation 5 that $D_L$ increases with $u_G$. This result agrees with the results obtained by Weiland and Onken (1981) for a co-current gas-liquid contactor with 0.1 m diameter, $u_G$ range of 0.02 to 0.05 m/s, and $u_L$ range of 0.21 to 0.32 m/s. Weiland and Onken (1981) showed that as $u_G$ increased from 0.02 to 0.05 m/s with $u_L$ increasing from 0.21 to 0.32 m/s, $D_L$ increased from 0.01 to 0.02 m$^2$/s. Roy and Joshi (2006) modeled $D_L$ for a $u_L$ range of 0.2 to 1.6 m/s at a fixed $u_G$ value of 0.045 m/s by using a computational fluid dynamics (CFD) model. Their results showed that $D_L$ decreased from 0.4 to 0.03 m$^2$/s as $u_L$ increased from 0.2 to 1.6 m/s. The value of $D_L$ was obtained from their results at a $u_L$ of 1.21 m/s (the same $u_L$ used in this study) and a $u_G$ of 0.045 m/s to be around 0.06 m$^2$/s, which was smaller than the corresponding $D_L$ values obtained from Equation 5 (about 0.1 m$^2$/s for opposing jets and 0.14 m$^2$/s for alternating jets). The higher value of $D_L$ of the multi-jet contactor was directly related to the injection of the 2-phase jets from several locations along the contactor, which enhanced the axial dispersion of the system.
Gas Mass Transfer

The overall mass transfer coefficient ($k_{La}$, s$^{-1}$) of the contactor, for each operating condition, was determined by solving the following partial differential equation numerically:

$$\frac{\partial C}{\partial t} + u \frac{\partial C}{\partial x} = D_L \frac{\partial^2 C}{\partial x^2} + k_{La}(C_s - C) \quad [6]$$

where $C$ is the dissolved oxygen in the system (mg/L), $t$ is time (s), $u$ is the average axial velocity of the liquid (m/s), $x$ is the axial distance along the contactor (m), and $C_s$ is the equilibrium concentration of dissolved oxygen in water (mg/L). The values of $C$ were determined at the described two locations (1.37 m apart) by using the DO probes. The value of $u$ is the ratio of $Q_L$ to the cross-sectional area of the contactor.
\((A = 0.008 \text{ m}^2)\). The value \(C_s\) in water can be determined as follows:

\[
C_s = \frac{P_{O_2}}{H} \tag{7}
\]

where \(P_{O_2}\) is the partial pressure of the oxygen in the gas phase (kPa), and \(H\) is Henry’s constant (kPa.L/mg). The value of \(H\) can be described as a function of the temperature as follows (Wilcock et al., 1977; Wilhelm et al., 1977; Sander, 1999):

\[
H = H_s e^{1500 \left(1 - \frac{x}{100} - \frac{1}{50T} \right)} \tag{8}
\]

where \(H_s\) is the Henry’s constant of oxygen at 25 °C (0.41 kPa.L/mg), and \(T\) is the water temperature (°C).

Once all the constants are determined, Equation 6 can be solved numerically for each operating condition by using the appropriate numerical method. The box method was utilized as it gives a consistent and unconditionally stable solution (Weid and Klette, 2001). It can be expressed as follows:

\[
\frac{C_j^{n+1} - C_j^n}{2\Delta t} + C_j^{n+1} + \frac{u}{2} \left( \frac{C_j^n - C_j^{n+1}}{\Delta x} + \frac{C_j^{n+1} - C_j^{n-1}}{\Delta x} \right) - \frac{D_L}{2} \left( \frac{C_j^n - C_j^{n-1}}{\Delta x^2} + \frac{C_j^{n+1} - C_j^{n-1}}{\Delta x^2} \right) - k_{La} \left( C_j - C_j^{n+1} \right) = 0 \tag{9}
\]

where \(C_j^n\) is the concentration of dissolved oxygen (mg/L) at axial grid node \(j\) and temporal time grid node \(n\) (\(j = 1, 2, \ldots, M\) and \(n = 0, 1, 2, \ldots, N-1\)), \(\Delta t\) is the time step (s), and \(\Delta x\) is the axial distance step (m). Equation 9 was solved by combining the results from the PLIF measurements for \(D_L\), the measured DO concentration at the two locations along the contactor, and by providing an initial guess for the value of \(k_{La}\). The corresponding overall mass transfer coefficient of oxygen \((k_{La-O_2}\), s\(^{-1}\)) was then calculated by using iteration until the minimum sum of the squared error was achieved. The overall mass transfer coefficient of the ozone \((k_{La-O_3}\), s\(^{-1}\)) was determined by using the obtained \(k_{La-O_2}\) by applying the following relationship which was introduced by Danckwerts (1970) and validated by Sherwood et al. (1975):

\[
\frac{k_{La-O_3}}{k_{La-O_2}} = \sqrt{\frac{D_{O_3}}{D_{O_2}}} \tag{10}
\]

where, \(D_{O_3}\) and \(D_{O_2}\) are the molecular diffusivities of ozone and oxygen gases, respectively, in water \((1.74 \times 10^{-9} \text{ and } 2.50 \times 10^{-5} \text{ m}^2/\text{s},\) respectively). The \(k_{La-O_3}\) values were obtained at 20 °C by applying the following relationship (Roustan et al., 1996):

\[
(k_{La-O_3})_T = (k_{La-O_3})_20 \times 1.024^{20-T} \tag{11}
\]

where \(T\) is the water temperature in the ozone contactor (°C). The value of \(T\) ranged from 20 to 23 °C during all DO measurements.

The overall mass transfer coefficient, \(k_{La}\), was found to be favorably affected by \(u_G\) at the fixed \(u_L\) value of 1.21 m/s, for both the opposing and alternating jets. As with the case of \(D_L\), the following empirical correlation, proposed by Deckwer et al. (1974), was used to show the effect of \(u_G\) on \(k_{La}\):

\[
k_{La} = \alpha_2 u_G^{\beta_2} \tag{12}
\]

Figure 14 shows the experimental results of the ozone-based \(k_{La}\) for different \(u_G\) values. The constants \(\alpha_2\) and \(\beta_2\) were found to be 1.27 and 0.48, respectively, for the opposing jets case with a \(R^2\) value of 0.98 compared to 2.27 and 0.69, respectively, for the alternating jets with a \(R^2\) value of 0.96. As the gas flow rate is increased (by increasing the number of side jets), the value of \(k_{La}\) increases.

This result can be attributed to the expected increase in the gas hold-up. This finding agrees with the findings of Briens et al. (1992) (who used a venturi ozone contactor with \(Q_G/Q_L < 1\) and \(u_L = 0.95 \text{ m/s}, k_{La} = 0.97u_G^{0.85}\)) and Mao et al. (1993) (who used a plunging jet ozone contactor, \(Q_G/Q_L < 1\) and \(u_G\) ranges from 0.025 to 0.25 m/s, \(k_{la} = 1.17u_G^{0.82}\)). It should be noted that the multi-jets ozone contactor outperformed the contactors used by these researchers. Moreover, the use of opposing jets yielded higher \(k_{La}\) values at the same \(u_G\) (about 15%). This result can be related to their better radial mixing and, hence, to the lower \(D_L\) values, as discussed earlier. Also, the bubble size obtained by using the opposing jets alignment is expected to be less than that obtained from the alternating jets alignment (at the same volume of gas) due to a possible bubble shear-off resulting from the impingement of the entering jets on each other. This result can lead to a higher interfacial area \((a, \text{ m}^{-2})\) and, hence, increases the total value of the \(k_{La}\).

The effect of \(u_L\) and \(u_G\) on \(k_{La}\) was also investigated for opposing jets (as they exert a higher mass transfer than that of the alternating jets) by varying the values of \(Q_L\) and \(Q_G\).

**FIGURE 14.** Overall mass transfer coefficient for alternating and opposing jets.
and fixing the number of the utilized side jets. The following expression for the ozone-based \( k_{La} \) was obtained for two opposing jets:

\[
k_{La} = 0.8 u_G^{0.36} u_L^{0.23}
\]

Figure 15 shows the obtained regression model (Equation 13) versus the experimental calculated \( k_{La} \) for two opposing jets. An excellent agreement is clearly observed with a \( R^2 \) value of 0.99. It was found that the mass transfer rate of the ozone increased with both \( u_G \) and \( u_L \). However, the effect of \( u_G \) was more pronounced. This result was expected as the gas flow rate was the driving force for such a process with the given range of the operating conditions. This finding agrees with those in several other studies of ozone contactors (Wang and Fan, 1978; Huynh et al., 1991; Zhou and Smith, 2000; Gamal El-Din and Smith, 2003).

The comparison of Equations 12 and 13 at \( u_L \) of 1.21 m/s for a \( u_G \) representing 4, 6 and 8 opposing jets (0.025, 0.037, and 0.049 m/s, respectively) shows that the multi-jets induce a higher mass transfer rate compared to that of the two opposing jets, only when \( u_G \) was higher than 0.025 m/s (i.e., when more than 4 opposing jets are used). This result can be related to the decrease of the available space in the mixing zone needed for higher gas mass transfer to occur in the case of two opposing jets with \( u_G \) greater than 0.025 m/s. Furthermore, as the value of \( D_L \) is higher for 8 opposing jets compared to that for 6 opposing jets, the use of the 6 opposing jets (3 from each side) may yield an enhanced \( k_{La} \) at high \( u_G \) values.

**CONCLUSIONS**

This study investigated the design of a pilot-scale of an in-line, multi-jets ozone contacting system. A two-dimensional laser flow map particle image velocimetry coupled with planar laser-induced fluorescence (PIV/PLIF) was used to characterize the hydrodynamics of the contactor under different operating conditions. By using the PIV system, velocity measurements of the two phases (liquid and gas), were taken at different locations along the contactor to examine the flow patterns within the system.

The results of velocity vectors, for both phases, at the mixing zone (the entrance zone of the side jets) showed that the opposing jets exerted better radial mixing compared to that of the alternating jets. This finding was supported by the findings obtained from the PLIF measurements as the axial dispersion coefficient resulting from the use of opposing jets was found to be smaller than that of the alternating jets. Furthermore, it was found that as the number of the 2-phase jets increased (i.e., as the gas-flow rate increased), the dispersion coefficient increased due to the dispersion effect of the bubbles.

The results obtained by using the PIV and PLIF systems were coupled with dissolved oxygen measurements, and monitored at two different locations along the contactor (upstream and downstream of the 2-phase side jets) to estimate the overall mass transfer coefficient (\( k_{La} \)) of the system under wide operational conditions. The value of \( k_{La-O_3} \) was found to increase with the number of jets (i.e., with higher gas-flow rates). Also, higher \( k_{La-O_3} \) values were observed when opposing jets were used compared to those observed when the same number of alternating jets were used. It was found that when \( u_G \) was higher than 0.025 m/s, the use of 6 and 8 opposing jets exerted higher \( k_{La} \) values than those exerted by the use of two opposing jets with the same \( u_G \) value. Hence, the use of 6 to 8 opposing jets is recommended at high gas flow rates for the studied ozone contactor at a liquid flow rate of 0.01 m\(^3\)/s.

**REFERENCES**


