Effect of distributor type on interphase mass transfer and gas axial dispersion in bubbling and turbulent fluidized-bed reactors

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Keywords: ozone decomposition-reaction; interphase mass transfer; transitional reactor model

Abstract—Three distributors consisting out of a perforated plate (baseline), a perforated plate with a porous cloth above (porous distributor) and a novel Multi-Vortex (MV) tuyere distributor were tested in a fluidised flatbed reactor (400mm x 25mm). Fluid Catalytic Cracking catalyst impregnated with ferric-oxide was used to decompose ozone. The porous-distributor resulted in large bubbles and poor reactor performance compared to the baseline. The MV-distributor resulted in a mean improvement in conversion efficiencies of 14.7% relative to the baseline. Bubble sizes were measured for the baseline and MV distributors using a visual as well as pressure fluctuation technique indicating slightly bigger bubble sizes for the MV distributor especially at higher superficial velocities. The bubbling-to-turbulent transition velocity ($U_c$) was determined to be 0.25m/s, 0.3m/s and 0.4m/s for the MV-, baseline- and porous- distributor, respectively. Conversion with the MV-distributor increased to near plug flow values at velocities exceeding $U_c$, which indicates a significant decrease in the axial dispersion in the bed.

INTRODUCTION
Low velocity catalytic gas-fluidisation (bubbling and turbulent regime) typically has lower overall reaction rates than predicted by the intrinsic kinetics of the catalyst. This phenomenon is primarily caused by interphase mass transfer limitations as a result of particle-fluid separation, which causes bubbles to form in the bed and induce bypassing of gas to occur. Increasing the gas flow rate typically increases the bubble size in the bed which reduces the mass transfer rate between the bubbles and the solids; this can affect the overall reaction rate.1-6 In addition to the interphase mass transfer, gas-backmixing in the emulsion phase can alter the reaction performance. Historically the main focus in rectifying this problem was to reduce the bubble sizes in the fluidized bed reactor (FBR).1-3,4,6-10

There are numerous methods to quantify the contribution of the mixing schemes in a FBR. One approach is to quantify changes in specific hydrodynamic parameters like bubble size, bubble velocity, gas hold-up etc. The measured changes (improvements) can then be used to speculate its contribution to the overall performance of a FBR. Alternatively, one can start from an overall performance measurement by employing a chemical reaction. For atmospheric cold-flow columns the ozone decomposition reaction has proven itself as an ideally suited reaction in this regard.11 The drawback of the overall performance measurement is that the underlying reason for the change (improvement) might not be clear causing scale-up of the novel mixing scheme to be uncertain without fundamental understanding. Reactor models in conjunction with literature correlations can be employed to investigate possible reasons, but the uniqueness of the specific system is not considered by the models and correlations might lead to an incorrect conclusion. It is therefore preferable to join the abovementioned methods, so that the analysis of the overall performance result is based on additional hydrodynamic measurements performed in the same bed.
The purpose of this study was to investigate distributor effects on the mixing schemes in a FBR using the technique of overall performance in conjunction with specifically measured hydrodynamic parameters. A porous distributor, perforated plate and a novel multi vortex (MV) distributor was used. The MV distributor is based on the concept of increased radial and tangential momentum by introducing the fluidising medium parallel to the distributor at different heights as well as perpendicular to the distributor. The porous- and MV-distributor were tested against the perforated plate (generally used in industry) with a comparable open area ratio and orifice velocities. The performance of the distributors was determined by fitting the Thompson et al. model to the experimentally determined conversions and analysing the fitted parameter values for physical significance. Additionally the bubble sizes were determined for the MV distributor and perforated plate distributor using a photographic method as well as a mathematical technique in which the pressure fluctuations in the bed were decomposed and analysed. The bubble properties were incorporated into the fitted model. The experimental procedure was performed in a two-dimensional (2-D) FBR with FCC catalyst, impregnated with Fe₂O₃ for the ozone decomposition reaction. The experiments were done at ambient conditions for superficial velocities ranging from 0.1 m/s to 0.55 m/s, which included the bubble-turbulent transition boundary for the systems.

**BACKGROUND**

Van Ommen et al. and Kleijn van Willigen et al. states that the disadvantages of fluidised bed reactors (FBRs) can be overcome by manipulating the hydrodynamics of FBRs, thereby decoupling some of the conflicting design objectives in the FBRs. Examples of these conflicting objectives are bubble size vs. high gas flow and conversion vs. backmixing in the FBR in the bed. This manipulation can take place by either altering the manner in which the fluidising medium is introduced or by changing the physical properties or state of the solid phase in the FBR. In both these cases a change in the dynamics or a geometric change can be applied to the specific FBR property, i.e. a change in the dynamics of the fluidising medium would constitute dynamic changes to the flow into the FBR, while a geometric change would involve redistribution of gas by internals in the FBR or a change to the distribution system of the FBR. Dynamic changes to the solid phase would entail a modification in the rate of change in the particle distribution in the bed while geometric changes would involve a change in the particle properties, e.g. the particle size distribution in the bed. Wong & Baird and Pence & Beasley altered the dynamics of the gas supply by applying continuous periodic variations and found a significant improvement in the quality of fluidisation, a measure of the phase distribution in the bed. Kleijn van Willigen et al. successfully reduced the bubble sizes in a FBR by manipulating the dynamics of the particles by applying an alternating electric field to the FBR, which polarised the particles and thereby increased the interparticle forces in the bed. Sun & Grace and Beetstra et al. studied the influence of particle size distribution, i.e. the geometry of the particles, on FBR performance and found that an increased fines content and a greater width of the particle size distribution resulted in the greatest improvement in performance and reduction in bubble sizes in the FBR. Geometric manipulation of the gas supply and flow patterns in the FBR has been studied extensively in literature. This method includes the installation of internals in the bed and the design of the distribution system for the FBR. Internals have been proven successful at bubble breakup, decreasing backmixing in the FBR by dividing a FBR into compartments, or by shedding the wake from the bubbles as they pass the internals (specifically wire mesh internals). The design of the distribution system to the FBR is considered crucial to the quality of fluidisation of the FBR and conversion of reactant in the distributor section of the bed; it has been proven that as much as 50% conversion can take place in the shallow section less than 10 cm above the distributor. This means that the distributor can act as a useful gasflow manipulator and has been used successfully in various studies to manipulate the hydrodynamics in the FBR.

Sobrino et al. tested the performance of bubble cap distributors as compared to a perforated plate distributor and found that the perforated plate distributor induced earlier onset of turbulent behaviour in the FBR due to jetting from the plate; this is preferred as the greatest gas mixing was observed when the turbulent fluidization regime is entered. Kleijn van Willigen et al. tested the performance of a secondary fractal injection distributor and found a decrease in the bubble size with an accompanying decrease in total number of bubbles and
therefore contact surface between the bubbles and particles.\textsuperscript{15} Yan \textit{et al.} compared a multi-orifice plate distributor to a multi-tube distributor and found that the axial hold-up and radial distribution of solids was more uniform for the multi-tube distributor when compared to the multi-orifice distributor.\textsuperscript{30} Ouyang \textit{et al.}, Chyang \textit{et al.}, Sreenivasan \& Raghavan and Wormsdecker \textit{et al.} compared the influence of horizontal injection of gas into the FBR, to induce a swirling bed FBR, to the normal axially injected distributor.\textsuperscript{7,8,9,10,31,32} They found that the deficiency of the bubbling regime can be addressed by increasing the horizontal momentum in the bed. The gas from conventional FBR distributors (porous plate, perforated plate and multi-vertical nozzle) possesses only axial momentum which causes the gas at any point in the FBR to possess an axial component (vertical component) which significantly exceeds the radial or tangential components (horizontal components). The axial component of the gas velocity is responsible for the fluidisation of the particles in the bed, while the horizontal component is responsible for horizontal momentum being transferred to the bed. A deficiency in horizontal velocity components reduces the fluid movement and therefore the mixing in the FBR. Injecting gas into a FBR at an angle of less than 90° to the horizontal, the horizontal component of the injected gas are increased which in turn promotes solids distribution and therefore mass transfer in the FBR. The reduction of the horizontal momentum higher up the bed can be counteracted by further decreasing the injection angle of gas to the FBR. It can therefore be deduced that the theoretical maximum horizontal momentum would be transferred to the FBR at an injection angle parallel to the distributor. Chyang \textit{et al.} further found that using a tuyére and horizontal nozzle distributor the lateral dispersion and lateral mixing in a FBR are significantly improved with an accompanying reduction of the dead zones in the FBR.\textsuperscript{9,32} Applying a radial force to the particles in the FBR by injecting gas at multiple points along the outer wall of the FBR was tested by De Wilde and De Broqueville.\textsuperscript{33} This gas distribution design induced rotation of the particles and therefore increased the tangential and centrifugal forces on the particles.

**EXPERIMENTAL**

**Reaction measurements**

The reactor used for the experimental study was a Plexiglas 2-D fluidised bed reactor with thickness 25 mm, width 400 mm and height 4.5 m. A volute-inlet primary cyclone, to handle the high solids loading at the upper gas velocities, and a tangential-inlet secondary cyclone were used. Excess fines that bypassed the secondary cyclone were captured in filter bags connected after the secondary cyclone. Figure 1 shows the piping and instrumentation for the experimental setup. The volumetric flow of reactor feed gas was controlled with a vortex flow meter with a linear velocity range of between 0.1 m/s and 0.6 m/s. Two high frequency pressure transmitters (Wika S-10, Range 0-1.6 barg, and maximum measurement frequency of 1000Hz) were inserted on the upper surface of the distributor and 0.3 m above the distributor. Constant humidity compressed air at 15°C was used as gas supply. The air was dosed with ozone (generated in a EcoTec MZV1000) in order to have an inlet ozone concentration varying between 20 ppm and 100 ppm. Pure oxygen was used as feed to the ozone generator to reduce NO\textsubscript{x} formation. The inlet concentration was determined by taking a sample from the plenum chamber. The sampling probe was inserted in the centre of the plenum chamber and glass beads were added to ensure good gas mixing and distribution. The sampling flowrate was negligible compared to the total flowrate. The outlet sample was drawn 4.1 m above the distributor, also from the centre of the column. The samples were continuously analysed using the 2B Technologies Inc. UV-106 ozone analyser. Ozone analysis is done by the well-established method of light absorption at a wavelength of 254 nm.

At a set superficial linear velocity the inlet ozone concentration was logged for 10 min to account for variations in the ozone concentration after which the analyser feed was changed to the outlet ozone concentration which was logged for 10 min. Data acquisition was done using National Instruments USB-608 analogue signal data loggers connected to a PC. The readings from the ozone analyser were measured at a rate of 5Hz, the velocity measurements were logged at a rate of 20Hz and the pressure fluctuations at a rate of 200Hz.

Initially 3.75 kg of catalyst was loaded to the reactor. It was estimated that 0.75 kg of catalyst remained in the return system and did not take part in the chemical reaction since no
ozone was present in the fluidising gas flow to the dipleg. This mass fraction remained more or less constant at all superficial velocities employed. The bed height prior to fluidisation was in the order of 400 mm.

The distributors used in the experimental setup were—

- A triangular pitch perforated plate distributor with $35 \times 2$ mm holes, with a porous cloth between the plenum chamber and the distributor, to prevent weepage. The cloth also increased the pressure drop over the distributor and therefore the gas distribution through the distributor. This distributor setup will be called the baseline distributor.

- The same triangular pitch perforated plate with the porous cloth above the distributor plate. This was done with the intention to try and break the characteristic jets formed with a perforated plate distributor and create a porous type distributor.

- A multi-vortex (MV) distributor consisting of $38 \times 1/16''$ outside diameter (OD) tuyères, with rectangular pitch, ejecting gas either horizontally or vertically in strategic directions. Glass wool was inserted into the nozzles to prevent weepage and increase the pressure drop for improved distribution. Figure 2 shows a photograph as well as the proposed gas flow pattern of the MV distributor used in the experimental study.

![Figure 1](image-url)

**Figure 1.** The piping and instrumentation for the experimental setup adapted from Saayman34
Connected in parallel to the FBR is a small test reactor (see Figure 1). It consists of a packed bed 50 mm in height and with a 16.4 mm inner diameter (10g of catalyst). This reactor was used to monitor the stability of the catalyst activity and to determine reaction rate constants. Axial dispersion in this reactor was estimated to be negligible [16] and accordingly a plug flow model was used to determine the kinetics.

The catalyst in this study was produced by adding commercial FCC catalyst (support particle) to a mixture of 10%(wt) Ferric Nitrate solution. After stirring for one hour the mixture was calcinated at 450 °C for approximately 1.4 hours, during which the NO₂ gasses were released from the mixture. According to literature, the ozone decomposition reaction is first order with respect to the ozone concentration [19,35-40].

![Figure 2. Photograph and front-view representation of the gas flow pattern of the MV distributor during operation](image)

**Bubble size measurements**

For the visible bubble size measurements a 12.1 megapixel digital camera (Sony Cybershot DSCW230 12.1MP) was used, with a frame rate of 30 fps. A standard 300 mm ruler, marked at 10mm intervals, was fixed to the column as a reference. Data analyses were done visually and logged in Microsoft® Excel 2007. The column was filmed for a period of 30s, with the camera mounted at the ruler height. The linear velocities in both distributor configurations ranged from 0.1 m/s to 0.35 m/s in intervals of 0.05 m/s.

A pressure fluctuation decomposition technique described by Van der Schaaf et al. was employed to determine the relative bubble sizes in the column at a height of 300 mm above the distributor plate [14]. The technique compares the coherent and incoherent pressure fluctuations in the bed, from which the comparative bubble sizes can be determined.

**RESULTS AND DISCUSSION**

**Catalyst and reaction**

The first order behaviour of the reaction was confirmed in the small packed bed experiment and results are given in Figure 3. The deactivation profile for a catalyst batch exposed to ambient conditions is given in Figure 4. The initial drop in activity is most likely related to desorption of water from the catalyst, since the activity plateau (reached after two hours) could be obtained without the use of ozone. The activity plateau exhibited a fluctuating behaviour ($k_r''''$ varies between 1.3 and 1.6 in Figure 4) and accordingly intermittent kinetic determinations of fluidised bed samples were done in the packed bed throughout the time span of a fluidised bed experimental run (where the catalyst inventory was not exposed to ambient conditions), this was done to ensure $k_r''''$ remained within an acceptable range. The samples tested were extracted from the bed at intervals of between three and five hours. From Figure 4
it is evident that the plateau fluctuations are of a random nature and accordingly a best fit single rate constant was used for each experimental run.

The catalyst particle size distributions were determined using a Malvern Mastersizer 2000 before and after each experimental run. For the porous-distributor experiments the Sauter mean diameter was constant, but 22% lower with a wider distribution and more fines than for the baseline case because a different batch of catalyst was used. For the MV experiment the same batch of catalyst was used as the baseline. The measured particle size distributions for the baseline and MV distributors were relatively constant; the Sauter mean diameter varied by 8.3% and 4.6% and the standard deviation varied by 4.8% and 5.9% for the baseline distributor and MV distributor respectively.

![Graph showing typical curves at different times during an experimental run.](image)

**Figure 3.** Typical curves at different times during an experimental run, measured in the packed bed reactor in parallel to the FBR. The time \( t = 0 \text{h} \) represent the time after deactivation \( i.e., 2\text{h} \) after start-up.

**Figure 4.** Typical curve for the deactivation of the catalyst over time.

**Fluidised bed results**

The transition velocity was determined to be 0.4m/s for the porous distributor, 0.3m/s for the baseline distributor and 0.25m/s for the MV distributor. All reaction measurements are reported in terms of a conversion efficiency, defined as the ratio of the measured conversion relative to the maximum (or plug flow) conversion at the given superficial velocity. It enables
comparison between runs of different activities and is useful to visualise the reactor performance. The experimental results for the porous-, MV- and baseline-distributor are shown in Figure 5. A total of three, six and four experimental runs were completed with the porous-, MV- and baseline-distributor, respectively. The error bars show one standard deviation from the average measured conversion, taken over the experimentally determined ozone measurements. As can be seen in Figure 5, there seems to be quite extensive scatter in the measured conversion efficiency, yet a prominent trend is exhibited in all cases. A very definite dip in the conversion efficiency takes place from 0.1 m/s to approximately 0.3 m/s. It is clear from Figure 5a that the porous distributor performs remarkably poor compared to the baseline despite the higher fines content. It is postulated that possible cloth stretching in the centre caused maldistribution due to increased apertures caused by stretching. The porous distributor runs were also visually characterised by large voids bypassing the bed in the centre of the reactor. Accordingly the remainder of the interpretation will focus on the other distributors.

![Figure 5](image)

*Figure 5. The average conversion efficiency for the measured conversions of the a) porous-, b) MV- and c) baseline-distributor. Error bars show one standard deviation from the average.*

To refine the results, i.e. filter the scatter to a more manageable trend, a moving average over 5 consecutive data points was applied to the data. The filtered data can be seen in Figure 6.

The data shows a definite increase in the conversion efficiency when using the MV distributor as compared to the baseline distributor. Figure 6b shows the percentage increase of the conversion efficiency for the MV-distributor. This increase varies between 0% and 20%, with a mean improvement of 14.7% and an increasing trend with increasing velocity. The data indicates that the improvement is related to interphase mass transfer as well as gas phase axial dispersion differences. The interphase mass transfer improvements are evident in the lower
velocity range where the conversion efficiency is lower than that of the ideal mixing (CSTR) scenario. The axial dispersion differences are evident above a velocity of 0.3 m/s where the MV distributor approaches plug flow performance, while the performance of the baseline distributor remains between the mixing and plug flow extremes (interphase mass transfer is unlikely to play a role at these high velocities since the bed is operating in the turbulent regime).

![Graph](image)

**Figure 6.** (a) The moving average of the measured conversion efficiencies for the MV distributor and the baseline distributors, taken over 5 consecutive data points. (b) The percentage improvement from the baseline to MV distributor

**Bubble size analysis**

Given the clear differences in interfacial mass transfer characteristics, bubble size quantification can assist in understanding these differences. Two separate measurement techniques were employed as discussed in the experimental section and the respective results are given in Figure 7 and Figure 8. Both sets of results indicate an increase in bubble size as the velocity increases. More interestingly, both sets indicate larger bubbles for the MV distributor which is an unexpected result given the conversion efficiency measurements. Since the standard deviation of the incoherent pressure fluctuations $\sigma_{xy,i}$ needs to be calibrated with measured bubbles sizes, the bubble growth slope with respect to gas velocity was used to compare the separate techniques. The MV distributor slope was found to be 1.4 times that of the baseline distributor case for the visual measurements while a factor of 1.5 was calculated from the pressure fluctuation measurements. It is therefore evident that the interface mass transfer difference cannot be attributed to bubble size differences, but rather to the mass transfer mechanism or total bubble fraction in the bed. Total bed height measurements at the same superficial velocity indicated no significant difference between the two distributors and accordingly the total bubble fraction argument is unlikely to explain the observed difference in conversion. With regards to the mass transfer mechanism one might speculate that the convective component of the exchange is more severe for the MV bubbles, but no quantification exists to substantiate this argument.
Figure 7. (a) and (b) The visually measured bubble diameters with the linear correlation of the bubble diameters for the baseline and MV distributor respectively. (c) The comparison of the best fit linear correlations from (a) and (b).

Model analysis

To extend the analysis the combined bubbling turbulent model of Thompson et al. was implemented to the system. Table 1 shows the comparative variables fitted for the different studies, with the fitted parameters shown in bold. The column on the far right (light grey) show values fitted by Thomson et al. for experimental results determined by Sun in a three
dimensional (3-D) FBR. The measured values for bubble size and transition velocity ($U_t$) were directly used in the model. The only fitted parameters used are the correction factors for the interphase mass transfer coefficient ($f_{k_{r}}$) and axial dispersion ($f_{D}$). With freedom on the value of these two parameters decent fits were obtained for both distributors (within 10% error) as can be seen in Figure 9, using an error minimisation technique in which the absolute average relative error percentages (AARE%) were minimised. The results of the minimisation are shown in Figure 10. The fitted values for $f_{k_{r}}$ show that an increase in mass transfer is apparent, the $f_{k_{r}}$ for the MV distributor is approximately 50% greater than for the baseline distributor.

Given the larger bubbles of the MV distributor this correction is required and can be attributed to the proposed improvement in the mass transfer mechanism. The value of $f_{k_{r}}$ for the 3-D bed in the Thompson study is significantly higher than the corresponding 2-D scenarios. From geometric considerations (2-D vs. 3-D) it can be deduced that the factor should be 1.4 times smaller for the 2D scenario but this is not sufficient to explain the total difference. Another interesting observation is the difference in the axial dispersion correction ($f_{D}$). The prediction for axial dispersion in the turbulent regime used in the Thompson study is the Bi et al. correlation. For the two dimensional bed in this study a vessel Peclet number of approximately 3 is predicted by the correlation, suggesting that the behaviour of the turbulent bed will be somewhere between ideal mixing and plug flow. The correction factor for the baseline 2-D distributor and the Thompson study indicates that the fitted dispersion is roughly in agreement with the correlation, while this is not nearly the case for the MV distributor where near plug flow behaviour ($P_e \approx R_e$) is observed at a velocity of 0.35 m/s. Similar to the interphase mass transfer difference the reason for the major improvement in vessel Peclet number can only be speculated upon.

Lastly one should quantify the interplay between interfacial mass transfer and axial dispersion. This was done by varying the two fitted parameters while monitoring the fitting error. From Figure 10 it is evident that variation in $f_{k_{r}}$ as well as $f_{D}$ are limited to a range of values. The $f_{k_{r}}$ could increase by as much as 100% while the $f_{D}$ changed by less than 10% when maintaining an AARE% of less than 5%. is indicates an independence of the mass transfer and axial dispersion in the model, which can be explained by the observation that the $f_{k_{r}}$ affects the bubbling regime during which time the two phase bubbling model dictates the prediction. The $f_{D}$ only affects the tail end of the fit (velocities higher that $U_t$) when the model transcends from the two phase model to the axially dispersed plug flow model and the designated asymptote between the CSTR and PFR extremes is set by the value of $f_{D}$.

### Table 1. The variables used for the simulation of the Thompson model

<table>
<thead>
<tr>
<th>Variable</th>
<th>Current Study</th>
<th>Thompson et al.(^{12})</th>
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</thead>
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<tr>
<td>Distributor</td>
<td>Perforated Plate</td>
<td>MV Distributor</td>
</tr>
<tr>
<td>Column Type</td>
<td>2-D</td>
<td>2-D</td>
</tr>
<tr>
<td>$\overline{k_{r}}$</td>
<td>1.6 s(^{-1})</td>
<td>1.33 s(^{-1})</td>
</tr>
<tr>
<td>$M_{sg}$</td>
<td>3.6</td>
<td>3.6</td>
</tr>
<tr>
<td>Catalyst</td>
<td>FCC</td>
<td>FCC</td>
</tr>
<tr>
<td>$d_{c}$</td>
<td>87 (\mu m)</td>
<td>84 (\mu m)</td>
</tr>
<tr>
<td>Bubble size</td>
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<td>2.2 cm – 7.7 cm</td>
</tr>
<tr>
<td>$f_{c}$</td>
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<td>$f_{k_{r}}$</td>
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<td>$f_{D}$</td>
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Table 1—continued

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<th>Current Study</th>
<th>Thompson et al.</th>
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<tbody>
<tr>
<td>$\mu_g$</td>
<td>2.0 x 10^{-5}$\text{Pa.s}$</td>
<td>2.0 x 10^{-5}$\text{Pa.s}$</td>
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<td>$\rho_s$</td>
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<td>1580 $\text{kg/m}^3$</td>
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<tr>
<td>$\rho_g$</td>
<td>1.20 $\text{kg/m}^3$</td>
<td>1.20 $\text{kg/m}^3$</td>
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</table>

Figure 9. (a) The Thompson model predictions for the measured conversion efficiencies for both the baseline and the MV distributor. Also shown is the CSTR model for the respective distributor cases. (b) and (c) are the % error plots for the baseline and MV distributor cases.
CONCLUSIONS

The 2-D FBR performance using a multi-vortex (MV) distributor and a perforated plate with porous cloth above (porous type distributor) was compared to a standard perforated plate distributor (baseline) with comparable open area and superficial velocity. In this study the ozone decomposition reaction in the bed was measured to determine the overall performance of the distributors and it was found that the MV distributor causes a significant improvement in the conversion efficiencies when compared to the baseline at all velocities tested. The improvement varied between 0% and 20%, with a mean improvement of 14.7% and an increasing trend with increasing velocity. The performance of the porous distributor was extremely poor due to the significant maldistribution.

Additionally, the bubble sizes were measured using a visual measurement technique as well as a pressure fluctuation analysis technique. The bubbles were found to grow at a rate approximately 1.4 times greater for the MV distributor than the baseline case. The bubbling to turbulent transition boundary ($U_b$) was determined to be 0.25 m/s for the MV distributor whereas the baseline distributor was at 0.3 m/s. The porous type distributor was at 0.4 m/s.

The improved overall performance for the MV distributor as compared to the baseline, despite the increase in bubble sizes, could indicate that the mass transfer mechanism (as opposed to the actual bubble sizes) influences the performance in this case, i.e., a possible convective component to the mass transfer for the MV bed, which exceeds that of the baseline could be present but no quantification is available to support the validity of this argument. The conversion for the MV distributor increases to near plug flow at velocities exceeding $U_b$, which
indicates a significant decrease in the axial dispersion in the bed; the predicted \( \mathcal{P}_f \approx \frac{1}{n} \) for the MV distributor at these velocities as compared to the baseline \( \mathcal{P}_f \approx 2 \) supports this.

### NOMENCLATURE

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
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<tbody>
<tr>
<td>( A_{open} )</td>
<td>Total open area on distributor [mm²]</td>
</tr>
<tr>
<td>( A_{or} )</td>
<td>Injection orifice cross-sectional area [mm²]</td>
</tr>
<tr>
<td>( C_{ozone,IN} )</td>
<td>Ozone inlet concentration [mg/L]</td>
</tr>
<tr>
<td>( C_{ozone.OUT} )</td>
<td>Ozone outlet concentration [mg/L]</td>
</tr>
<tr>
<td>( d_p )</td>
<td>Average particle size [m]</td>
</tr>
<tr>
<td>( f_c )</td>
<td>Fraction of total solids in cyclones and dipleg during normal operation [(-)]</td>
</tr>
<tr>
<td>( f_k )</td>
<td>Interphase mass transfer fitting parameter [(-)]</td>
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<tr>
<td>( f_{Pe} )</td>
<td>Peclet number fitting parameter [(-)]</td>
</tr>
<tr>
<td>( ID )</td>
<td>Inner diameter of injection orifice [mm]</td>
</tr>
<tr>
<td>( k_{r''} )</td>
<td>Reaction rate constant [s⁻¹]</td>
</tr>
<tr>
<td>( M_{SO} )</td>
<td>Total solids inventory in bed [kg]</td>
</tr>
<tr>
<td>( N_{or} )</td>
<td>Number of injection orifices [(-)]</td>
</tr>
<tr>
<td>( P_e )</td>
<td>Vessels Peclet number [(-)]</td>
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<tr>
<td>( Q )</td>
<td>Volumetric flowrate into the reactor [m³/s]</td>
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<tr>
<td>( U_0 )</td>
<td>Linear superficial inlet velocity [m/s]</td>
</tr>
<tr>
<td>( U_c )</td>
<td>Bubbling to turbulent regime transition velocity [m/s]</td>
</tr>
<tr>
<td>( U_{or} )</td>
<td>Orifice superficial velocity [m/s]</td>
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<tr>
<td>( V )</td>
<td>Volume of the reactor [m³]</td>
</tr>
<tr>
<td>( x )</td>
<td>Conversion of ozone to oxygen [(-)]</td>
</tr>
<tr>
<td>( \mu_g )</td>
<td>Gas viscosity [Pa.s]</td>
</tr>
<tr>
<td>( \rho_g )</td>
<td>Gas density [kg/m³]</td>
</tr>
<tr>
<td>( \rho_s )</td>
<td>Solid particle density [kg/m³]</td>
</tr>
<tr>
<td>( \sigma_{ozone,IN} )</td>
<td>Standard deviation in ( C_{ozone,IN} ) [mg/L]</td>
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<tr>
<td>( \sigma_{ozone.OUT} )</td>
<td>Standard deviation in ( C_{ozone.OUT} ) [mg/L]</td>
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<tr>
<td>( \sigma_x^2 )</td>
<td>Variance in the conversion of ozone to oxygen [(-)]</td>
</tr>
<tr>
<td>( \sigma_{xy,i} )</td>
<td>The incoherent standard deviation of pressure fluctuations [kPa]</td>
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<tr>
<td>( \Phi_{LO} )</td>
<td>Solid volume fraction in the bubble when operating in the bubbling regime [(-)]</td>
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### REFERENCES


