Hydrodynamic characteristics of cocurrent upflow and downflow of gas and liquid in a fixed bed reactor


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Abstract

The residence time distribution (RTD) of a bench-scale multiphase reactor has been investigated in the present study using air as a gaseous phase and water as a liquid phase. Two different sizes (1.1 and 0.3 mm) of silicon carbide were used as diluent for the catalyst bed. In the entire study, equal volumes of catalyst and diluent were used in the bed. The ranges of air and water velocities were kept at such levels as to simulate the liquid hourly space velocities and hydrogen/oil ratios of typical bench-scale hydroprocessing units. The experiments were conducted in both upflow and downflow modes of operation in the reactor. The two modes of operation showed wide differences in RTD when a larger size of diluent was used. However, the differences in RTD in both the modes of operation were eliminated by using a smaller size of diluent. The effect of gas and liquid velocities on different hydrodynamic parameters such as bed Peclet number of liquid, mean residence time of liquid and operating liquid holdup were also investigated for these two levels of diluent size.

Keywords: RTD studies; Diluent; Upflow reactor; Trickle bed reactor

1. Introduction

Trickle bed reactors with cocurrent downflow of gas and liquid have found wide application in oil industries for the hydroprocessing of petroleum fractions. The successful design of commercial reactors involves generation of reliable data in laboratory-scale reactors and scaling up of these data for larger units. Similarly, scaling down of a commercial unit to a smaller size reactor is also required to study newer catalysts as well as alternative feedstocks for an established process. The laboratory or bench-scale units with lower capital investments are preferred for these investigations. Studies in these units also involve lower operating cost, lesser inventory of chemicals, safer operation and a reduced risk of hazards. Owing to these advantages, the small-scale reactors are gaining importance daily for the scaling-up and scaling-down of commercial reactors [1].

These reactors have their own inherent drawbacks. The length and diameter of a laboratory unit are much smaller than those of a commercial reactor. Thus, if catalysts, in their commercially applied shape and size, are to be tested in these reactors, the ratio of the diameter of the reactor to that of catalyst particles becomes very low. Because of this low ratio, the wall effect in these reactors becomes significant. The liquid superficial velocity in these reactors is also very low because of the small amount of catalyst used in these reactors as compared to that in commercial units for the same values of liquid hourly space velocity (LHSV) in these units [2]. The lower liquid velocity, along with the wall effect, leads to maldistribution of liquid and hence, incomplete wetting and partial utilization of catalyst in these reactors [3,4]. In addition, the ratio of the catalyst bed length to the diameter of the catal等 particle is also very low in these reactors, which along with lower liquid velocity leads to axial backmixing of liquid as against the prevalence of the almost plug type of liquid flow in commercial reactors [1.4–5]. Thus, owing to the differences in their hydrodynamics, a small-scale reactor cannot be treated as an exact replica of a commercial unit [1.4]. Therefore, the data generated in such a reactor may give misleading information for scale-up and scale-down activities and hence, lose their significance.

Sic [1] has given guidelines regarding the maximum size
of particles that can be used for different types of small-scale reactors in order to remove their drawbacks. The use of small catalyst particles obtained by crushing larger ones having commercially applied size and shape in laboratory reactors can increase the ratio of reactor diameter to catalyst particle diameter as well as the ratio of the length of the catalyst bed to the diameter of the catalyst particle, and, therefore, can solve the problems associated with these reactors. However, the practical relevance of data generated in this way is doubtful for reactions like hydroprocessing of atmospheric and vacuum gas oil, which are affected by intraparticle diffusion. Hence, this technique of testing catalysts in their crushed form in a small-scale reactor has seldom been suitable to obtain reliable data for industrial applications [1].

Two approaches have been recommended to overcome the above-mentioned problems during the testing of commercial catalysts in small reactors. The first one is to use these catalyst particles in a downflow trickle bed reactor but diluted with non-porous inert fine diluent particles. The second approach is to operate the fixed bed catalytic reactor in the upflow mode where wetting of the catalyst is almost complete. Though both these approaches have been used by different researchers for the evaluation of catalysts in laboratory or bench-scale units, these two methodologies differ in their basic nature and performance. For example, in the upflow mode of operation, liquid is in a continuous phase and gas remains in the dispersed phase whereas the situation is reversed in downflow operation. The downflow mode of operation, though resembling the commercial reactor in terms of continuous and dispersed phases, has the disadvantage of poor catalyst wetting and, to some extent, backmixing of liquid. A number of researchers have demonstrated that use of the proper size of fine diluent can overcome the limitations of laboratory trickle bed reactors, and hence, meaningful data can be generated in such reactors using the dilution technique [1,6–8]. The hydrodynamic behavior of a laboratory trickle bed reactor with catalyst diluted with fines has also been studied by a number of researchers [6,9–12].

The upflow mode of operation may not portray the commercial trickle bed reactor in terms of continuous and dispersed phases. The upflow mode of operation, though it ensures almost complete wetting of the catalyst, suffers from the serious drawback of non-ideal flow of liquid and formation of a stagnant zone inside the catalyst bed. The full wetting of the catalyst in an upflow reactor is obtained because of increased liquid holdup, which could perhaps alter the relative rates of homogeneous and heterogeneous reactions [9]. Moreover, the availability of hydrogen in a hydroprocessing process is possibly decreased in the upflow mode of operation. Few researchers have reported the superior performance of an upflow mode of operation in spite of its limitations as compared to the downflow mode [13]. Several researchers have also studied the hydrodynamic behavior of an upflow fixed bed reactor [14–16]. However, the particle sizes used in most of these studies are larger (diameter >2 mm) as compared to those of commercial hydroprocessing catalysts, and the velocities of different phases are much higher than those typically used in laboratory reactors.

Few reports are available comparing the hydrodynamics of the two modes of operation [17–20]. Khudilkar et al. [21] have conducted a comparative study between the performance of a laboratory trickle bed and an upflow reactor using the hydrogenation of alpha-methyl styrene as the test reaction. They have concluded that the superiority of any mode of operation depends on whether the reaction is liquid- or gas-limited. For example, the performance of the downflow trickle bed reactor is superior for a gas-limited reaction whereas the upflow fixed bed reactor gives advantages for liquid-limited reaction. Dudukovic et al. [22] have reviewed the comparison between the two modes of operation for different cases of gas- and liquid-limited reactions. However, most of the above studies have been conducted in such a way that the upflow mode was operated without consideration of proper diluent size.

There are only a few studies available in literature comparing the performances of upflow and downflow fixed bed reactors for different sizes of diluent. De Wind et al. [13] have compared experimentally the performances of upflow and downflow modes of operation with and
Table 1

Data on parameters used in the present investigation

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Values</th>
</tr>
</thead>
<tbody>
<tr>
<td>Average particle size of diluent (mm)</td>
<td>0.3 and 1.1</td>
</tr>
<tr>
<td>LHSV (h⁻¹)</td>
<td>0.75, 1.25, 2.0, 3.0</td>
</tr>
<tr>
<td>Gas/liquid ratio (l/l)</td>
<td>150, 500, 750, 1000</td>
</tr>
</tbody>
</table>

without using a diluent. They have reported the superior performance of the upflow mode of operation when diluted with 0.5 mm size of silicon carbide particles. However, the diluent size used in their study was higher than that used by other researchers in removal of the shortcomings of the downflow reactor of similar capacity. Wu et al. [23] have recently compared the performances between upflow and downflow modes of operation using 0.2 mm size of diluent. They did not observe any differences in the performance of the two modes when 0.2 mm size of diluent was used. Myrstad et al. [24] have also observed an equal performance of the upflow and downflow modes using 0.5–0.71 mm size of silicon carbide as diluent. However, the length and the diameter of the catalyst bed used in their study were not the same for both the modes of operation. Since there are only a few published data available comparing the performances of both upflow and downflow modes of operation over different sizes of diluent, there is a need for a systematic study in this direction.

Therefore, the main objective of this study is to compare the hydrodynamic behavior of a fixed bed reactor in the upflow and downflow modes of operation for two different sizes of diluent in the catalyst bed. Another aim of this work is to study the effects of liquid and gas flow rates on different hydrodynamic parameters for these two modes of operation using two different sizes of diluent. The hydrodynamic parameters such as operating liquid holdup, mean residence time and axial dispersion in a multiphase reactor are dependent on the geometry of the system such as reactor size and shape of the catalyst particles. These hydrodynamic parameters were obtained for the two modes of operation by the residence time distribution (RTD) technique using a catalyst volume of 100 ml and using hydrochloric acid as a tracer.

It has been reported in literature [1,6,23] that a diluent size in the range of 0.2–0.5 mm could eliminate poor catalyst wetting and axial backmixing in a bench-scale downflow trickle bed reactor. Therefore, a diluent size of 0.3 mm was selected in this study. Also, another larger diluent size with average particle diameter of 1.1 mm was used to compare its effect. The liquid and gas velocities used in this study were maintained close to those typically used for testing different hydroprocessing catalysts in a bench-scale unit. The levels of these flow rates depend on the type of catalyst to be tested. For example, if the experiments are planned for testing an atmospheric gas oil hydroprocessing catalyst, the liquid and gas flow rates may be selected to obtain an LHSV of 1–3 h⁻¹ and a gas/liquid ratio (commonly known as hydrogen/oil ratio) of ~150–250 (v/v). Similarly, vacuum gas oil hydrocracking or atmospheric residue hydrocracking catalyst requires typical ranges of LHSV and gas/liquid ratio of 0.75–1.5 h⁻¹ and 500–750 (v/v), respectively. On the other hand, the testing of vacuum gas oil hydrocracking and vacuum residue hydrocracking catalyst require an LHSV in the range of 0.5–0.75 h⁻¹ and a gas/liquid ratio of ~1000 (v/v). The liquid and gas flow rates in this study cover the entire range of testing all these hydroprocessing catalysts, and are given in Table 1.

2. Experimental

The simplified schematic diagrams of the experimental setup used for the upflow and downflow modes of operation are shown in Fig. 1(a) and (b), respectively. The setup consists of a reactor, a system for feeding controlled and measured amounts of gas and liquid, a pneumatic system for injecting the tracer, and a conductivity measurement instrument for detecting the concentration of the tracer.

The reactor was a stainless steel tube with length and
internal diameter of 70 and 2.5 cm, respectively. Cylindrical pellets of alpha alumina with 1.2 mm diameter was used as catalyst for packing the reactor. The alpha alumina used in pellet form was highly inert having a surface area of \(-10 \text{ m}^2/\text{g}\). Preliminary tests were performed to confirm that alpha alumina did not react with dilute hydrochloric acid. Because of the inertness of alpha alumina towards dilute hydrochloric acid, the chemisorption of hydrochloric acid on alpha alumina could be neglected. Two different sizes of silicon carbide were used as diluent. 100 ml each of alumina and silicon carbide were loaded one after another in lots of 10 ml each with intermittent vibration of the reactor. This method of loading the reactor provides the advantage of avoiding the segregation by size and mass that would result from pre-mixing the particles and then trying to fill the bed. Silicon carbide (having an average particle size of 0.3 mm) and glass wool were also used at the top and bottom of the reactor for the proper distribution of liquid as well as supporting the alumina bed.

In the past air has been used as gas phase and water as liquid phase for carrying out RTD studies of packed beds since these fluids are inexpensive and non-hazardous [14,15,17]. Therefore, in the present study air and demineralized water were used as the gas phase and liquid phase, respectively. The flow rate of air from a cylinder was controlled by a precision needle valve and measured either by a mass flow meter (for flow range of 0–150 l/h) or by a rotameter (for flow range of 150–300 l/h). One pre-calibrated positive displacement pump was used to feed water into the reactor. The hydrochloric acid as a tracer was introduced into the reactor as a pulse through a pneumatic injection system. The amount of hydrochloric acid injected was adjusted during some preliminary experiments by changing the quantity of hydrochloric acid and observing the response pattern of the same at the outlet. The concentration of the tracer at the upstream point (for upflow mode) or down-stream point (for downflow mode) was measured by a conductivity-measuring instrument (see Fig. 1(a) and (b)). The distance between the injection system and tip of the conductivity-measuring probe was kept constant at 35 cm for both upflow and downflow modes of operation.

Before starting the experiment, the bed was fully wetted by passing water at slightly higher rate of 300 ml/h. Then, the flow rates of air and water were adjusted to the desired values. It took approximately 30 min to attain steady state. The attainment of steady state was determined by measuring the flow rates of gas and liquid at the outlet. It was seen that the flow rates did not change after 30 min of operation. The pulse of the tracer was then injected through the pneumatic injection system. The concentration of the tracer at the upstream or downstream point was determined by the conductivity-measurement instrument.

The operating liquid holdup of the packed bed was determined as follows. The bed was fully wetted and then the reactor was run at the set required liquid and gas flow rates for 30 min. Then, the gas and liquid flows to the reactor were simultaneously stopped by closing the inlet and outlet valves. The total free liquid in the reactor was drained in a liquid collector and measured. The operating holdup of liquid was expressed as volume of liquid/total volume of the packed bed.

3. Results and discussion

The reproducibility of the injection system was verified by repeatedly carrying out one experiment. The concentration vs. time data of the hydrochloric acid tracer obtained from the experiments was normalized and plotted as an E-curve. The values of bed Peclet number and mean residence time of the liquid were calculated from the RTD data.

3.1. Theoretical calculations

In this study, the RTD of the tracer is represented by the E-curve, which describes the normalized concentration data of the tracer against time. Such a normalized E-curve is defined as:

\[
\int_0^\infty E(t) \, dt = 1
\]  

(1)

Two parameters, viz. mean residence time and bed Peclet number of liquid, were calculated from the RTD data of the tracer. Peclet number was selected as a measure of axial dispersion of the liquid phase. The axial dispersion model has been used for calculating the mean residence time and Peclet number of liquid [25,26]. The mean residence time, \(t_m\), is obtained from the following equation:

\[
t_m = \frac{\int_0^\infty t C \, dt}{\int_0^\infty C \, dt}
\]  

(2)

The bed Peclet number (henceforth only Peclet number) of liquid is the reciprocal of the dispersion number \((D/uL)\), i.e.

\[\text{Pe} = \frac{1}{D/uL}\]  

(3)

The dispersion number is calculated as follows:

\[
\frac{D}{uL} = \frac{\sigma_{\bar{v}}^2}{2}
\]  

(4)

and

\[
\sigma_{\bar{v}}^2 = \left(\frac{u}{L}\right)^2 \sigma^2
\]  

(5)

where \(\sigma^2\) is the variance of the E-curve and is calculated from the following equation.

\[
\sigma^2 = \frac{\int_0^\infty t^2 C \, dt}{\int_0^\infty C \, dt} - (t_m)^2
\]  

(6)
Since most of the peaks were symmetrical in nature, the use of Eq. (4) can be justified. The operating liquid holdup of liquid, $h$, is defined as the ratio of the volume of the free-drained water to the total volume of the packed bed.

### 3.2. Effect of diluent size on the nature of the E-curve

The RTD studies for upflow and downflow modes of operation using 0.3 and 1.1 mm size of diluent were carried out at two different levels of LHSV at a constant gas/liquid volume ratio of 500. The $E$-curves (i.e., the normalized concentration curves) for all these cases are shown in Figs. 2–5 along with the values of mean residence time and Peclet number of liquid for these two modes of operation. For example, the downflow mode provided a much lower residence time (34 min) to the liquid as compared to that (88 min) for the upflow mode of operation, probably owing to channeling of liquid in the former case. However, the liquid backmixing was much higher in the upflow mode as indicated by the lower values of Peclet number. On the hand, the upflow mode of operation gave a much higher liquid holdup as compared to the downflow mode (see Table 2). The differences between the two modes of operation also existed at a higher LHSV of 3 h⁻¹ but to a slightly lower extent (see Fig. 3). The higher superficial velocity at 3 h⁻¹ reduced the channeling of liquid in downflow operation and reduced backmixing in upflow mode. Also, a higher liquid holdup at higher LHSV was
obtained in the downflow mode of operation. These phenomena resulted in a slight decrease in the difference between the two modes of operation.

When the catalyst was loaded with a smaller size of diluent (0.3 mm), the values of mean residence time, Péclet number and liquid holdup increased for the downflow mode (see Figs. 4 and 5). As a result of this, the hydrodynamic behavior for both upflow and downflow modes of operation was improved. Sie [1] has also recommended that the use of about 0.8 mm diluent size could remove the shortcomings of a bench-scale trickle bed unit containing 150 ml of catalyst. Van Klinken and van Dongen [6] have also reported that the use of 0.2 mm size of diluent in a bench-scale trickle bed unit increased the mean residence time and liquid holdup as compared to that in an undiluted bed, and reduced axial backmixing.

The present study indicated the change in the behavior of the upflow mode on using a smaller size of diluent. For example, the use of a smaller size of diluent increased the values of Péclet number (see Figs. 4 and 5) and moderated excessive liquid holdup (see Table 2) and thus eliminated the limitations of the upflow mode of operation. The differences in the nature of E-curves for the two modes of operation under similar operating conditions of liquid and gas velocities thus disappeared for the smaller size of diluent (see Figs. 4 and 5). The values of mean residence time, Péclet number and liquid holdup were nearly the same for the two modes of operation. Thus, the discrepancies

Fig. 4. E-curves for upflow and downflow modes of operation at an LHSV of 0.75 h$^{-1}$ and a gas/liquid ratio of 500 l/l using 0.3 mm size of diluent.

Fig. 5. E-curves for upflow and downflow modes of operation at an LHSV of 3.0 h$^{-1}$ and a gas/liquid ratio of 500 l/l using 0.3 mm size of diluent.
between the two modes of operation could be removed by using a diluent size of 0.3 mm. Recently, Wu et al. [23] have also reported that a diluent having an average particle size of about 0.2 mm does neutralize the differences between the two modes of operation for both gas- and liquid-limited reactions. Thus, the use of a smaller size of diluent could remove the drawbacks of both upflow fixed bed and trickle bed reactors and make them suitable tools for generating reliable and meaningful data for scale-up and scale-down activities.

### 3.3. Effect of liquid space velocity and gas/liquid ratio on axial dispersion

The effect of liquid space velocity (0.75–3.0 h⁻¹) on liquid Peclet number was studied for both sizes of diluent at a constant gas/liquid volume ratio of 500. The results are given in Fig. 6. The figure indicates that the values of Peclet number showed dramatic improvement when a smaller size of diluent is used for both modes of operation and for the entire range of space velocities. Stieigel and Shah [14] have reported that during the upflow mode of operation of the fixed bed, the Peclet number of liquid increased with decrease in particle size. It is also apparent from the figure that the values were almost equal for both the modes and close to the values where almost plug flow of the liquid can be assumed. However, the Peclet number value increased with the increase in LHSV for both the modes of operation indicating the reduction of backmixing with higher liquid flow rate. Also, the values of the Peclet number for the downflow while using larger diluent size were higher than those with upflow mode for the entire range of space velocities studied indicating the presence of higher liquid backmixing in the former case. However, the figure also indicates that the use of a fine size of diluent also reduced the dependency of Peclet number on space velocity. Hiuta et al. [17] have also reported that for a downflow trickle bed reactor using a larger size of catalyst particles, the values of the dispersion coefficient, which is inversely proportional to the Peclet number, decreased by increasing the liquid flow rate. Similarly, Stieigel and Shah [14] have found an increasing trend in the Peclet number with increasing liquid velocity for the upflow mode of operation when a larger size of particles was used.

The effect of gas/liquid ratio was studied at a constant LHSV of 0.75 h⁻¹. Fig. 7 represents the variation of Peclet number with gas/liquid ratio. The figure shows that the values of Peclet number were higher for the downflow mode as compared to those obtained for the upflow mode for the range of gas/liquid ratio studied. The gas flow rate had negligible effect on the Peclet number for both the larger and smaller sizes of diluent when the reactor was operated in the downflow mode. However, the Peclet number was a very strong decreasing function of gas/liquid ratio for the upflow mode when a larger size of diluent was used. This might be due to the increase of circular motion of liquid causing backmixing with increasing gas flow rate. Cassanello et al. [18] have also reported that the gas velocity affects the axial dispersion coefficient for upflow operation while its effect is not significant for downflow. It is also

### Table 2
Operating liquid holdups for upflow and downflow modes of operation using gas/liquid ratio of 500 l/l

<table>
<thead>
<tr>
<th>Flow pattern</th>
<th>Experimental condition</th>
<th>LHSV (h⁻¹)</th>
<th>Operating liquid holdup (ml/ml)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Upflow</td>
<td>1.1</td>
<td>0.75</td>
<td>0.29</td>
</tr>
<tr>
<td>Downflow</td>
<td>1.1</td>
<td>0.75</td>
<td>0.055</td>
</tr>
<tr>
<td>Upflow</td>
<td>1.1</td>
<td>3.0</td>
<td>0.032</td>
</tr>
<tr>
<td>Downflow</td>
<td>1.1</td>
<td>3.0</td>
<td>0.13</td>
</tr>
<tr>
<td>Upflow</td>
<td>0.3</td>
<td>0.75</td>
<td>0.2</td>
</tr>
<tr>
<td>Downflow</td>
<td>0.3</td>
<td>0.75</td>
<td>0.18</td>
</tr>
<tr>
<td>Upflow</td>
<td>0.3</td>
<td>3.0</td>
<td>0.22</td>
</tr>
<tr>
<td>Downflow</td>
<td>0.3</td>
<td>3.0</td>
<td>0.19</td>
</tr>
</tbody>
</table>

![Fig. 6. Effect of LHSV on liquid Peclet number for upflow and downflow modes of operation at a gas/liquid ratio of 500 l/l for two sizes of diluent.](image)
observed in the present study that the use of a smaller size of diluent moderates the effect of gas flow rate on the liquid Peclet number.

3.4. Effect of liquid space velocity and gas/liquid ratio on mean residence time of liquid

The effect of LHSV on mean residence time of the liquid was studied at a constant gas/liquid volume ratio of 500 and the results are shown in Fig. 8. The mean residence time of the liquid decreased with increase in liquid space velocity for both the modes of operation. However, the mean residence time was a stronger function of space velocity for the upflow mode of operation as compared to the downflow mode when a larger size of diluent was used. The higher values of mean residence time for the upflow mode could be due to excessive liquid holdup and formation of a stagnant zone inside the catalyst bed. The higher mean residence time in the upflow mode could definitely provide a better utilization of catalyst. At the same time, the liquid would also spend undesired longer residence time when not in contact with the catalyst. As a result, a number of undesirable thermal reactions would take place during this period. On the other hand, in the downflow mode of operation, the liquid would not get the required amount of time for reaction. These discrepancies for both the modes of operation could be removed when a smaller size of diluent was used. The mean residence time for the downflow mode was increased due to the higher liquid holdup resulting in better utilization of catalyst. The use of a smaller size of diluent also decreased the porosity of the bed, which in turn reduced the excessive residence time of liquid in the upflow mode of operation. This could help in the reduction of undesirable non-catalytic reactions in the upflow mode of operation. Interestingly, the mean residence time for both the modes of operation nearly coincided for a diluent size of 0.3 mm. However, the trend of increasing residence time with space velocity was also observed for a smaller size of diluent.

Fig. 9 represents the variation of liquid mean residence time with gas velocity at a constant LHSV of 0.75 h⁻¹. It is
evident from the figure that when a larger size of diluent was used, the mean residence time increased with gas/liquid ratio for the upflow mode of operation but remained nearly constant for the downflow mode. The increased gas flow rate in the upflow mode perhaps induced circulatory motion of liquid inside the catalyst bed so that the liquid spent more time in the reactor. The independent nature of the mean residence time of liquid with gas flow rate for the downflow mode using a larger size of particles could be attributed to the separate independent pathways followed by the two phases. On the other hand, when a smaller size of diluent was used, the mean residence time decreased slowly with gas flow rate. This could be due to the slow draining effect of liquid caused by the high flow rate of gas. The values of the residence time for the two modes of operation approached each other for the smaller size of diluent.

3.5. Effect of liquid space velocity and gas/liquid ratio on operating liquid holdup

The operating liquid holdup is the portion of liquid that drains out of the catalyst bed when both gas and liquid flows are stopped. The operating liquid holdup is an important parameter influencing the rate of reaction in a gas–liquid–solid multiphase reactor. In the present investigation, the operating holdup was determined by the method of free draining as discussed earlier in Section 2. The effect of liquid space velocity on holdup was determined for a fixed gas/liquid volume ratio of 500. The values of operating liquid holdup were determined for two sizes of diluent and the results are plotted in Fig. 10. The figure shows that operating holdup increased with liquid space velocity for both the modes of operation when the catalyst was diluted with larger particle size.
Thus, higher liquid flow rate could increase the reaction rate when the catalyst was diluted with larger particle size. The use of a smaller size of diluent increased the holdup for the downflow mode of operation. Van Klinken and van Dongen [6], De Wind et al. [13], and Al-Dahan and Dudukovic [9] have reported similar results. On the other hand, the holdup for the upflow mode of operation was reduced when a smaller size of particles was used. This is slightly contrary to the results obtained by De Wind et al. [13]. They have reported that the use of a smaller size of diluent did not have any effect on holdup during the upflow mode of operation. However, Stiegel and Shah [14] have reported the decrease of liquid holdup with the decrease in particle size for the upflow mode of operation. It has been observed in this study that when the catalyst bed was diluted with a smaller size of particles, the effect of space velocity on operating liquid holdup was very small or negligible for both the modes of operation. De Wind et al. [13] have also reported that for the diluted bed of catalyst, the effect of space velocity on liquid holdup was negligible. However, they have observed an increasing trend in liquid holdup with space velocity for the downflow mode. This could be due to the larger size of diluent (diameter of 0.5 mm) used in their study.

The effect of gas/liquid ratio on operating liquid holdup for a constant value of LHSV (0.75 h\(^{-1}\)) is shown in Fig. 11. The results showed that for the downflow mode of operation, the liquid holdup was independent of gas flow rate for both the sizes of diluent. On the other hand, the liquid holdup decreased with increasing gas flow rate for the upflow mode when the bed was packed with a larger size of diluent. The effect of gas flow rate on liquid holdup in the upflow mode could be removed by using a smaller size of diluent. However, when a smaller size of particle was used as diluent in the catalyst bed, the values of liquid holdup for both the modes of operation were nearly the same for the entire range of gas/liquid ratios studied.

4. Conclusions

The following conclusions are drawn from the present study:

(i) Large differences in hydrodynamics between upflow and downflow modes of operation in a fixed bed reactor existed when a larger size of particle (average diameter of 1.1 mm) was used as diluent in the catalyst bed.

(ii) The differences between the two modes of operation were reduced slightly at higher liquid flow rate even when a larger size of diluent was used.

(iii) The use of a smaller size of diluent with average particle diameter of 0.3 mm neutralized mostly all the differences between the two modes of operation. Hence, both the upflow as well as the downflow modes of operation using 0.3 mm of diluent particle can be used for the scale-up and scale-down activities.

(iv) The liquid flow rate had a strong influence on the liquid Peclet number for both modes of operation when a larger size of diluent was used. The effect was reduced substantially in the case of smaller particle size. The gas flow rate had a significant effect on the Peclet number for the upflow mode of operation using a larger size of diluent.

(v) The use of a smaller size of particles moderated the effect of liquid and gas flow rates on the mean residence time and operating hold of the liquid phase.
References