Investigation of the Hydrodynamics of Fixed Bed Reactor: Co-current Upflow

by

Nurhafizati Bt. Abdul Manan

Dissertation submitted in partial fulfillment of
the requirements for the
Bachelor of Engineering (Hons)
(Chemical Engineering)

JULY 2005

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CERTIFICATION OF APPROVAL

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A project dissertation submitted to the
Chemical Engineering Programme
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Approved by,

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Project Supervisor

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July 2005

CERTIFICATION OF ORIGINALITY

This is to certify that I am responsible for the work submitted in this project, that the original work is my own except as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

NURHAFIZATI ABDUL MANAN

ABSTRACT

The main objective of this project is to investigate the hydrodynamic characteristics of co-current upflow of gas and liquid in a fixed bed reactor. This is an experimental based project utilizing the packed bed reactor for residence time distribution (RTD) studies. In this project, the RTD of a bench-scale multiphase was studied using air as a gaseous phase and water as a liquid phase. The ranges of air and water velocities are kept at such levels as to simulate the hydrogen/oil ratios of typical bench-scale hydroprocessing units. The experiments are conducted in upflow mode of operation in the reactor, with increasing gas/liquid ratio. The effects of gas and liquid velocities on different hydrodynamic parameters such as pressure drop and operating liquid holdup are investigated. Three moments analysis, which are mean residence time, variance, and skewness, are evaluated in order to characterize the RTD. Other parameters such as bed Peclet number of liquid and stagnant zone volume were also investigated with variation of gas and liquid velocities as to measure the efficiency of the reactor. The discrepancies in experimental results suggested that there are conditions to be altered in order to eliminate the inconsistency.

ACKNOWLEDGEMENTS

The author would like to be grateful to The Most Merciful, Allah S.W.T, for giving the chance to conduct and successfully complete this final year research project.

The author's outmost appreciation is for the project supervisor, Mrs. Norhayati bt. Mellon, who has given the trust and supervision to conduct such outstanding experiment for this final year research project.

To Universiti Teknologi PETRONAS, the author is thankful for all the equipment provided in order to complete the experiments. Also to laboratory technicians, who have helped in various ways in providing sufficient assistance and supports.

To supportive colleagues and families, the motivations and advises are mostly appreciated by the author.

Also to all individual involved directly or indirectly in this project, the author is thankful for the contribution in progression and completion of this final year research project.

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ABBREVIATIONS AND NOMENCLATURES

LPM	Liter per minute
d_c	Column diameter
d_p	Particle diameter
ΔP	Pressure drop
H_o	Operating liquid holdup
L_c	Height of column
ε	Fractional void volume
μ_g	Absolute viscosity of gas
μ_l	Absolute viscosity of liquid
U_g	Superficial gas velocity
U_l	Superficial liquid velocity
D_p	Effective particle diameter
G_g	Mass flow rate of gas
G_l	Mass flow rate of liquid
ρ_g	Density of gas
ρ_l	Density of liquid
g	Gravitational acceleration
Re	Reynolds number
D	Axial dispersion coefficient
L	Distance between the tracer injection point and conductivity measurement probe
u	Mean real liquid velocity
t_m	Mean residence time
σ^2	Variance of the E curve

s^3	Skewness of the E curve
σ_θ^2	Variance of the E curve for dimensionless time units
C	Concentration
C_i	Concentration at time t_i
HRT	Hydraulic retention time
T_d	Theoretical mean residence time
Q	Volumetric flow rate of liquid
V	Volume of column

CHAPTER 1
INTRODUCTION

1.1 BACKGROUND OF STUDY

Simultaneous gas-liquid flow through packed beds is frequently encountered in chemical process equipment and is practiced as countercurrent flow, co-current downflow (trickle bed) or co-current upflow. Countercurrent flow is preferred for mass transfer operations, while the co-current flow of gas and liquid phases is frequently adopted in multiphase reactors, since the throughput is not limited by flooding.

Multiphase packed-bed reactor with two-phase upflow is used in gas-liquid and gas-liquid-solid processes that require a high ratio between the liquid and gas flow rates. It is also used for processes with relatively large liquid residence time in order to achieve the necessary degree of conversion. It is also used when the heat of reaction is high, due to the large liquid holdup and their improved radial liquid mixing and radial heat transfer. The upflow operation is also advantageous in cases where the ratio of column diameter over particle diameter, d_c / d_p is relatively small, because then the liquid-solid contact is more effective than in trickle-bed operation. A ratio of the column to the catalyst diameter smaller than 15 inch the other types of fixed-bed reactor such as trickle beds, cause unsatisfactory liquid distribution due to wall bypassing. One of the drawbacks of the upflow fixed-bed reactors is that the flow behavior of the liquid is non ideal and that backmixing is considered to be more important than in trickle beds. This may give better heat transfer, but larger axial mixing would give poorer conversion.

Considerable work has been reported in literature on co-current downflow of phases; very little however has been reported on co-current upflow in spite of the specific advantages of the downflow model; i) the liquid distribution is radially uniform; this promote efficient distribution of heat and its transfer to and from the wall when desired and prevents the formation of dry spots, ii) the larger liquid holdup gives higher production rate for a given size of reactor, and iii) liquid-side mass transfer coefficients are higher.

Few reports are available comparing the hydrodynamics of the two modes of operation. It is concluded that the superiority of any mode of operation depends on whether the reaction is liquid- or gas-limited, i.e. the performance of trickle bed reactor is superior for a gas-limited reaction, whereas upflow fixed bed reactor gave advantages for liquid-limited reaction.

1.2 PROBLEM STATEMENT

1.2.1 Problem Identification

Hydrodynamics study in a fixed bed reactor is vital in obtaining relevant data for reactor scale up as well as investigating the performance of the reactor. Co-current upflow arrangement received considerable interest due to the fact that almost complete catalyst particle wetting is achieved which enhance the reaction rate of the process. However, the continuous liquid phase and dispersed gas phase will probably result to non-ideal flow and possibility of stagnant zone in the reactor, thus resulted poor reactor performance.

1.2.2 Significant of Project

The RTD of a bench-scale multiphase reactor has been investigated mainly for industrial scale-up purposes. The successful design of commercial reactors involved generation of reliable data in laboratory-scale reactors and scaling up of these data for larger units. The study of effects of gas and liquid flow rates on various hydrodynamics parameters utilizing the RTD technique using tracer is important for the performance of the reactor. Future work may be based on the development of this study.

1.3 OBJECTIVE AND SCOPE OF STUDY

1.3.1 Objectives

- To investigate the effect of gas/liquid ratio on pressure drop, operating liquid holdup, axial liquid dispersion and stagnant zone volume.
- To characterize residence time distribution (RTD) of reactor by three moments analysis (mean residence time, variance, and skewness).

1.3.2 Scope of Study

The scope of this project is to utilize the existing laboratory experiment on RTD study in packed bed reactor, at the Reaction Teaching Lab. However, few modifications are done in order to study the hydrodynamics of the packed bed reactor. The modifications are mainly on the variation of the gas and liquid flow rates, to suit the scaled down of commercial reactors. These variations are analyzed based on the response of hydrodynamics parameters.

1.3.3 Feasibility of the Project within the Scope and Time frame

This experiment is conducted as nearly similar to the existing procedures. Due to time constraint, the RTD study may not be extended to the effect of different catalyst particle and diluent size, or effect of different packing.

The experiments emphasized mainly on the investigation of the hydrodynamic characteristics of co-current upflow of gas-liquid reactor. With the available equipment, the RTD study can be conducted to evaluate the three moments analysis, to evaluate the effects of gas/liquid ratio on axial dispersion and stagnant zone volume. Consecutively, the effect of gas/liquid ratio on operating liquid holdup and pressure drop are also investigated.

CHAPTER 2

LITERATURE REVIEW / THEORY

2.1 PRESSURE DROP AND OPERATING LIQUID HOLDUP IN PACKED BED REACTOR

2.1.1 Pressure Drop Correlation

The pressure loss accompanying the flow of gas through packed columns has been the subject of many theoretical analysis and experiment investigations to try to find a suitable mathematical expression to predict the pressure drop caused by both kinetic and viscous energy losses.

A very successful attempt is that of Ergun [1] which is included in the Perry's Handbook. The Ergun equation is;

$$\frac{\Delta P}{L_c} g \quad = \quad 150 \frac{(1-\varepsilon)^2}{\varepsilon^3} \frac{\mu U_g}{D_p^2} \quad + \quad 1.75 \frac{1-\varepsilon}{\varepsilon^3} \frac{G_g U_g}{D_p} \qquad (1)$$

The Ergun equation gave very good results in the whole range of Reynolds numbers from 1 to 100,000. Also, it should be noted that the effective diameter is equal to real diameter only when the particles are spherical; for all other shapes the D_p is define as $D_p = 6V_p/A_p$, where V_p is the volume of particles and A_p is the external surface of particle. Ergun equation assumed equivalent pressure drop regardless of any type of flow regimes.

Turpin and Huntington [2] also gave a single relation for pressure drop valid for all the regimes, in terms of a dimensionless parameter, $Z = \mathrm{Re}_g^{1.167} / \mathrm{Re}_l^{0.767}$. On the other hand, Varma et al. [3] developed an empirical equation for predicting the transition from one flow regime to another. It presented typical variation of frictional pressure drop with liquid and gas flow rates respectively for bubble flow, pulse flow and spray flow. It is seen that though the pressure drop increases with the gas and liquid rates in all the regimes, its variation differs for the different flow regimes. For example, the pressure drop increases rapidly with the gas rate in the spray flow as compared to its increase in pulse flow and in bubble flow.

However, it is noted that the transition between the flow regimes is not sharp and occurred over a small range in gas and liquid flow rates. Thus, all flow regimes in co-current upflow can be assumed equivalent in this experiment, as in Ergun [1] principle, which has been used widely in several researches.

In experiment, pressure drop is directly obtained from the differential pressure reading at control panel or via the Data Acquisition System (DAS). The pressure drop reading is taken at time interval of one minute, and readings are averaged for one value of pressure drop for every variation of gas and liquid flow rates.

2.1.2 Operating Liquid Holdup

Liquid holdup may well be considered as the basic liquid-side dependent variable in packed tower operation. Holdup has a direct influence on factors such as liquid-phase mass transfer, loading behavior, and gas-phase pressure gradient. Researchers measured holdup, with or without gas flow, and have produced empirical description of their results. Only the correlation of Buchanan [4] is in dimensionless form and can claim any generality over a wide range of Reynolds numbers (from 0.01 to 1000). This correlation is appropriate for experimental research. It applied to ring packing operating below the load point and correlated all literature data to about ±20%. The Buchanan equation is consisted of two dimensionless terms, the 'Film number' and 'Froude number';

$$H_o = 2.2\left(\frac{\mu_l U_l}{g\rho_l D_p^{\,2}}\right)^{1/3} + 1.8\left(\frac{U_l^2}{gD_p}\right)^{1/2} \qquad (2)$$

Experimentally, the operating liquid holdup of liquid is the portion of liquid that is drained 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. The operating liquid holdup of liquid is defined as the ratio of the volume of the free-drained water to the total volume of the packed bed.

Chander et al. [5] determined the effect of liquid space velocity on holdup and proved that the operating liquid holdup increased with liquid space velocity. Thus, higher liquid flow rate could increase the reaction rate. Also, the studies also showed that liquid holdup for the upflow mode of operation was reduced when smaller size of particles was used. Stiegel and Shah [6] also have reported the decrease of liquid holdup with the decrease in particle size for the upflow mode of operation. It is also observed that studies by Chander et al. [5] showed that when catalyst bed was diluted with smaller size of particles, the effect of space velocity on operating liquid holdup was very small or negligible.

Chander et al. [5] also studied the effect of gas/liquid ratio on operating liquid holdup, which resulted that the liquid holdup decreased with increasing gas flow rate for the upflow mode when the bed was packed with a larger size of diluent.

2.2 RESIDENCE TIME DISTRIBUTION (RTD) STUDY IN REACTOR

The RTD of a reactor is a characteristic of the mixing that occurs in the chemical reactor. There is no axial mixing in a plug-flow reactor (PFR), and this omission is reflected in RTD which is exhibited by this class of reactors. The CSTR (constant stirred type reactor) is thoroughly mixed and possesses a far different kind of RTD than the plug-flow reactor. The RTD exhibited by a given reactor yields distinctive clues to the type of mixing occurring within it and is one of most informative characterizations of the reactor.

2.2.1 Measurement of the RTD

The RTD is determined experimentally by injection of an inert chemical, molecule, or atom, called a tracer, into the reactor at some time, $t = 0$ and then measuring the tracer concentration, C, in the effluent stream as a function of time. In addition to being a non-reactive species that is easily detectable, the tracer should have physical properties similar to those of the reacting mixture and be completely soluble in the mixture. The latter requirements are needed so that the behavior of tracer will honestly reflect that of the material flowing through the reactor. The two most used methods of injection are pulse input and step input;

2.2.1.1 Pulse Input

In a pulse input, an amount of tracer N_o is suddenly injected in one shot into the feedstream entering the reactor in as short a time as possible. The outlet concentration is then measured as a function of time. Typical concentration-time curves at the inlet and outlet of an arbitrary reactor are shown in ***Figure 2.1***. The effluent concentration-time curve is referred to as the C curve in RTD analysis. The injection of a tracer pulse shall be analyzed for a single-input and single-output system in which only flow (i.e. no dispersion) carries the tracer material across system boundaries. First, an increment of time Δt is chose to be sufficiently small that the concentration of tracer, $C(t)$, exiting between time t and $t + \Delta t$ is essentially constant. The amount of tracer material, ΔN, leaving the reactor between time t and $t + \Delta t$ is then

$$\Delta N = C(t) v \Delta t \qquad (3)$$

where v is the effluent volumetric flow rate. In other words, ΔN is the amount of material that has spent time between time t and $t + \Delta t$ in the reactor. If the term is divided by the total amount of material that was injected into reactor, N_o, then

$$\frac{\Delta N}{N_o} = \frac{vC(t)}{N_o} \Delta t \qquad (4)$$

which represents the fraction of material that has a residence time in the reactor between time t and $t + \Delta t$.

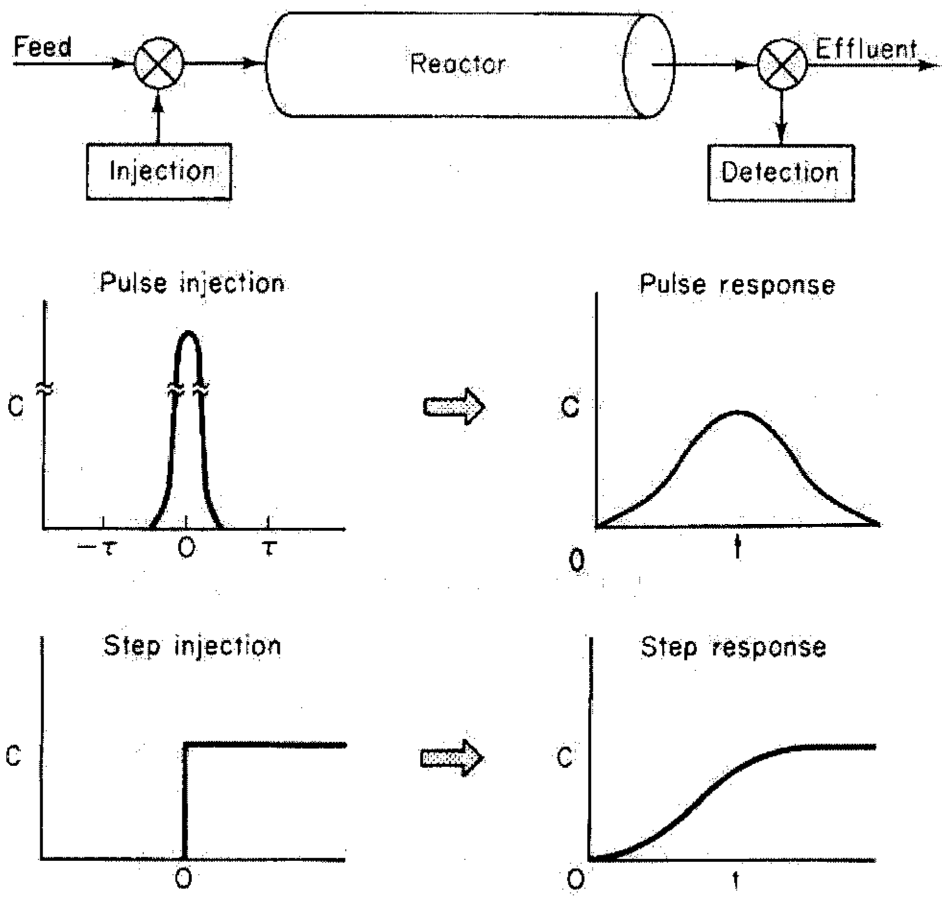

Figure 2.1 RTD measurements.

For pulse injection, it is defined

$$E(t) = \frac{vC(t)}{N_o} \qquad (5)$$

so that

$$\frac{\Delta N}{N_o} = E(t)\Delta t \qquad (6)$$

The quantity $E(t)$ is called the residence-time distribution function. It is the function that describes in a quantitative manner how much time different fluid elements have spent in the reactor.

If N_o is not known directly, it can be obtained from the outlet concentration measurements by summing up all the amounts of materials, ΔN, between time equal to zero and infinity. Writing equation (3) in differential form yields,

$$dN = vC(t)dt \qquad (7)$$

and then integrating,

$$N_o = \int_0^\infty vC(t)dt \qquad (8)$$

The volumetric flow rate is usually constant, so $E(t)$ can be defined as

$$E(t) = \frac{C(t)}{\int_0^\infty C(t)dt} \qquad (9)$$

The integral in the denominator is the area under the C curve.

An alternative way of interpreting the residence-time function is in its integral form:

$$\boxed{\begin{array}{l}\text{Fraction of material leaving the}\\ \text{reactor that has resided in the}\\ \text{reactor for time between } t_1 \text{ and } t_2\end{array}} = \int_{t_1}^{t_2} E(t)dt \qquad (10)$$

It is known that the fraction of all the material that has resided for a time t in the reactor between $t = 0$ and $t = \infty$ is 1; therefore,

$$\int_0^{\infty} E(t)dt = 1 \tag{11}$$

The principal potential difficulties with the pulse technique lie in the problems connected with obtaining a reasonable pulse at a reactor's entrance. The injection must take place over a period which is very short compared with residence times in various segments of the reactor or reactor system, and there must be a negligible amount of dispersion between the point of injection and the entrance to the reactor system. If these conditions can be fulfilled, this technique represents a simple and direct way of obtaining the RTD.

There are problems when the concentration-time curve has a long tail because the analysis can be subject to large inaccuracies. This problem principally affects the denominator of the right-hand side of equation (9), i.e. the integration of the $C(t)$ curve. It is desirable to extrapolate the tail and analytically continue the calculation. The tail of the curve may sometimes be approximated as an exponential decay. The inaccuracies introduced by this assumption are very likely to be much less than those resulting from either truncation or numerical imprecision in this region.

2.2.2 Characteristics of the RTD

Sometimes $E(t)$ is called the exit-age distribution function. If the 'age' of an atom is regarded as the time it has resided in the reaction environment, the $E(t)$ concerns the age distribution of the effluent stream. It is the most used of the distribution functions connected with reactor analysis because it characterizes the lengths of time various atoms spend at reaction conditions.

Figure 2.2 illustrates typical RTDs resulting from different reactor situations. ***Figure 2.2 (a)*** and ***(b)*** correspond to nearly ideal PFRs and CSTRs respectively. In ***Figure 2.2 (c)***, it is observed that a principal peak occurs at a time smaller than the space-time, $\tau = V/v$ (i.e. early exit of fluid) and also that fluid exits at a time greater than space time τ. This curve is representative of the RTD for a packed-bed reactor with channeling and dead zones. One scenario by which this situation might occur is shown in ***Figure 2.2 (d)***. ***Figure 2.2 (e)*** shows the RTD for the CSTR in ***Figure 2.2 (f)*** which has dead zones and bypassing. The dead zone serves to reduce the effective reactor volume indicating that the active reactor volume is smaller than expected.

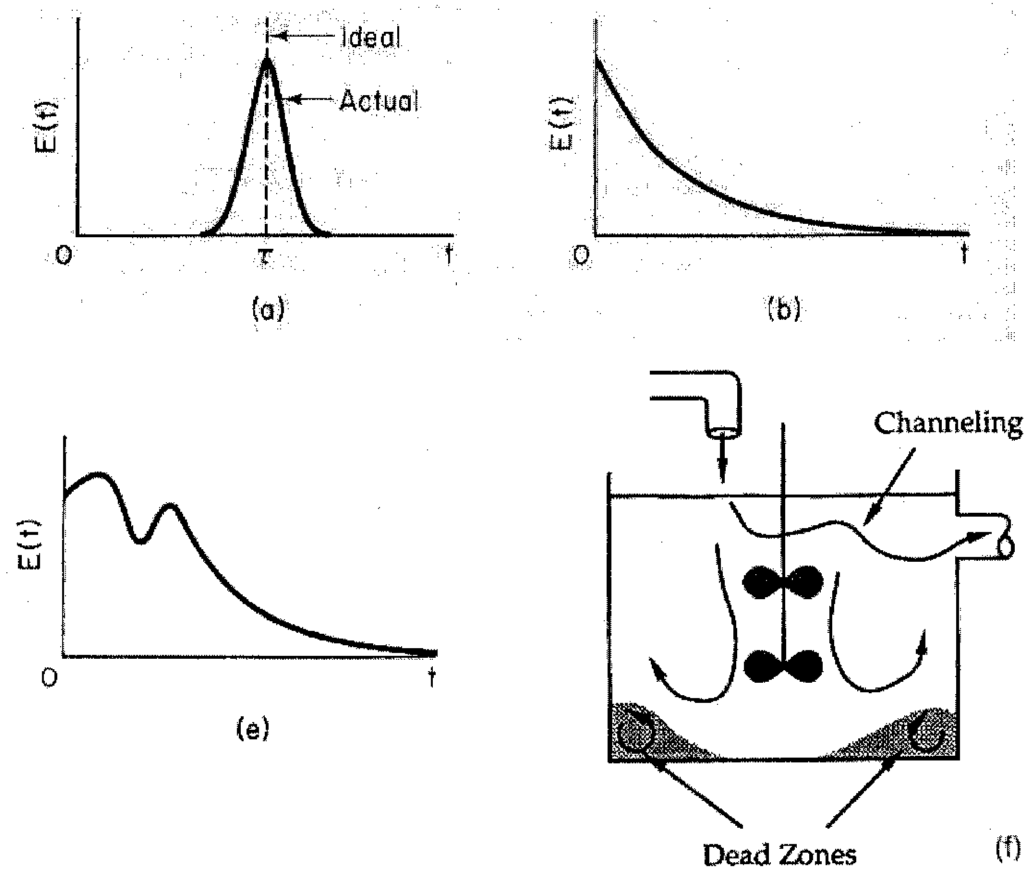

Figure 2.2 (a) RTD for near plug flow reactor; (b) RTD for near perfectly mixed CSTR; (c) RTD for packed-bed reactor with dead zones and channeling; (d) packed-bed reactor; (e) tank reactor with short-circuiting flow (bypass); (f) CSTR with dead zone.

2.2.2.1 Mean Residence Time

A parameter frequently used in analysis of ideal reactors is the space-time or average residence time τ, which is defined as being equal to V/v. It can be shown that no matter what RTD exists for a particular reactor, ideal or non-ideal, this nominal holding time τ, is equal to the mean residence time, t_m.

As is the case with other variables described by its distribution functions, the mean value of the variable is equal to the first moment of the RTD function, $E(t)$. Thus, the first moment is the mean residence time,

$$t_m = \frac{\int_0^\infty tE(t)dt}{\int_0^\infty E(t)dt} = \int_0^\infty tE(t)dt = \frac{\int_0^\infty tC(t)dt}{\int_0^\infty C(t)dt} = \frac{\sum t_i C_i \Delta t_i}{\sum C_i \Delta t_i} \qquad (17)$$

Chander et al. [5] determined the effect of liquid hourly space velocity on mean residence time of the liquid. It is reported that the mean residence time of the liquid decreased with increase in liquid space velocity. However, the mean residence time was a stronger function of space velocity for the upflow mode of operation. 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.

Chander et al. [5] also studied the variation of liquid mean residence time with gas velocity at constant liquid hourly space velocity. The study showed that when a larger size of diluent was used, the mean residence time increased with gas/liquid ratio for the upflow mode of operation. 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.

2.2.2.2 Other Moments of the RTD

It is very common to compare RTDs by using their moments instead of trying to compare their entire distributions.

The second moment commonly used is taken about the mean and is called the variance, or square of the standard deviation. It is defined by

$$\sigma^2 = \int (t - t_m)^2 E(t) dt \qquad (18)$$

alternatively,

$$\sigma^2 = \frac{\int_0^\infty (t - t_m)^2 C(t) dt}{\int_0^\infty C(t) dt} = \frac{\sum (t_i - t_m)^2 C_i \Delta t_i}{\sum C_i \Delta t_i} \qquad (19)$$

The magnitude of this moment is an indication of the 'spread' of the distribution as it passes the vessel exit and has units of (time)^2; the greater the value of this moment, the greater a distribution's spread.

It is particularly useful for matching experimental curves to one of a family of theoretical curves. ***Figure 2.3*** illustrates these terms.

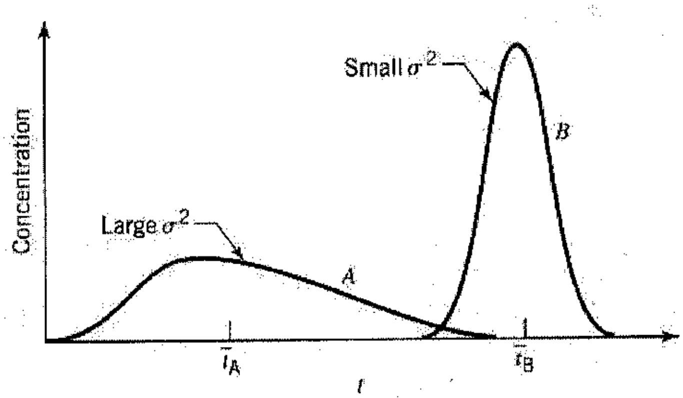

Figure 2.3 Variance for matching theoretical curves.

The third moment is also taken about the mean and is related to the skewness. The skewness is defined by

$$s^3 = \frac{1}{\sigma^{3/2}} \int_0^\infty (t - t_m)^3 E(t) dt \qquad (20)$$

The magnitude of this moment measures the extent that a distribution is skewed in one direction or another in reference to the mean.

Rigorously, for complete description of a distribution, all moments must be determined. Practically, these three (t_m, σ^2, s^3) are usually sufficient for a reasonable characterization of an RTD.

2.3 RTD ANALYSIS ON AXIAL DISPERSION AND STAGNANT ZONE VOLUME

2.3.1 Axial Dispersion

Suppose an ideal pulse of tracer is introduced into the fluid entering a reactor. The pulse spreads as it passes through the vessel, and to characterize the spreading according to dispersion model (***Figure 2.4***), it is assumed a diffusion-like process superimposed on plug flow. This is called dispersion or longitudinal dispersion to distinguish it from molecular diffusion. The dispersion coefficient D (m^2/s) represents this spreading process. Thus

- large D means rapid spreading of the tracer curve
- small D means slow spreading
- $D = 0$ means no spreading, hence plug flow

Also, $\left(\frac{D}{uL}\right)$ is the dimensionless group characterizing the spread in the whole vessel.

D or D/uL is evaluated by recording the shape of the tracer curve as it passes the exit of the vessel. In particular, t_m (mean time of passage, or when the curve passes by the exit) and σ^2 (variance, or a measure of the spread of the curve) are measured.

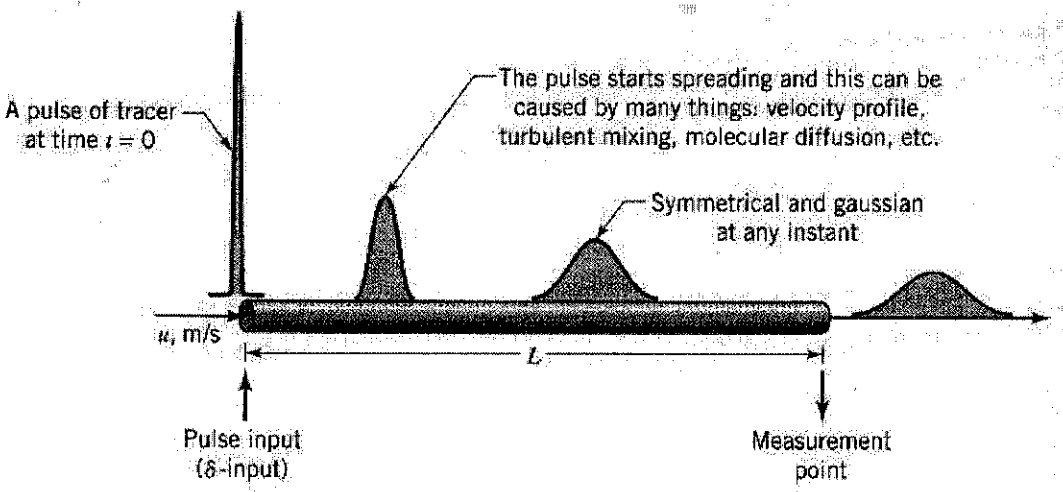

Figure 2.4 The spreading of tracer according to the dispersion model.

These measures, t_m and σ^2, which are earlier mentioned, are directly linked by theory to D and D/uL.

Consider plug flow of a fluid, on top of which is superimposed some degree of backmixing, the magnitude of which is independent of position within the vessel. This condition implies that there exist no stagnant pockets and no gross bypassing or short-circuiting of fluid in the vessel. This is called the dispersed plug flow model, or simply the dispersion model. ***Figure 2.5*** shows the conditions visualized. Note that with varying intensities of turbulence or intermixing the predictions of this model should range from plug flow at one extreme to mixed flow at the other. As a result, the reactor volume for this model will lie between those calculated for plug and mixed flow.

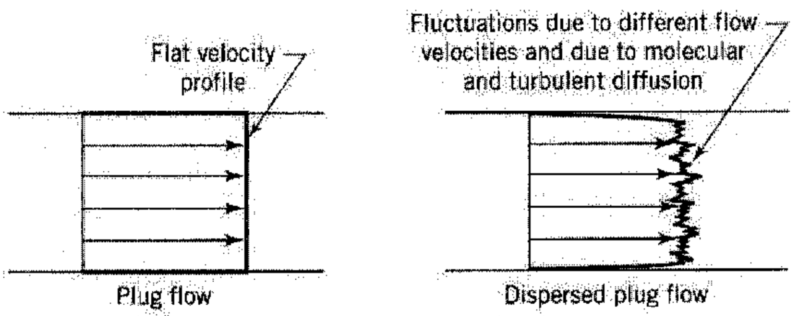

Figure 2.5 Representation of the dispersion (dispersed plug flow) model.

Since the mixing process involves a shuffling or redistribution of material either by slippage or eddies, and since this is repeated many, many times during the flow of fluid through the vessel, these disturbances are considered to be statistical in nature, somewhat as in molecular diffusion. For molecular diffusion in the x-direction, the governing differential equation is given by Fick's law;

$$\frac{\partial C}{\partial t} = \mathrm{D} \quad \frac{\partial^2 C}{\partial x^2} \tag{21}$$

where D , the coefficient of molecular diffusion, is a parameter which uniquely characterizes the process. In an analogous manner, it can be considered that all the contributions to intermixing of fluid flowing in the x-direction to be described by a similar form of expression, or

$$\frac{\partial C}{\partial t} = D\frac{\partial^2 C}{\partial x^2} \tag{22}$$

where the parameter D, which is called the longitudinal or axial dispersion coefficient, uniquely characterizes the degree of backmixing during flow. The terms longitudinal and axial are used because it is to distinguish mixing in the direction of flow from mixing in the lateral or radial direction, which is not the primary concern. These two quantities may be quite different in magnitude. For example, in streamline flow of fluids through pipes, axial mixing is mainly due to fluid velocity gradients, whereas radial mixing is due to molecular diffusion alone.

In dimensionless form where $z = (ut + x)/L$ and $\theta = t/t_m = tu/L$, the basic differential equation representing this dispersion model becomes

$$\frac{\partial C}{\partial \theta} = \left(\frac{D}{uL}\right)\frac{\partial^2 C}{\partial z^2} - \frac{\partial C}{\partial z} \tag{23}$$

where the dimensionless group $\left(\frac{D}{uL}\right)$, called the vessel dispersion number, is the parameter that measures the extent of axial dispersion. Thus

$\left(\frac{D}{uL}\right) \rightarrow 0$ negligible dispersion, hence plug flow

$\left(\frac{D}{uL}\right) \rightarrow \infty$ large dispersion, hence mixed flow

The dispersion model usually represents quite satisfactory flow that deviates not too greatly from plug flow, thus real packed bed and tubes (not long ones if flow is streamline).

The bed Peclet number (henceforth only Peclet number) of liquid is the reciprocal of the dispersion number, $\left(\frac{D}{uL}\right)$, i.e.

$$\mathrm{Pe} = \frac{1}{D/uL} \tag{24}$$

which the dispersion number is also defined by

$$\left(\frac{D}{uL}\right) = \frac{\sigma_\theta^2}{2} \tag{25}$$

and

$$\sigma_\theta^2 = \left(\frac{u}{L}\right)^2 \sigma^2 \tag{26}$$

where σ^2 is the variance of the E curve.

Stiegel and Shah [6] have reported that during the upflow mode of operation, the bed Peclet number of liquid increased with decrease in particle size. However, Peclet number value increased with the increase in liquid space velocity indicating the reduction of backmizing with higher liquid flow rate. Chander et al. [5] indicated that the use of fine size of diluent also reduced the dependency of Peclet number on space velocity. Similarly, Stiegel and Shah [6] have found an increasing trend in Peclet number with increasing liquid velocity for the upflow mode of operation when a larger size of particles was used.

Chander et al. [5] also studied the variation of Peclet number with gas/liquid ratio. The study proved that 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. Cassanello et al. [9] have also reported that the gas velocity affects the axial dispersion coefficient for upflow operation.

2.3.2 Stagnant Zone Volume

The hydrodynamics in a reactor is an important factor that influences the efficiency of a reactor. The existence of stagnant zone greatly reduces the efficiency and performance of the reactor. RTD analysis provides a good indication on the presence of stagnant zone as well as the flow pattern through the reactor.

The deviation of the ideal flow can be determined by obtaining a complete velocity distribution profile measured through the reactor. However, the approach is rather impractical. So, there is a need to study the age of distribution of fluid exiting the reactor by the stimulus response technique. The concept of the technique is to introduce a tracer at the inlet or some point within the reactor. Then, at some point along the reactor or at an exit, the tracer is collected to measure the concentration subsequent time interval. In order to illustrate the RTD of the actual flow, the stimulus-response experiment can be conducted with an appropriate choice of tracer. The packed bed reactor presumably behaves as a plug flow reactor. However, deviation from the ideal plug flow can occur due to short-circuiting, channeling or an existence of dead zone (***Figure 2.6***). Arrangement of packing and adequate distribution of liquid can disrupt the ideal behavior of plug flow due to the channeling of liquid.

Sata et al. [10] considered an ideal plug flow behavior in which the tracer should emerge in the exit until $T_i = T_d$ at the same concentration of the entrance. The mean residence time, t_m is calculated from RTD analysis, previously mentioned.

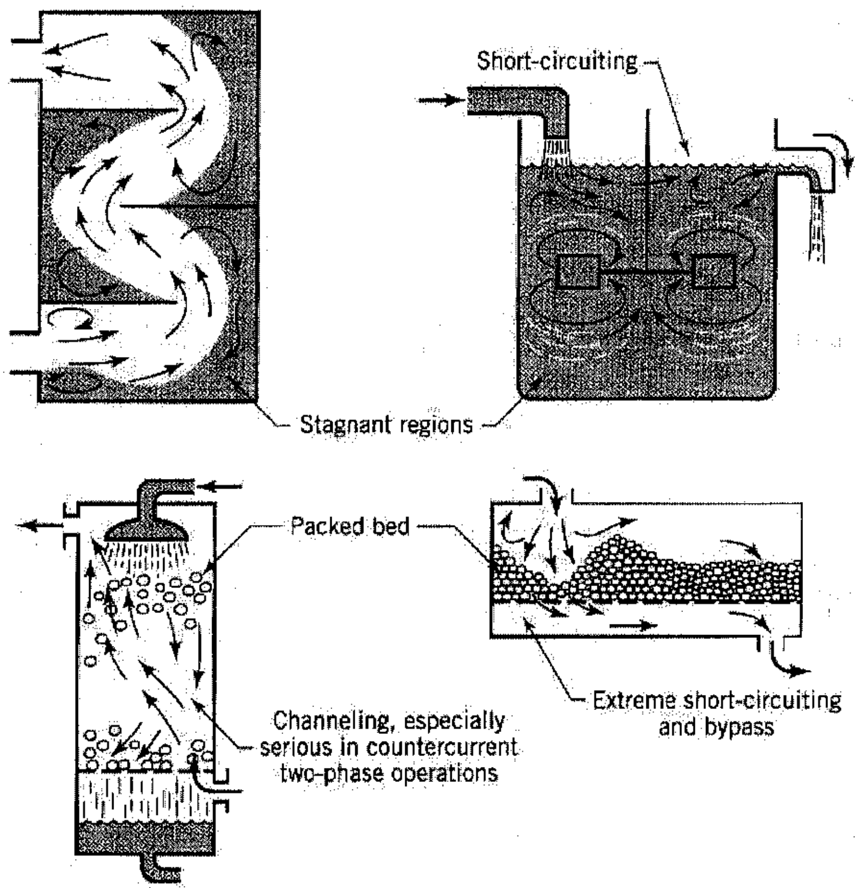

Figure 2.6 Non ideal flow patterns which may exist in process equipment

The mean residence time can be determined by the equation:

$$t_m = \frac{\sum(t_i C_i \Delta t_i)}{\sum C_i \Delta t_i} \qquad (27)$$

The mean residence time can also be defined as the reactor volume-volumetric flow rate ratio:

$$T_d = \frac{V}{Q} \tag{28}$$

The stagnant zone volume can be estimated based on the ratio of actual, t_m and theoretical HRT, T_d:

$$V_{stagnant} = V\left(1 - \frac{t_m}{T_d}\right) \tag{29}$$

Study by Sata et al. [10] have reported that if the tracer peak emerged earlier than the predicted theoretical HRT, this meant that the effective volume of the reactor is reduced due to a form of channeling in the packing media, which will give low t_m/T_d ratio.

It is also observed that the peak of higher flow rate will appear first, which indicated the phenomenon of channeling. Another deviation is the tailing effect of the tracer toward longer time, which indicated recycling effect and tracer accumulation in the reactor.

CHAPTER 3
METHODOLOGY / PROJECT WORK

3.1 GENERAL EXPERIMENTAL EQUIPMENT

The SOLTEQ RTD studies in Tubular Reactor (Model BP 112) is utilized for this experiment and is designed for experiment on residence time distribution (RTD) in a packed bed reactor. Process diagram of this experimental instrument is illustrated in ***Figure 3.1***. The unit consisted of a reactor, a system for feeding controlled and measured amounts of gas and liquid, tracer injection, and conductivity measurement instrument for detecting the concentration of the tracer. The liquid phase is de-ionized water and the gas phase is air.

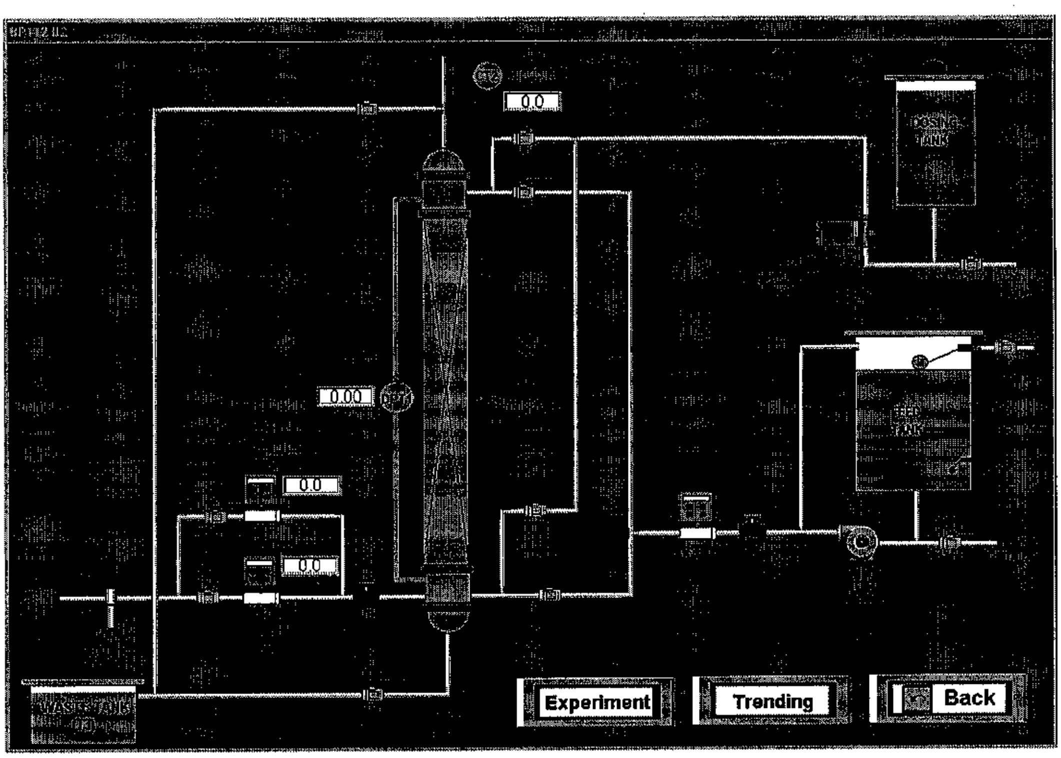

Figure 3.1 Process Diagram for RTD Studied in Tubular Reactor (BP 112).

a) *Reactor*

A column made of borosilicate glass packed with 8 x 8 mm Raschig rings. Column OD: 100 mm; ID: 82 mm; Height: 1500 mm. Top and bottom caps made of stainless steel fitted with appropriate inlet and outlet ports. A differential pressure tapping is also provided on both caps.

b) *Feed Tank*

20-L stainless steel cylindrical tank, equipped with circulation pump. The tank is fitted with a level switch to protect the pump from dry run.

c) *Dosing Tank*

20-L stainless steel cylindrical tank equipped with a metering pump.

d) *Waste Tank*

50-L rectangular tank made of stainless steel.

e) *Instrumentations*

Air Flowmeter;

Fluid	:	Compressed air (0.34 MPa)
Range	:	0 to 50 LPM
Output	:	0 to 5 VDC
Display	:	LCD digital display

Liquid Flowmeter;

Fluid	:	De-ionized water
Range	:	0 to 5 LPM
Output	:	0 to 5 VDC
Display	:	LCD digital display

Conductivity Meter;

Sensor Range	:	0 to 200 mS/cm
Sensors	:	CT2 (Co-current Upflow)
Output	:	4 to 20 mA
Display	:	Conductivity controller with digital display

g) *Data Acquisition System (DAS)*

DAS is consisted of a personal computer, ADC modules and instrumentations for measuring the process parameters. A flowmeter with 0 to 5 VDC output signal is supplied for feed flowrate measurement. Conductivity sensors with controller are provided for monitoring the tracer concentration in each reactor. All analog signals from the sensors are then converted by the ADC modules into digital signals before being sent to the personal computer for display and manipulation.

3.2 EXPERIMENTAL GAS AND LIQUID FLOW RATES

The gas and liquid flow rates used in the experiment are maintained close to those typically used for testing different hydroprocessing catalysts in a bench-scale unit. The level of these flow rates depend on the type of catalyst to be tested.

For atmospheric gas oil hydrotreating catalyst, gas/liquid ratio (hydrogen/oil ratio) of approximately 150 – 250 (v/v) is required. The values of gas/liquid ratio are used in this experiment, which is summarized in ***Table 3.1***.

Gas/Liquid Ratio (LPM/LPM)	**Gas Flow Rates (LPM)**	**Liquid Flow Rates (LPM)**
150	7.5	0.05
	15.0	0.10
	22.5	0.15
200	10.0	0.05
	20.0	0.10
	30.0	0.15
250	12.5	0.05
	25.0	0.10
	37.5	0.15

Table 3.1 Experiment gas and liquid flow rates with specified gas/liquid ratio.

3.3 PROCEDURE IDENTIFICATION

3.3.1 General Start-up Procedure

1. A quick inspection is performed to ensure that the equipment is in proper working condition.
2. 10 liter of 0.2M NaCl solution is prepared for tracer solution.
3. The system is flushed with de-ionized water until no traces of salt is detected.
4. The equipment is ready to be run.

3.3.2 Determination of Experimental Pressure Drop and Liquid Holdup

1. General start-up procedure is performed.
2. Valves are set appropriately for co-current upflow mode.
3. Gas and liquid flow rates are adjusted to obtain desired gas/liquid ratio.
4. Conductivity reading is observed and stabilized at low value.
5. The system is maintained approximately 30 minutes to attain steady state, which the flow rates did not change after 30 minutes of operation.
6. The pressure drop is obtained from the differential pressure reading at control panel or via the DAS.
7. The liquid holdup is determined as follows;
   a. After attaining steady state, the gas and liquid flows are stopped simultaneously.
   b. The total free liquid in the reactor is drained in a liquid collector and measured.
   c. The liquid holdup is expressed as (Volume of Liquid) / (Volume of Column).
8. Gas and liquid flow rates are varied to obtain required gas/liquid ratio as per commercial reactors.
9. Data are recorded in Appendices.

3.3.3 Residence Time Distribution (RTD) Analysis

1. The general start-up procedure is performed.
2. The valves are set appropriately for co-current upflow mode.
3. Gas and liquid flow rates are adjusted to obtain desired gas/liquid ratio.
4. Conductivity reading is observed and stabilized at low value.
5. Tracer is introduced in the system for two minutes. The conductivity reading is recorded at 1 minute interval, until the reading is constant.
6. Experiment is stopped by closing the inlet and outlet valves simultaneously.
7. From the concentration-time data from experiment, E curve and F curve are constructed.
8. The values of hydrodynamics parameters are determined by utilizing the RTD analysis. Mean Residence Time (t_m), Variance (σ^2), Skewness (s^3), Axial Dispersion (expressed by Peclet number of liquid, Pe) and Stagnant Zone Volume ($V_{stagnant}$) are calculated as outlined in Chapter 2, and summarized in Chapter 4.
9. The experiment is repeated with different gas and liquid flow rates.
10. Data are recorded in Appendices.

CHAPTER 4
RESULTS AND DISCUSSION

4.1 EFFECT OF GAS/LIQUID RATIO ON PRESSURE DROP AND OPERATING LIQUID HOLPUP

4.1.1 Effect of Gas/Liquid Ratio on Pressure Drop

Pressure drop analysis across the reactor is done by investigating the effect of gas/liquid ratio, as well as comparing the experimental value with Ergun [1] correlation. The pressure drop throughout the experiment is recorded and the result is shown in ***Figure 4.1***. The pressure drop increases with increasing gas/liquid ratio, both experimentally and theoretically.

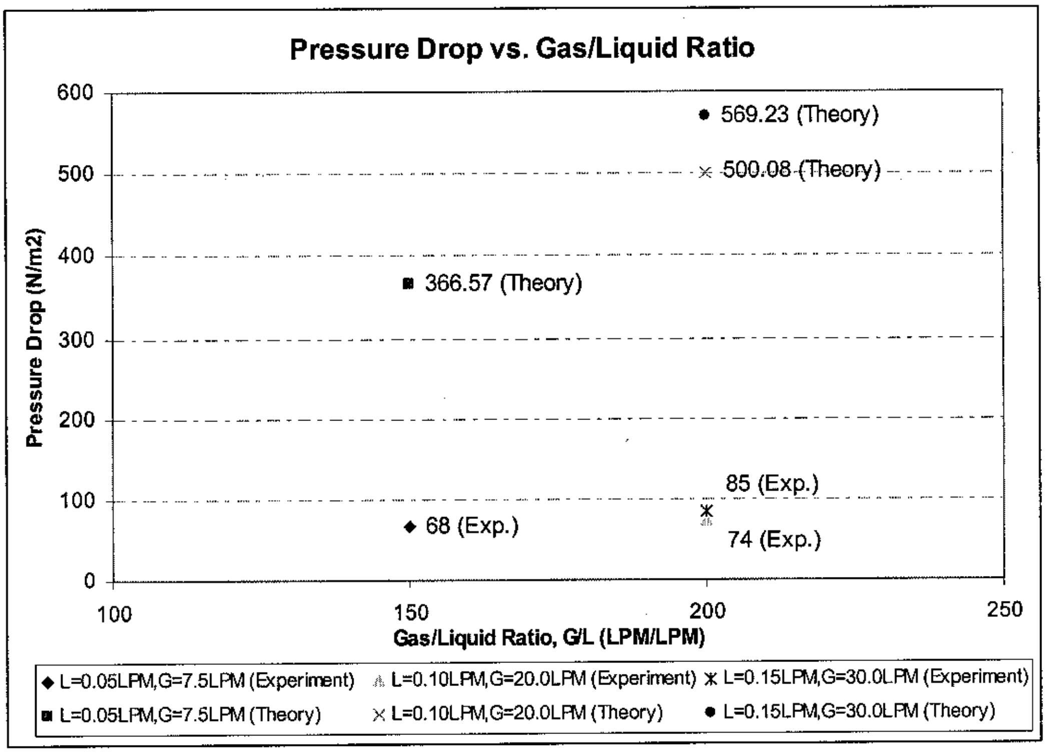

Figure 4.1 Effect of gas/liquid ratio on pressure drop

Observing the values of pressure drop, the pressure increases with increasing variation of gas and liquid flow rates, for both experimental value and Ergun [1] correlation.

The results obtained in this experiment is in strong agreement with results obtained by Varma et al. [3], who shows that the pressure drop increased with the gas and liquid flow rates in all the regimes in upflow mode of operation.

However, there is a large deviation between experimental values and the values calculated from Ergun [1] correlation. This may be because of Ergun [1] correlation consider only gas phase. Ergun [1] theory revealed a decrease in pressure drop accounted by the gas flow rate. Gas may help to push liquid thru column thus reduces the pressure drop. If considering the accuracy of theories, Varma et al. [3] accounted both liquid and gas flow rates on pressure drop, unlike in Ergun [1] equation.

4.1.2 Effect of Gas/Liquid Ratio on Operating Liquid Holdup

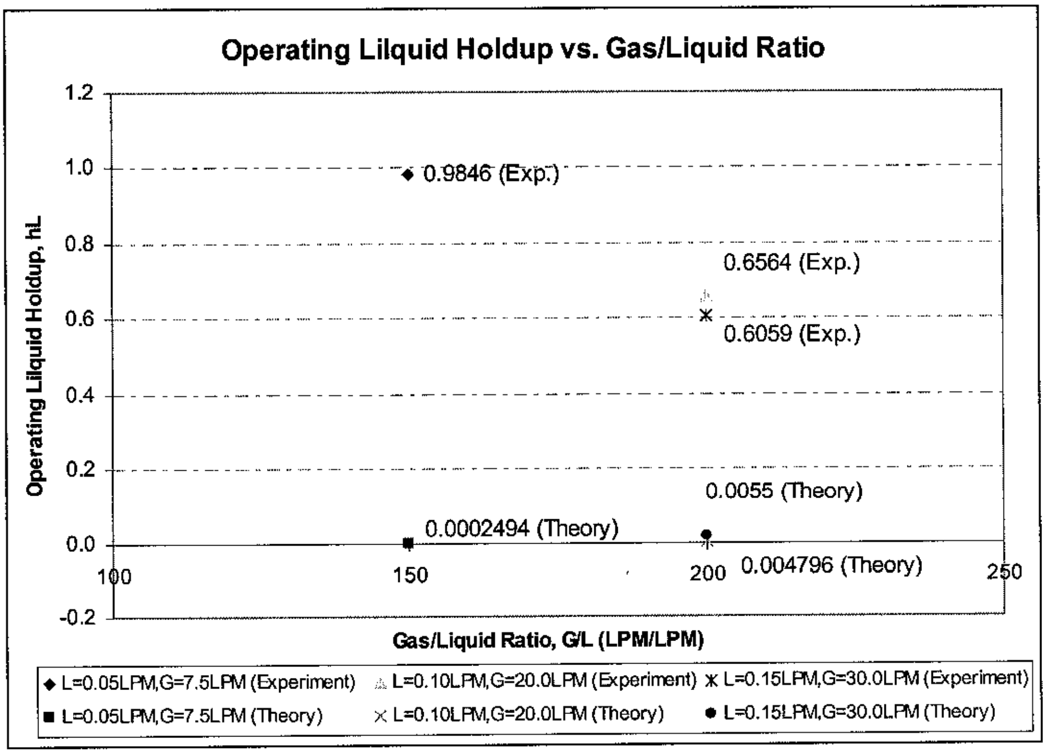

Figure 4.2 Effect of gas/liquid ratio on operating liquid holdup

Figure 4.2 shows that the experimental value of operating liquid holdup decreases with increasing gas/liquid ratio and liquid flow rates. This is similar to the findings by Chander et al. [5].

However, Buchanan [4] correlation does not agree with the experimental results. It is observed that operating liquid holdup is increasing with increasing liquid flow rates and gas/liquid ratio. This can be due to the basis of this correlation which only emphasized on only liquid phase in equation (2), and not both gas and liquid phases. Also, it might be because the correlation has used different condition against the condition being used in this experiment during that time. This can also be true if the correlation agrees with Chander et al. [5], which proved that the effect of gas flow rate on liquid holdup in upflow mode could be removed, if the catalyst bed was diluted with a smaller size of diluent.

From the experimental results, operating the packed bed reactor at higher gas/liquid ratio (200) with higher gas and liquid flow rates (G = 30.0 LPM, L = 0.15 LPM) is more desirable.

4.2 EFFECT OF GAS/LIQUID RATIO ON MOMENTS OF RTD

For this experiment, RTD experiment with pulse input is used. An amount of tracer (NaCl) is injected in one shot into the feedstream entering the reactor in as short time as possible. The outlet conductivity is then measured as a function of time. The effluent concentration-time curve is referred as *C* curve in RTD analysis. However, the consideration is more to the *E* curve and the three moments of RTD.

4.2.1 *E* Curve

Figure 4.3 and ***Figure 4.4*** showed that the *E* curve exhibited deviation of mixed flow behavior for both gas/liquid ratio of 150 and 200. Further experiments should be done on different gas/liquid ratio also gas and liquid flow rates to justify the trend.

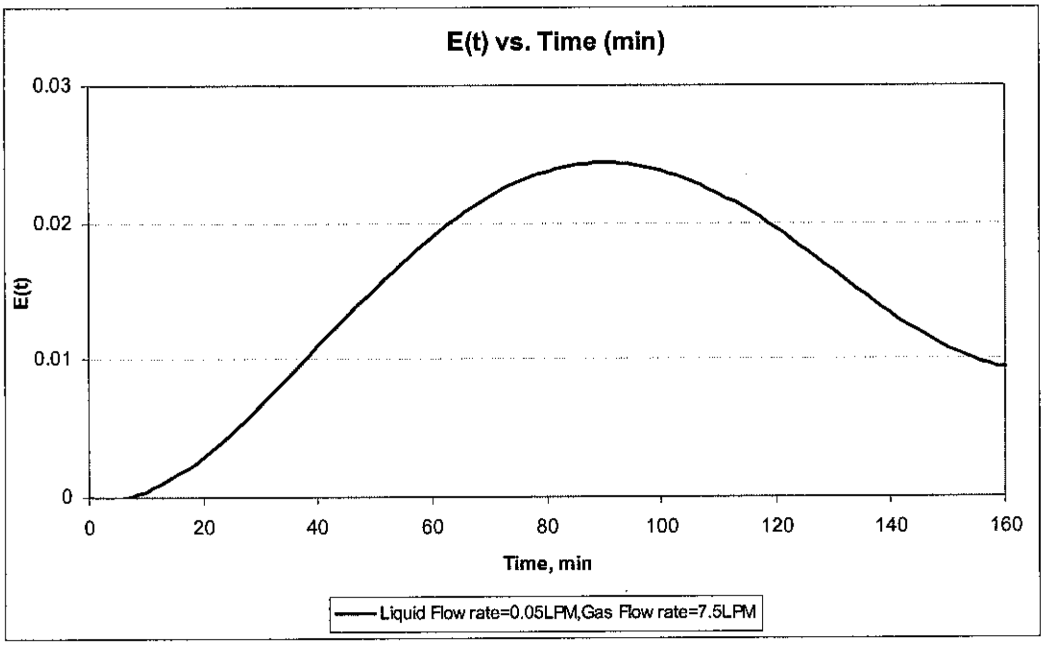

Figure 4.3 *E* curves for gas/liquid ratio of 150

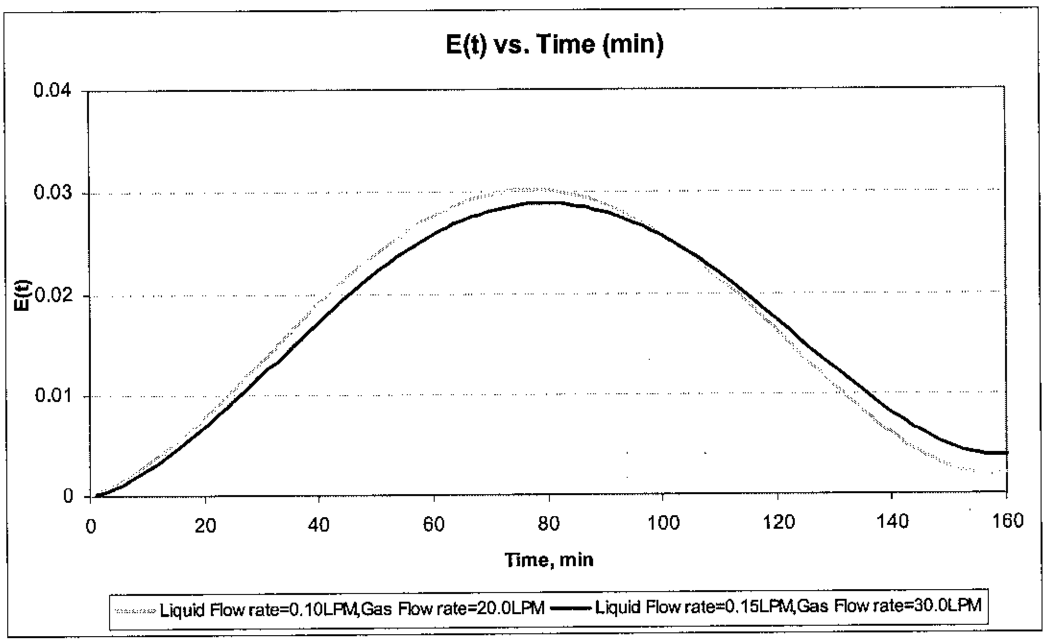

Figure 4.4 *E* curves for gas/liquid ratio of 200

4.2.2 Three Moments Analysis of RTD

4.2.2.1 First Moment Analysis: Mean Residence Time, t_m

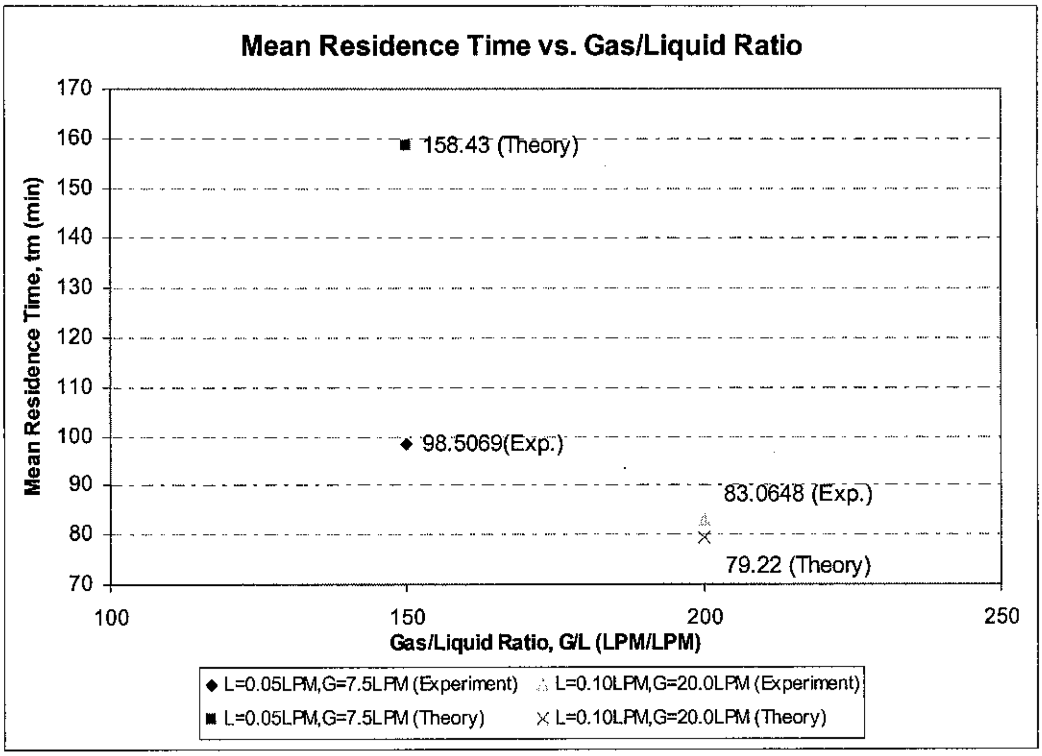

Figure 4.5 Effect of gas/liquid ratio on mean residence time of liquid

Experimentally, it is observed that there is a considerable decrease in mean residence time as the gas/liquid ratio is increased, illustrated in ***Figure 4.5***. This result does not agree with Chander et al. [5].

Chander et al. [5] reported that the mean residence time increases with gas/liquid ratio for upflow mode of operation. The increasing gas flow rate maybe due to induced circulatory motion of the liquid inside the bed. Therefore the liquid would spend more time in the reactor. The longer mean residence time may lead to better kinetics as liquid spends longer time in the reactor allowing for better catalyst utilization.

Further analysis, the mean residence time of the reactor obtained from experiment show the same trend as calculated from theory. However, there is a large difference between experiment and theoretical value of mean residence time for gas/liquid ratio of 150. Mellon, N. et al. [13] reported that this suggests that there is a stagnant zone inside the reactor. In theory, in the presence of heat transfer, these stagnant zones may develop hot spots inside the reactor. Depending on the extent of the availability of stagnant zones, this may, in fact cause an operational problem later.

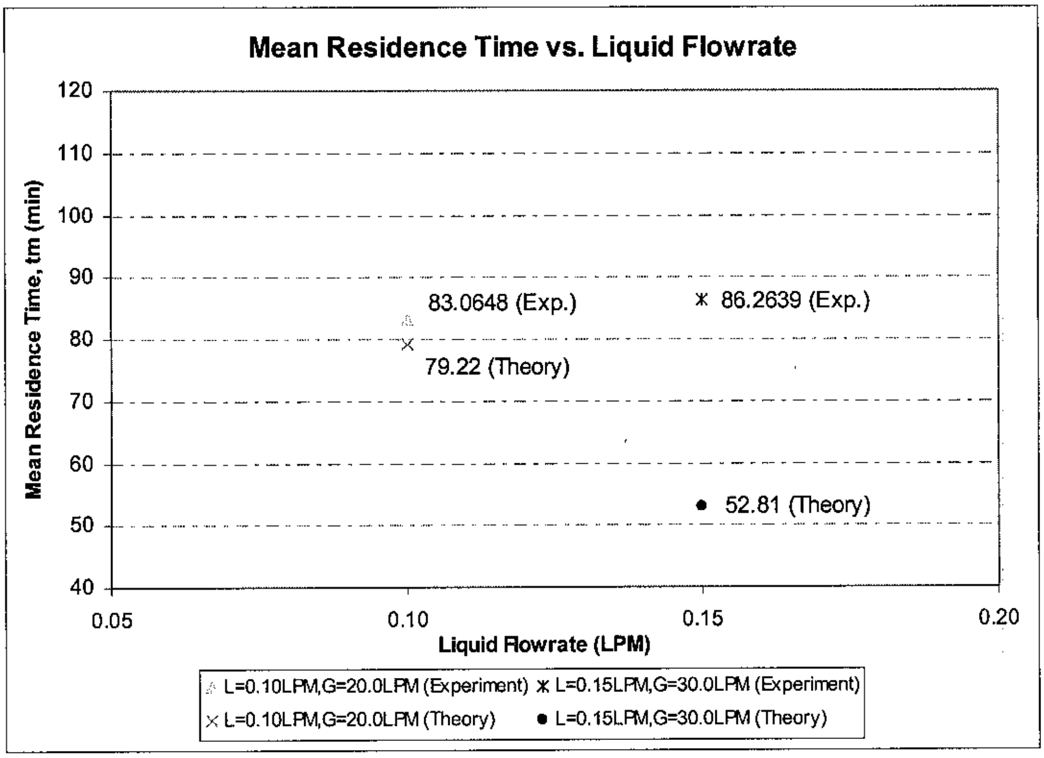

Figure 4.6 Effect of liquid flowrate on mean residence time at gas/liquid ratio of 200

Experimentally, the effect of liquid flowrate on mean residence time of the liquid was studied at a constant gas/liquid ratio of 200 and the result is shown in ***Figure 4.6***. Experimentally, the mean residence time of the liquid increase with increase in liquid flowrate for upflow mode of operation. The result however, opposing with the value calculated theoretically and with studied by Chander et al. [5].

The analysis shows that the mean residence time for this reaction is high. Furthermore, the peak of the *E* curves (***Figure 4.3*** and ***Figure 4.4***) occurs earlier than the mean residence time (early mixing). This is an indication of the possibility of stagnant zone in the reactor. It also could be due to excessive liquid holdup inside the catalyst bed. According to Chander et al. [5], the higher mean residence time would 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. In this case, the emerging peak of E curves which is earlier than the mean residence time suggests that there would be insufficient contact with catalyst and also, undesirable thermal reaction would take place during this period.

Further experiments with different gas and liquid flow rates need to be performed to clarify the significant effect of gas and liquid flow rates and effect of gas/liquid ratio.

4.2.3.2 Second Moment Analysis: Variance, σ^2

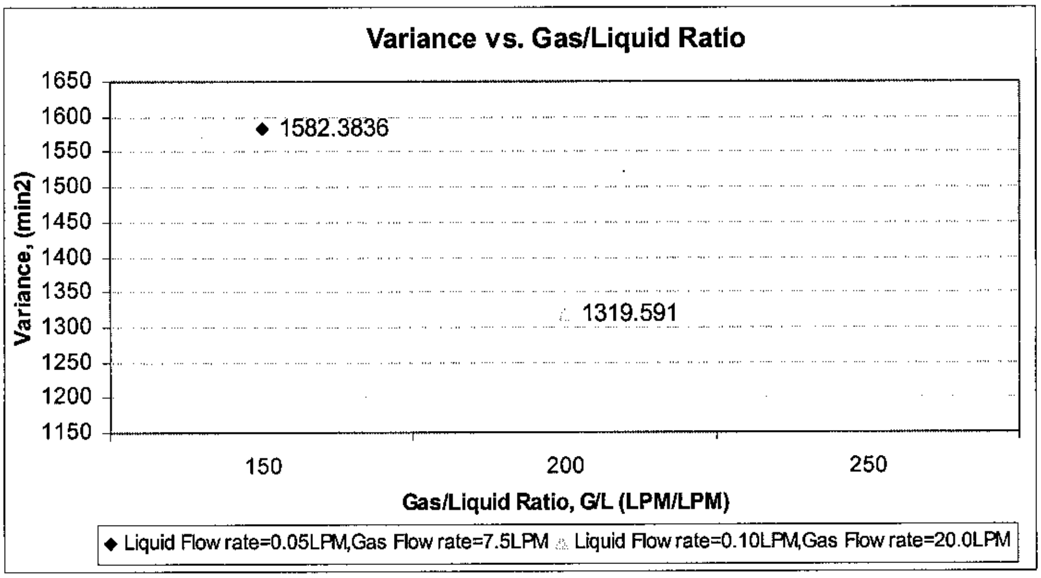

Figure 4.7 Effect of gas/liquid ratio on variance

Experimentally, from ***Figure 4.7***, the variance decreased as the gas/liquid ratio increases.

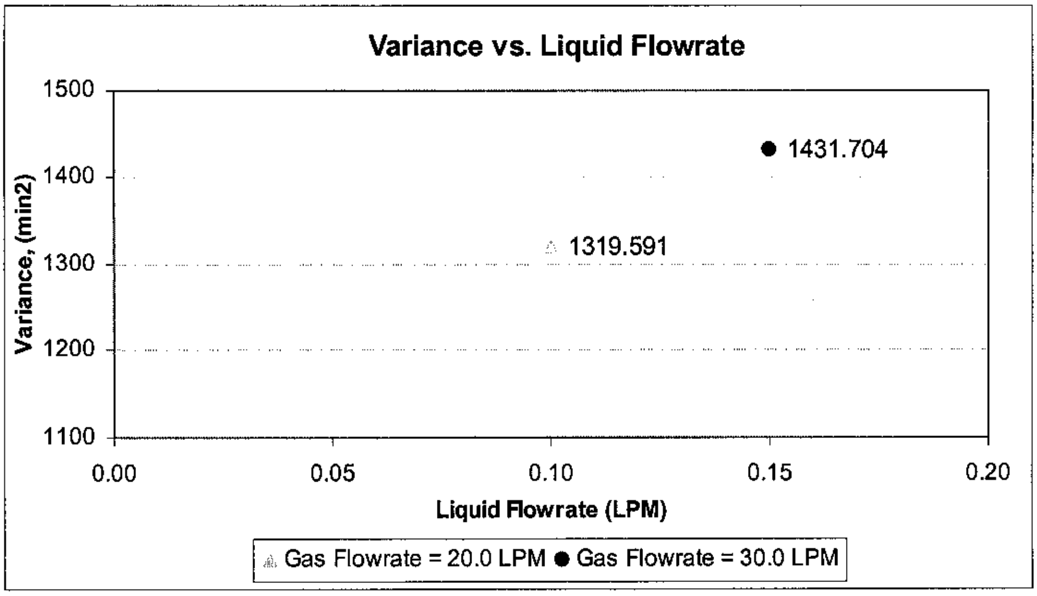

Figure 4.8 Effect of liquid flowrate on variance at gas/liquid ratio of 200

And from ***Figure 4.8***, the variance decrease with decreasing gas and liquid flowrate.

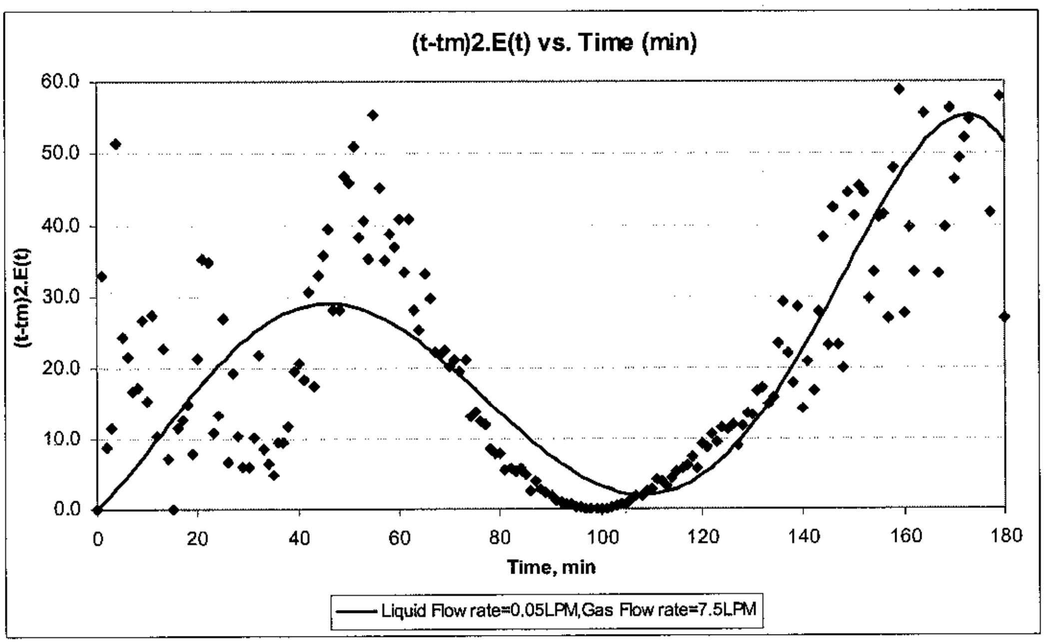

Figure 4.9 Variance Curve for Gas/Liquid Ratio of 150 (7.5 LPM / 0.05 LPM)

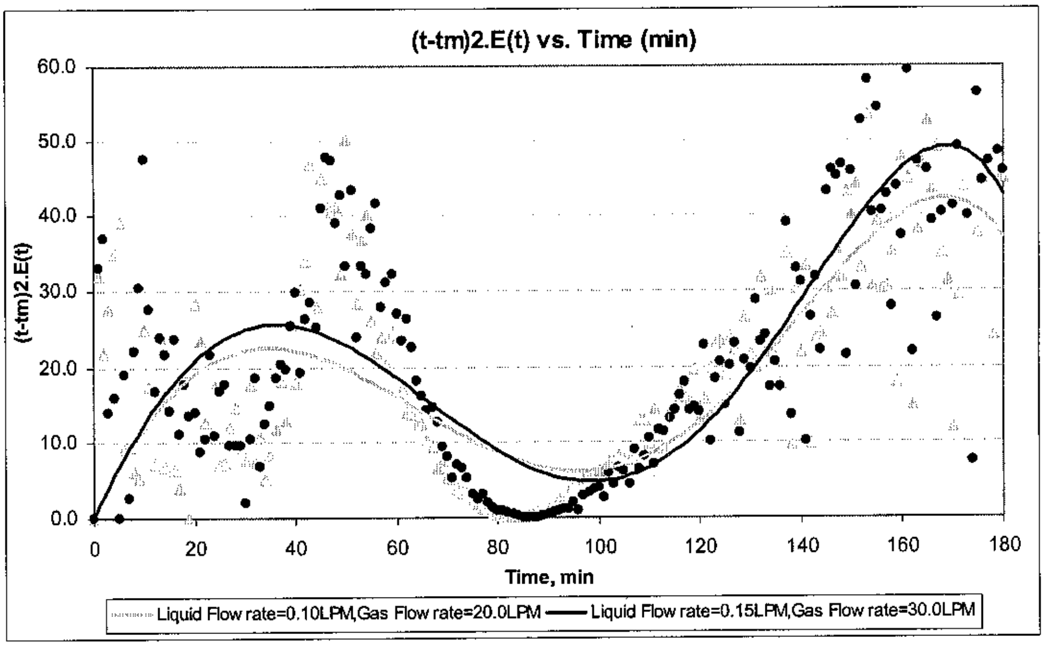

Figure 4.10 Variance Curve for Gas/Liquid Ratio of 200 (20.0 LPM / 0.10 LPM and 30.0 LPM / 0.15 LPM)

These results are also consistent with the *E* curve in ***Figure 4.3*** and ***Figure 4.4***; the *E* curve for liquid flow rate of 0.10 LPM has higher peak compare to liquid flowrates of 0.15 LPM, which indicates smaller variance, and vice versa. Thus, it is proved that the smaller the variance, the smaller the distribution's spread.

4.2.3.3 Third Moment Analysis: Skewness, s^3

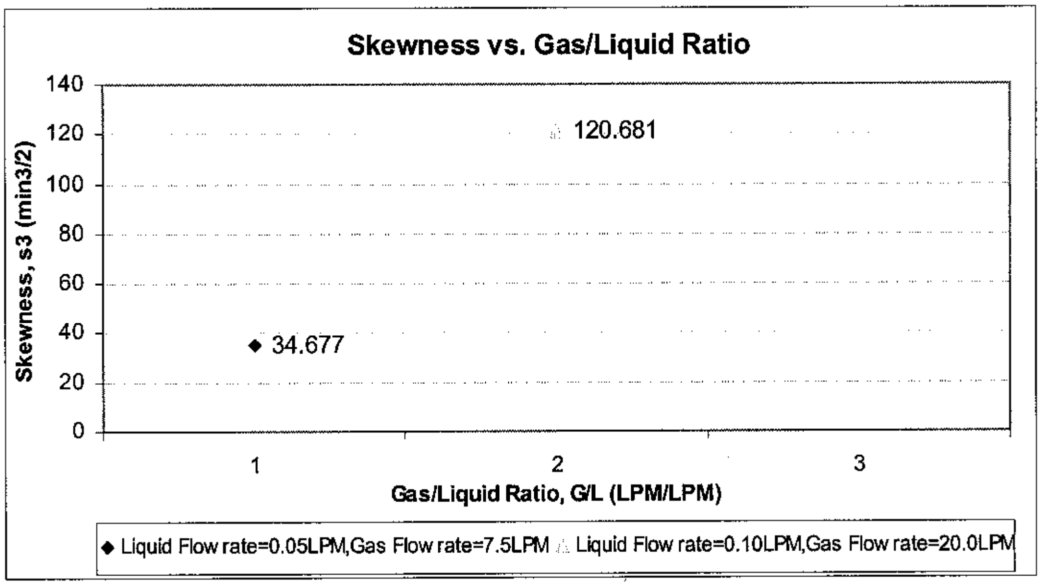

Figure 4.11 Effect of gas/liquid ratio on skewness

From ***Figure 4.11***, it is observed that the skewness increases as the gas/liquid ratio increases. In order to achieve plug flow behavior, it is desirable to have smaller value of skewness.

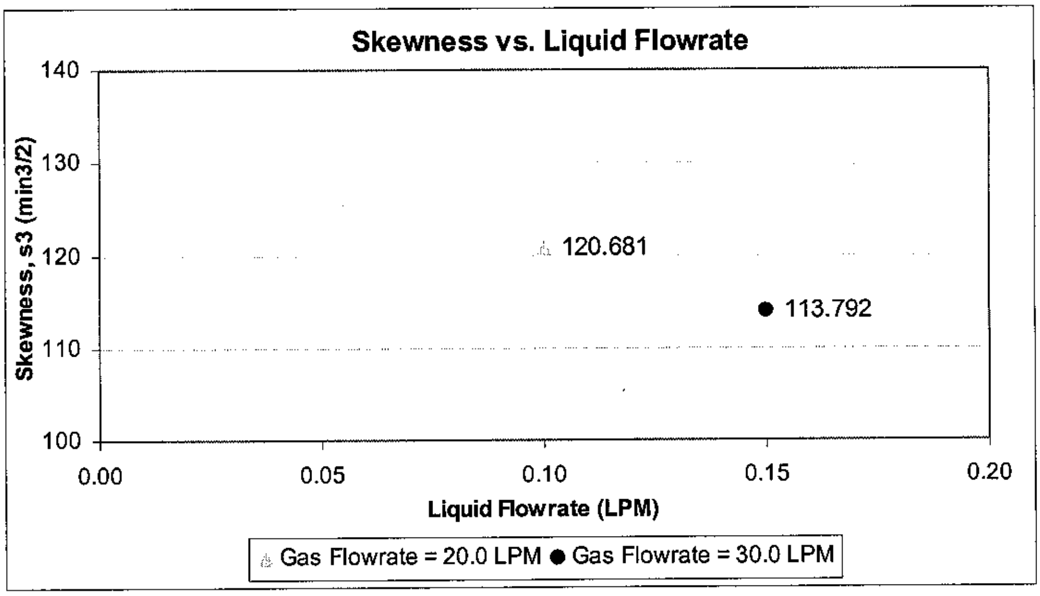

Figure 4.12 Effect of liquid flowrate on skewness at gas/liquid ratio of 200

And from ***Figure 4.12***, the skewness decreased as liquid flowrate increased.

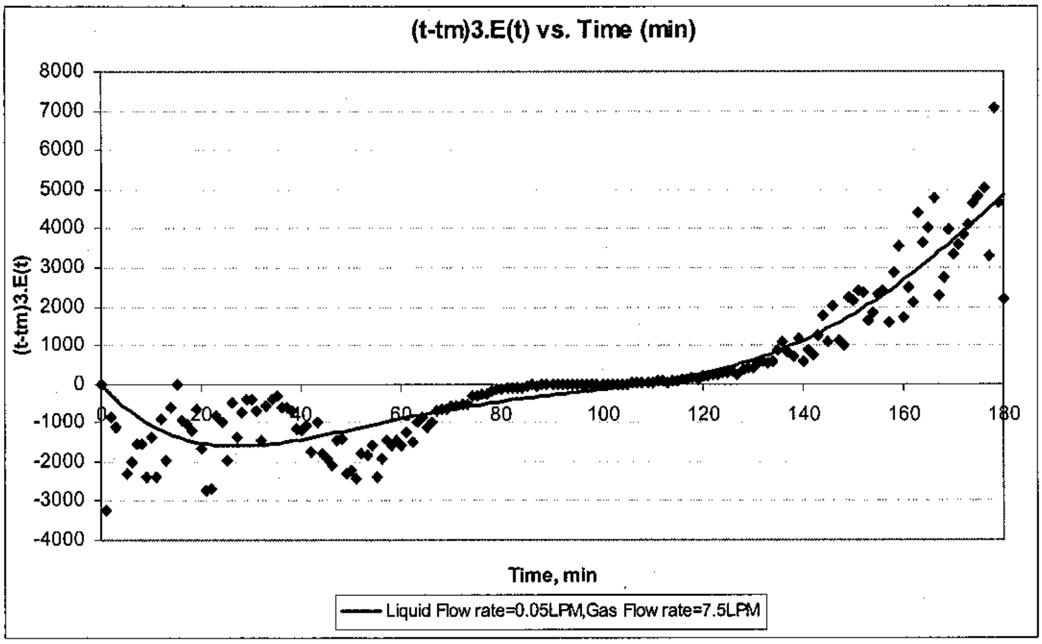

Figure 4.13 Skewness for Gas/Liquid Ratio of 150 (7.5 LPM / 0.05 LPM)

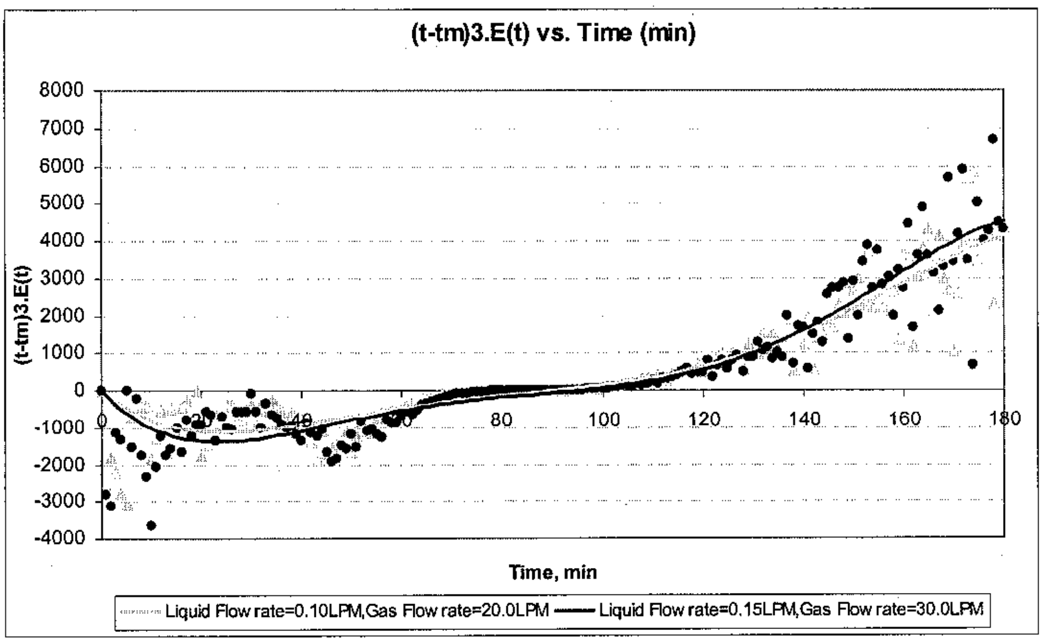

Figure 4.14 Skewness for Gas/Liquid Ratio of 200 (20.0 LPM / 0.10 LPM and 30.0 LPM / 0.15 LPM)

Again, these results are reflected by the E curve in ***Figure 4.3*** and ***Figure 4.4***; the E curve for liquid flowrate of 0.15 LPM is not skewed far from the reference of mean compared to liquid flowrate of 0.10 LPM (skewed left), which indicates lowest skewness, and vice versa. The lower the skewness, the less skewed the distribution is, from its mean.

4.3 EFFECT OF GAS/LIQUID RATIO ON AXIAL DISPERSION AND STAGNANT ZONE VOLUME BY RTD ANALYSIS

4.3.1 Effect of Gas/Liquid Ratio on Axial Dispersion

Another analysis in RTD study is the degree (intensity) of liquid-phase axial dispersion. This axial dispersion is conveniently expressed as Peclet number in the analysis. Lower value of Peclet number indicates higher degree of dispersion in the system.

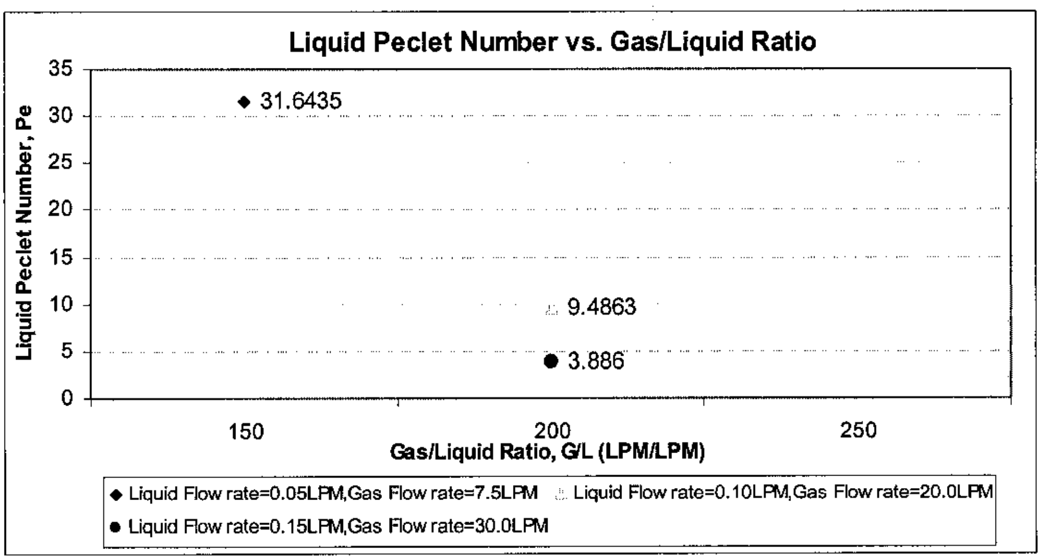

Figure 4.15 Effect of gas/liquid ratio on liquid Peclet number

Experimentally, the Peclet number relatively decreases with increasing gas/liquid ratio and gas liquid flowrates as illustrated in ***Figure 4.15***. These results agrees with studies done by Chander et al. [5], who also reported that the Peclet number is a very strong decreasing function of gas/liquid ratio for upflow mode. This might due to the increase of circulatory motion of liquid causing backmixing with increasing gas flow rate. Study by Cassanello et al. [9] also agrees that the gas velocity affects the Peclet number value for upflow mode of operation. Thus, the reduction of backmixing can be achieved at low gas/liquid ratio (150) with lower gas and liquid flow rates (G = 7.5 LPM, L = 0.05 LPM).

4.3.2 Effect of Gas/Liquid Ratio on Stagnant Zone Volume

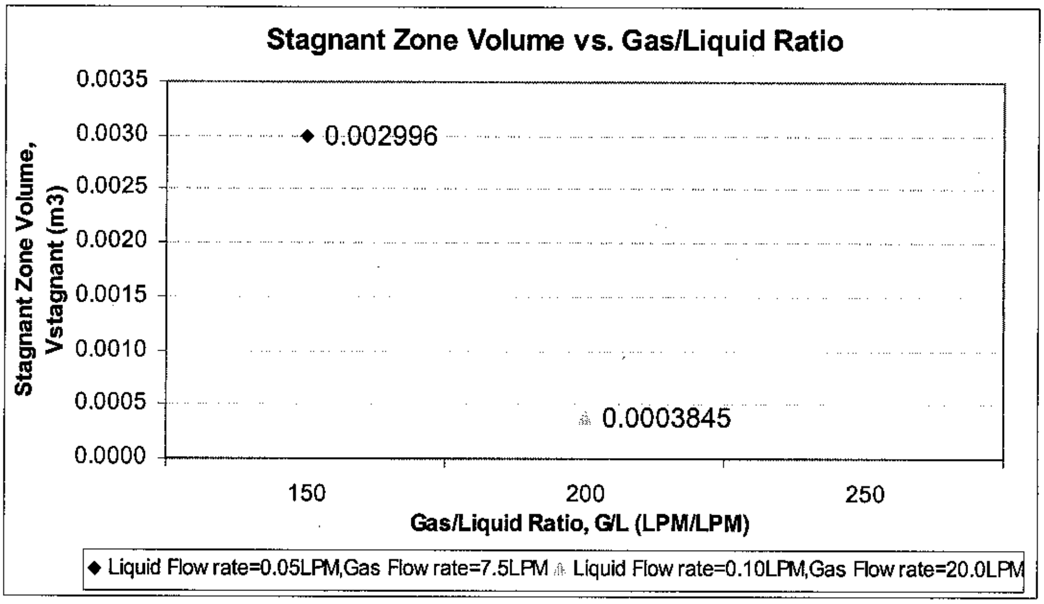

Figure 4.16 Effect of gas/liquid ratio on stagnant zone volume

From experiment, it is observed that stagnant zone volume can be reduced if the gas/liquid ratio is increased, as illustrated by ***Figure 4.16***.

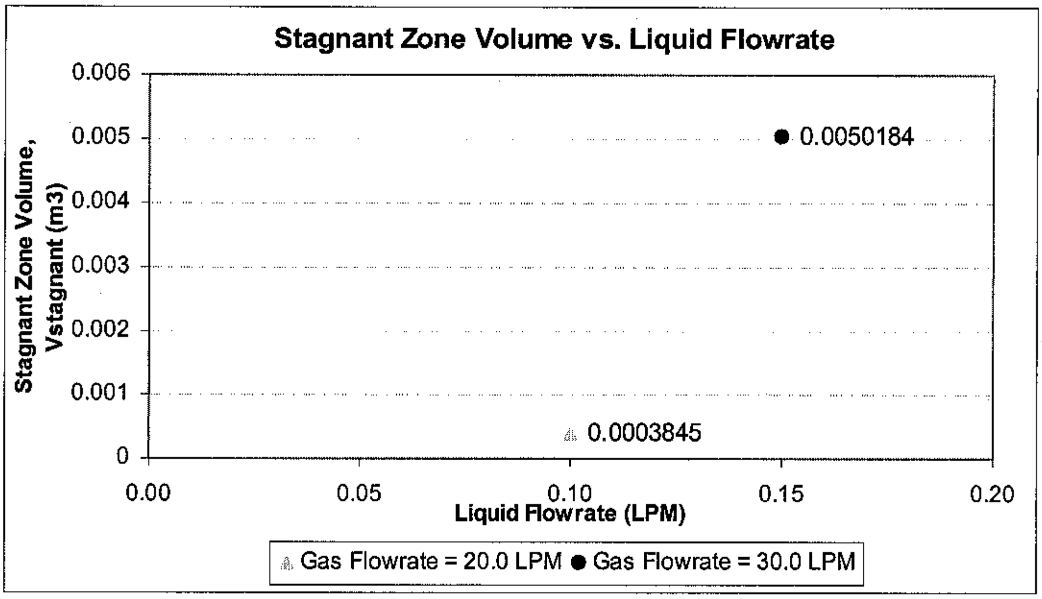

Figure 4.17 Effect of liquid flowrate on stagnant zone volume at gas/liquid ratio of 200

Also, the stagnant zone volume reduces with decreasing value of liquid flow rates as shown in ***Figure 4.17***. This is with agreement with the theoretical HRT proposed by Sata et al. [10] in equation (28) and (29), which is only influenced by liquid flow rate.

At low liquid flow rate, the mean residence time is lower than the predicted theoretical HRT, which means that the effective volume of the reactor is reduced due to a form of channeling in the packing media, which will give low t_m / T_d ratio. Another deviation is the tailing effect of the tracer towards longer time, which indicated recycling effect and tracer accumulation in the reactor.

CHAPTER 5
CONCLUSIONS AND RECOMMENDATIONS

5.1 CONCLUSIONS

Experimentally, pressure drop, ΔP can be reduced with decreasing gas/liquid ratio. Thus, gas and liquid flow rates should be maintain low as to reduce pressure drop across the reactor. Also, the theories introduced in this project are considered parallel with the experimental results.

Operating liquid holdup, H_o is decreasing with increasing gas/liquid ratio in experiment. Also, the liquid holdup is decreased with increasing liquid flow rates. Thus, for a desirable process, liquid holdup must be minimized, which can be achieved at high gas/liquid ratio (200) with high gas and liquid flowrate (G = 30.0 LPM, L = 0.15 LPM). However, the theory neglected the effect of gas flow rate and thus constant for same liquid flow rate. Thus, the theory used in this project is considered inappropriate and required the need of other theories in future.

Mean residence time, t_m is increased with decreasing gas/liquid ratio. This is desirable for better utilization of catalyst. However, too high mean residence time would also result undesirable thermal reaction. Mean residence time is also increased with increasing liquid flowrate. Thus, the desirable operation can be done at low gas/liquid ratio (150) with lower gas and liquid flowrate (G = 7.5LPM, L = 0.05 LPM).

Variance, σ^2 is decreased when gas/liquid ratio is increase. Also, the variance is decreased with decreasing of liquid flow rates. Variance reflects the spread of distribution. The more the distribution spread, the higher the value of variance, which also results more towards mixed flow behavior. Small variance is desired for a fixed bed reactor to behave more towards plug flow. High gas/liquid ratio (200) but with lower gas and liquid flow rates (G = 20.0 LPM, L = 0.10 LPM), will effect in decreasing of variance.

Skewness, s^3 measured the extent that the distribution is skewed in one direction or another in reference to its mean. From the experiment, skewness can be decreased as the gas/liquid ratio is decreased and with decreasing of liquid flow rates. Skewness is undesirable because the higher the value of skewness, the further the distribution is skewed from the mean, which will also deviate from plug flow characteristics. Thus, lower value of skewness is preferred to operate the packed bed reactor towards plug flow behavior. Low gas/liquid ratio (150) with lower gas and liquid flowrate (G = 7.5LPM, L = 0.05 LPM) give lower skewness.

Axial dispersion of liquid, expressed by Peclet number, Pe can be increased with decreasing gas/liquid ratio and decreasing of liquid flowrate. The increase of circulatory motion of liquid can cause backmixing which is due to the increasing gas flowrate. Thus, it is proved that backmixing can be reduced if this packed bed reactor is operated at low gas/liquid ratio (150) with lower gas and liquid flow rates (G = 7.5 LPM, L = 0.05 LPM).

Stagnant zone volume, $V_{stagnant}$ can be reduced with higher gas/liquid ratio, as well as higher liquid flowrate. This is mainly because the mean residence time is lower than the predicted HRT. Thus, the effect of non ideal reactor, which is caused by channeling, dead zones, or short-circuiting, can be reduced at higher gas/liquid ratio (200) flow rates with higher gas and liquid flowrate (G = 30.0 LPM, L = 0.15 LPM).

As overall conclusion, in order to be an ideal reactor, certain requirements must be fulfilled. From the experiment;

1. As gas/liquid ratio and gas liquid flow rates increases, the **operating liquid holdup** as well as the **stagnant zone volume** decreases. (G/L = 200, G = 30.0 LPM, L = 0.15 LPM)

2. Lower gas/liquid ratio with lower gas and liquid flow rates, effect in decreasing of **pressure drop** and in increasing of **Peclet number (axial dispersion)**. (G/L = 150, G = 7.5 LPM, L = 0.05 LPM)
3. High gas/liquid ratio (200) but with lower gas and liquid flow rates (G = 20.0 LPM, L = 0.10 LPM), will effect in decreasing of **mean residence time** and **variance**.
4. Low gas/liquid ratio (150) and low gas and liquid flow rates (G = 7.5 LPM, L = 0.05 LPM), give lower **skewness**.

Further experiments with different gas and liquid flow rates are to be performed to clarify the significant effect of gas and liquid flow rates and effect of gas/liquid ratio.

5.2 RECOMMENDATIONS

Recommendations outlined here is based on studies that can be done or extended for future development of RTD analysis, or rather the investigation of hydrodynamic characteristics of fixed bed reactor or packed bed reactor.

5.2.1 Comparative Study between Co-current Upflow and Downflow Mode of Operation

Trickle bed reactors with co-current downflow of gas and liquid have found wide application in oil industries for the hydroprocessing of petroleum fractions. However, there are several drawbacks for scaling up and scaling down of commercial reactor. Owing to the differences in the hydrodynamics, a small-scale reactor cannot be treated as an exact replica of a commercial unit.

Two approaches can be recommended to overcome the drawbacks during the testing of commercial catalysts in small reactor. 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. However, these two approaches differ in their basic nature and performance; in the upflow mode of operation, liquid is in continuous phase and gas remains in the dispersed phase, whereas the situation is reversed in downflow operation. The upflow mode of operation, though it ensures almost complete wetting of catalyst, suffers from serious drawback of non-ideal flow of liquid and formation of a stagnant zone inside the catalyst bed.

Thus, there is a need to compare the hydrodynamic behavior of a fixed bed reactor in the upflow and downflow modes of operation. There are several theories regarding this comparative study. For example, the downflow mode provided much lower residence time to the liquid as compared to that for the upflow mode of operation, probably owing to channeling of liquid in the former case. On the other hand, the upflow mode of operation gave a much higher liquid holdup as compared to the downflow mode. It is predicted that higher liquid hourly space velocity reduced the channeling of liquid in downflow operation and reduced backmixing in upflow mode. Also, a higher liquid holdup at higher liquid hourly space velocity may be obtained in the downflow mode of operation.

5.2.1 Effect of Diluent Size

When the catalyst was loaded with smaller size of diluent, the values of mean residence time, Peclet number and liquid holdup is expected to increase for the downflow mode. As a result of this, the hydrodynamics behavior for both upflow and downflow modes of operation can be improved.

Since the project study did not discuss the effect of diluent size, future study can be made on investigating the change in the behavior of the upflow mode on using a smaller size of particle as diluent in the catalyst bed. The use of smaller size of diluent can increase the value of Peclet number and moderate the excessive liquid holdup, 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 can also be eliminated for the smaller size of diluent. The values of mean residence time, Peclet number and liquid holdup are predicted nearly the same for the two modes of operation. Thus, the use of a smaller size of diluent could remove the drawbacks of both upflow fixed bed and trickle bed reactors, which will provide suitable tools for generating reliable data for scale-up and scale-down activities.

The use of a smaller size of diluent can also decreased the porosity of the bed, which in turn reduced the excessive mean residence time of liquid in the upflow mode of operation. This could help in the reduction of undesirable non-catalytic reaction in the upflow mode.

5.2.2 Effect of Packing Types

The hydrodynamics comparison between different packing types can be done with same variation of gas and liquid flow rates. The effect of packing type can be investigated by RTD analysis to study the hydrodynamics of the reactor at different types of packing.

The comparison can also made by the estimation of stagnant zone volume or by investigation of the effect of packing types on nature of E curve. The E curve can be a direct indication whether there is non-ideal behavior in the reactor. It is seen if any peak of E curve emerges first compared to others, which can be attributed by to the channeling.

Moreover, different types of packing introduced different types of distribution in the reactor. As an example, Raschig rings and Pall rings have different shapes and sizes, which in turn affect the distribution. Also, the different packing types can be studied against pressure drop, which also an effect of the different distribution in the reactor.

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