Investigations of Hydrodynamics of Fixed Bed Reactor: Counter-Current Flow

by

Norafneeza Norazahar

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

JANUARY 2006

Universiti Teknologi PETRONAS
Bandar Seri Iskandar
31750 Tronoh
Perak Darul Ridzuan

CERTIFICATION OF APPROVAL

Investigations of Hydrodynamics of Fixed Bed Reactor: Counter-Current Flow

by

Norafneeza Norazahar

A project dissertation submitted to the
Chemical Engineering Programme
Universiti Teknologi PETRONAS
in partial fulfillment of the requirement for the
BACHELOR OF ENGINEERING (Hons)
(CHEMICAL ENGINEERING)

Approved by,

Ms. Nurhayati Mellon
Project Supervisor

UNIVERSITI TEKNOLOGI PETRONAS
TRONOH, PERAK

January 2006

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.

NORAFNEEZA NORAZAHAR

ABSTRACT

The residence time distribution (RTD) of a pilot plant scale packed bed reactor was studied in the present study using air – water system to investigate the effect various gas/liquid volumetric ratio on the residence time distribution. The ranges of air – water ratio is chosen to simulate the hydrogen/oil ratio of typical bench scale hydro-processing units, noting the difference that might arise due to the different physical properties of the liquid. The experiments were conducted in counter-current mode and the effects of gas/liquid velocities on the hydrodynamics of the reactor were investigated. This study is done using the tracer method – specifically the pulse input method. The packed bed reactor used in this study is packed with 8 x 8 mm Raschig rings. Results show that the increasing gas/liquid ratio increases the mean residence time. However, operating liquid hold-up and the value of Peclet number, which gives an indication to the degree of axial dispersion in the system, increase. Second moment analysis of the reactor, namely the variance of the system, shows a reduction in value with decreasing gas/liquid ratio suggesting wider liquid distribution through the system at low gas flow rates. 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, Ms. Nurhayati 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 Reaction Engineering laboratory technicians, who have helped in various ways in providing sufficient assistance and supports.

To supportive colleagues and families, the author mostly appreciates the motivations and advices.

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.

TABLE OF CONTENTS

LIST OF FIGURES AND TABLE

LIST OF FIGURES

LIST OF TABLE

ABBREVIATIONS AND NOMENCLATURES

LPM	Liter per minute
LHSV	Liquid Hour Space Velocity
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 Study

Multiphase systems are very often applied in biotechnology (immobilized microbes or enzymes on a solid support used as a catalyst for microbic or enzyme fermentations) and chemical processes where the reactants and catalyst are present in different phases (gas–liquid and solid–catalyst). The reaction rate of such processes usually depends on the specific contact area between the phases, such as fluid and solid, as well as on the mass transfer rate from one phase to the other. External mixing in a reactor of different and specific construction and usage could generate phase dispersion.

The flow of phases in continuous processes is usually not ideal and in some cases, such deviation from the ideal flow pattern (plug flow or ideal mixing) might be substantial. Such deviations from the ideal type of fluid flow are usually channeling, recycle flow or stagnant zone formation inside the reactor or some other type of equipment used in the chemical process industry (distillation columns, absorbers, adsorbers, heat exchanger). When such non-ideal behavior of fluid flow occurs, an important decrease of process unit productivity may be detected.

Knowing the effect of non-ideal flow is essential in the case of scale-up design. In many cases, it is enough to have information about the residence time distribution of a fluid in a reactor determined by stimulus-response methods with the appropriate tracer. The diminishing effects of non-ideal flow in the design of industrial units might lead to enormous errors in their exploitation.

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 operations and reduced risk of hazards.

Packed bed reactors have found wide application in industrial practice and are considered more efficient than the other types of contactors, because of increased efficiency brought about by the use of packings. Packed bed reactors are vertical towers equipped with suitable liquid and gas distributors and have grids for the support of packing materials, which help to vary the local velocity of the continuous phase and restrict the re-circulation and back-mixing in the phases considered.

A vast amount of research has been carried out in three phase packed bed reactors. All liquid residence time distribution (RTD) measurements were conducted at room temperature and near atmospheric pressure. The tracer technique was chosen for determination of the needed liquid flow parameters. The change in electrical conductivity of a potassium chloride (KCl) tracer was simultaneously measured downstream/upstream of the injection point at both the top and the bottom of the bed using computer-monitored conductivity sensors. A brief tracer pulse was syringe-injected into the liquid inlet pipe. The signals of the sensors were amplified and transmitted to a computer by a data acquisition system (DAS). A residence time distribution model to describe the liquid flow in a fixed bed reactor packed with porous particles and operated under both partially and fully wetted conditions was proposed based on a simple representation of the liquid flow structure. This model assumes that the external liquid stream is divided into two zones: a dynamic zone in which the liquid flows through the bed as a piston flow with axial dispersion and a stagnant zone in direct contact with the dynamic zone and the solid particle. The dynamic evolution of the tracer concentration in the particles was described in terms of diffusion phenomena [1].

1.2 Problem Statement

Hydrodynamics play an important part on the performance of packed bed reactor. The knowledge of the real flow pattern and the active volume of packed bed reactor are critical information to scale-up and modeling purposes. The performance of packed bed reactor relies on hydrodynamics, such as mixing, mass and heat transfer, other than its microbial kinetics and physical phenomena. As for mixing, there exist two categories, which are micromixing and macromixing. The effective method of examining the macromixing behavior of reactor is Residence Time Distribution (RTD) analysis.

1.3 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.

1.4 Objectives

The objectives of investigations of hydrodynamic in a packed bed reactor of counter-current flow are:

1. To investigate the effect of liquid and gas flow rate on pressure drop and operating liquid hold-up.
2. To characterize residence time distributions (RTD) of reactor by three moments pertaining mean residence time, variance, and skewness.
3. To perform the residence time study (tracer study) in order to analyse the effect of gas/liquid volumetric ratio on the hydrodynamics of the reactor.

1.5 Project Scope

The scope of research, which was successfully accomplished:

1. Understand the principle and characteristics of hydrodynamics.
2. Investigate the effect of gas and liquid flow rates towards residence time distribution (RTD) and further investigation on three moments of RTD

3. Determine the liquid hold-up in a packed bed reactor at gas and liquid flow rates.
4. Familiarize the concept of back-mixing and stagnant zone volume.
5. Understand the concept of axial dispersion on gas/liquid ratio to observe the mixing behavior of reactor.
6. Determine pressure drop in packed bed reactor at different gas and liquid flow rates.

CHAPTER 2
LITERATURE REVIEW

2.1 Packed Bed

Packed bed columns for gas-liquid contacting are used extensively in chemical process, especially in separation and reaction. Usually the columns are filled with a randomly oriented packing material, but for an increasing number of applications the packing is very carefully positioned in the column. The packed column is characteristically operated with counter flow of the phases.

The packed column is a simple device as shown in ***Figure 2.1***. A typical column consists of a cylindrical shell, containing a support plate for the packing material and a liquid distributing device designed, to provide good liquid distribution on the system. Devices may be added to the packed bed to provide redistribution of liquid that might channel down the wall or otherwise become maldistributed. Several beds, each with liquid distributor and support device, may be used within the same column shell. For example, a distillation column with rectifying and stripping zones requires a minimum of two beds.

Packed columns are usually specified when plate devices would not be feasible because of undesirable fluid characteristic or some special design requirement. Conditions favouring packed columns are:

- For columns less than 0.6-m (2.0-ft) diameter, packings are usually cheaper than plates unless alloy-metal packings are required.
- Acids and many other corrosive materials can be handled in packed column because construction can be of ceramic, carbon, or other resistant materials.
- Packings often exhibit desirable efficiency-pressure-drop characteristic for critical vacuum distillations.

- Liquids, tending to foam, may be handled more readily in packed columns because of the relatively low degree of liquid agitation by the gas.
- Hold-up of liquid can be quite low in packed columns, an advantage when the liquid is thermally sensitive.

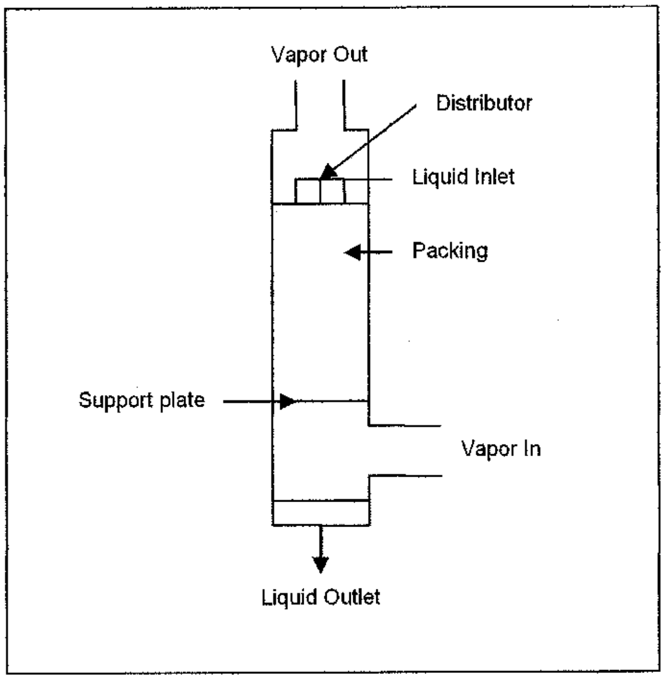

Figure 2.1 Schematic of packed column

2.2 Packed Column Hydraulics

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 [2] which is included in the Perry's Handbook. The Ergun equation is:

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

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 [3] 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. [4] 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 counter-current downflow can be assumed equivalent in this experiment, as in Ergun [2] 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, in unit of mbar, 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.

2.3 Operating Liquid Hold-up

The dynamic liquid holdup at different flow rates of the phases, refers to Iluita et al [1], is the volume of the liquid drained after the inflow was stopped after reaching steady state. The static liquid holdup was obtained from the difference between the external void volume of the bed and that of the liquid drained after the reactor had been totally flooded with liquid.

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. Besides, 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.

2.4 Residence Time Distribution (RTD)

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 constant stirred type reactor (CSTR) is thoroughly mixed and possesses a far different kind of RTD than the plug-flow reactor. The RTD of a reactor yields distinctive clues to the type of mixing occurring within it and is one of most informative characterizations of the reactor.

2.5 Measurement of 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;

2.5.1 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 2.2***. The effluent concentration-time curve is referred to as the C curve in RTD analysis. The injection of a tracer pulse shall be analysed 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

$$\Delta N = C(t)v\Delta t \qquad (2)$$

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

$$\frac{\Delta N}{N_o} = \frac{vC(t)}{N_o}\Delta t \qquad (3)$$

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

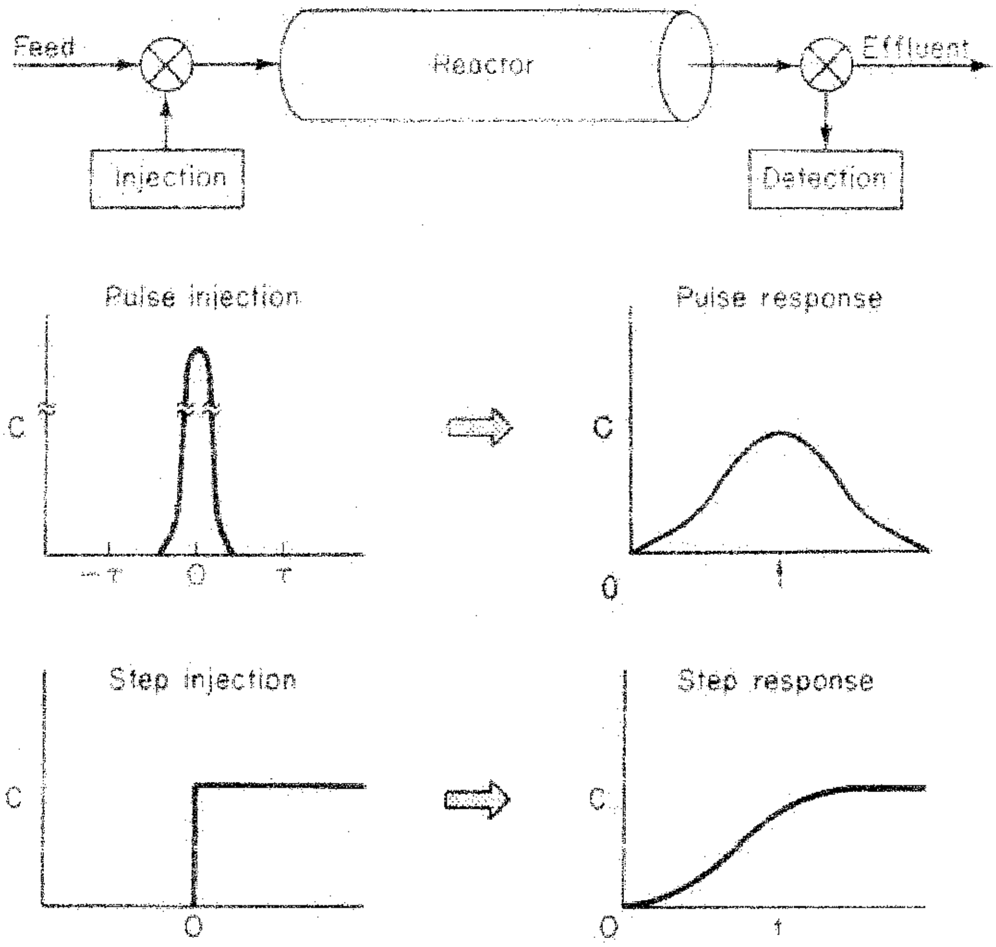

Figure 2.2 RTD measurements

For pulse injection, it is defined

$$E(t)=\frac{vC(t)}{N_o} \tag{4}$$

So that,

$$\frac{\Delta N}{N_o}=E(t)\Delta t \tag{5}$$

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 (2) in differential form yields,

$$dN = vC(t)dt \tag{6}$$

And then integrating,

$$N_o = \int_0^{\infty} vC(t)dt \tag{7}$$

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

$$E(t) = \frac{C(t)}{\int_0^{\infty} C(t)dt} \tag{8}$$

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:

$$\boxed{\begin{array}{l}\text{Fraction of material leaving the}\\ \text{reactor that has resided in the reactor}\\ \text{for time between } t_1 \text{ and } t_2\end{array}} = \int_{t_1}^{t_2} E(t)dt \tag{9}$$

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,

$$\int_0^{\infty} E(t)dt = 1 \tag{10}$$

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 (8), 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.

2.5.2 Step Tracer

The meaning of the RTD curve is previously discussed, now a more general relationship between a time varying tracer injection and the corresponding concentration in the effluent will be formulated. It should be stated without development that the output concentration from a vessel is related to the input concentration by the convolution integral.

$$C_{out}(t) = \int_0^t C_{in}(t-t')E(t')dt' \qquad (11)$$

The inlet concentration most often takes the form of either perfect pulse input (Dirac delta function), imperfect pulse injection, refer to ***Figure 2.2***, or step input.

Step input in the tracer concentration will be analysed for a system with a constant volumetric flow rate. Consider a constant rate of tracer addition to a feed which is initiated at time $t = 0$. Before this time, no tracer was added to the feed. Stated symbolically,

$$C_o(t) = 0 \qquad t < 0$$

$$C_o(t) = \text{constant} \qquad t \geq 0$$

The concentration of the tracer in the feed to the reactor is kept at this level until the concentration in the effluent is indistinguishable from that in the feed;

the test may then be discontinued. A typical outlet concentration curve for this type of input is shown in ***Figure 2.3***.

Because the inlet concentration is constant with time, C_o, the integral is taken outside the integral sign, i.e.,

$$C_{out} = C_o \int_0^t E(t')dt'$$

Dividing by C_o

$$\left[\frac{C_{out}}{C_o}\right]_{step} = \int_0^t E(t')dt' = F(t) \qquad (12)$$

This expression is differentiated to obtain the RTD function $E(t)$:

$$E(t) = \frac{d}{dt}\left[\frac{C_{out}}{C_o}\right]_{step} \qquad (13)$$

The positive step is usually easier to carry out experimentally that the pulse test, and it has the additional advantage that the total amount of tracer in the feed over the period of the test does not have to be known as it does in the pulse test. One possible drawback in this technique is that it is sometimes difficult to maintain a constant tracer concentration in the feed. Obtaining the RTD from this test also involves differentiation of the data and presents an additional and probably more serious drawback to the technique, because differentiation of data, on occasion, can lead to large errors. A third problem lies with the large amount of tracer required for this test. If the tracer is very expensive, a pulse test is usually used to minimize the cost.

2.6 Characteristics of 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 2.3 illustrates typical RTDs resulting from different reactor situations. ***Figure 2.3 (a)*** and ***(b)*** correspond to nearly ideal PFRs and CSTRs respectively. In ***Figure 2.3 (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 2.3 (d)***. ***Figure 2.3 (e)*** shows the RTD for the CSTR in ***Figure 2.3 (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.

2.6.1 Integral Relationships

The fraction of the exit stream that has resided in the reactor for a period of time shorter than a given value t is equal to the sum over all times less than t of $E(t)\Delta t$, or expressed continuously,

$$\int_0^t E(t)dt = \boxed{\begin{array}{l}\text{Fraction of effluent, which}\\ \text{has been in reactor for less}\\ \text{than time } t\end{array}} = F(t) \tag{14}$$

Analogously,

$$\int_t^\infty E(t)dt = \boxed{\begin{array}{l}\text{Fraction of effluent, which}\\ \text{has been in reactor for}\\ \text{longer than time } t\end{array}} \quad 1 - F(t) \tag{15}$$

Because t appears in the integration limits of these two expressions, equation (14) and (15) are both functions of time. $F(t)$ can be calculated at various times t from area under the curve of $E(t)$ versus t plot.

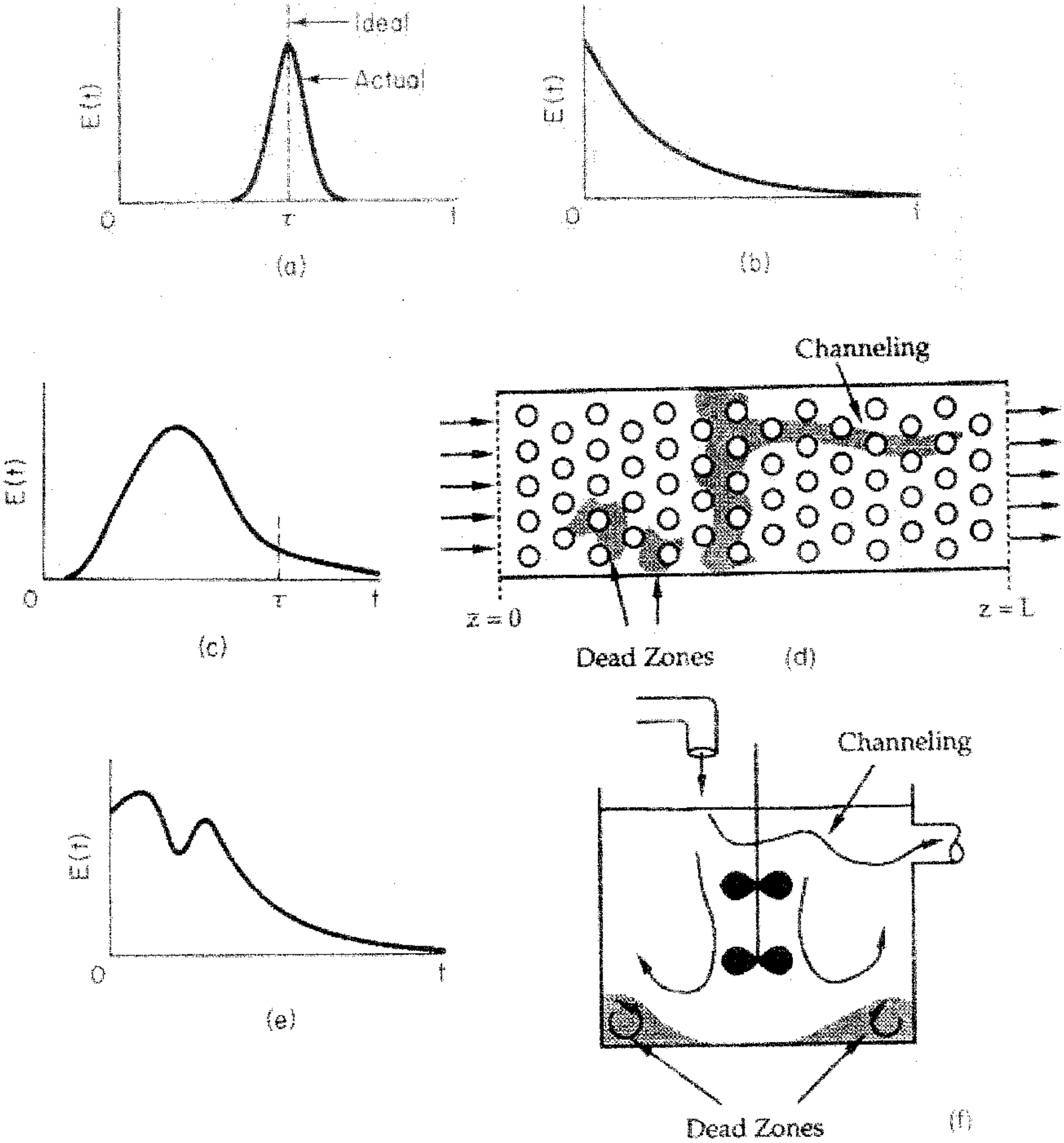

Figure 2.3 (a) RTD for near plug flow reactor; (b) RTD for near perfectly mixed CSTR; (c) RTD for packed-bed reactor with dead zones and channeling; (d) packed-bed reactor; (e) tank reactor with short-circuiting flow (bypass); (f) CSTR with dead zone.

2.6.2 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,

$$t_m = \frac{\int_0^\infty tE(t)dt}{\int_0^\infty E(t)dt} = \int_0^\infty tE(t)dt = \frac{\int_0^\infty tC(t)dt}{\int_0^\infty C(t)dt} = \frac{\sum t_i C_i \Delta t_i}{\sum C_i \Delta t_i} \qquad (16)$$

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 utilisation 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.

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.

2.6.3 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

$$\sigma^2 = \int (t - t_m)^2 E(t) dt \tag{17}$$

Alternatively,

$$\sigma^2 = \frac{\int_0^\infty (t - t_m)^2 C(t) dt}{\int_0^\infty C(t) dt} = \frac{\sum (t_i - t_m)^2 C_i \Delta t_i}{\sum C_i \Delta t_i} \tag{18}$$

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 2.4*** illustrates these terms.

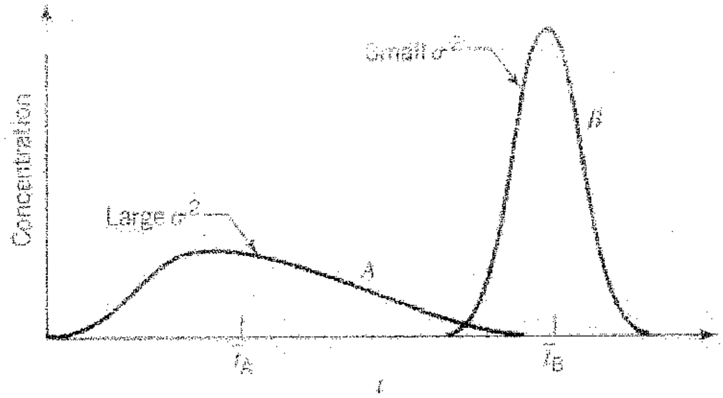

Figure 2.4 Variance for matching theoretical curves

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

$$s^3 = \frac{1}{\sigma^{3/2}} \int_0^\infty (t - t_m)^3 E(t) dt \qquad (19)$$

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.

2.7 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. Therefore, to characterize the spreading, according to dispersion model (***Figure 2.5***), 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.

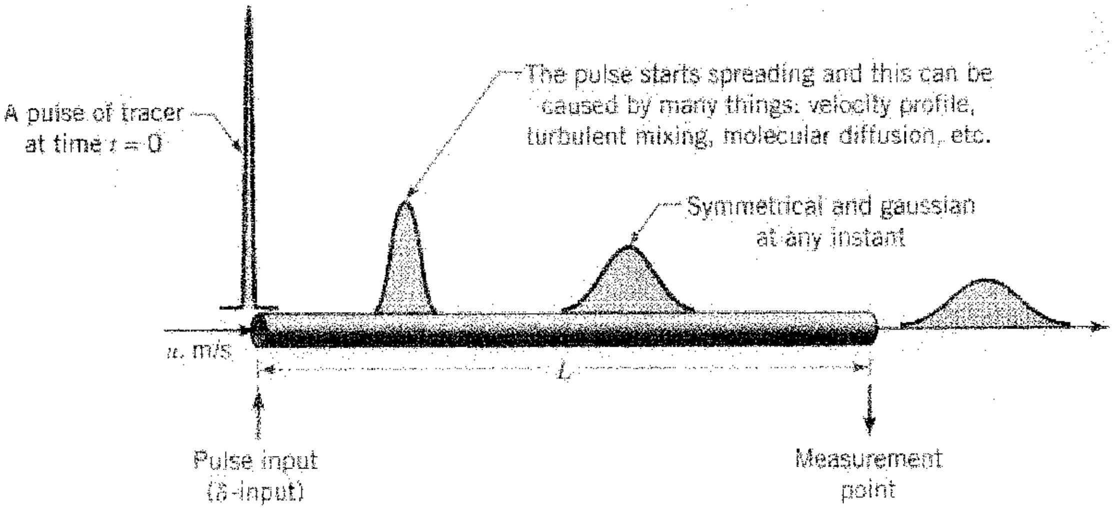

Figure 2.5 The spreading of tracer according to the dispersion model

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 2.6*** 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.

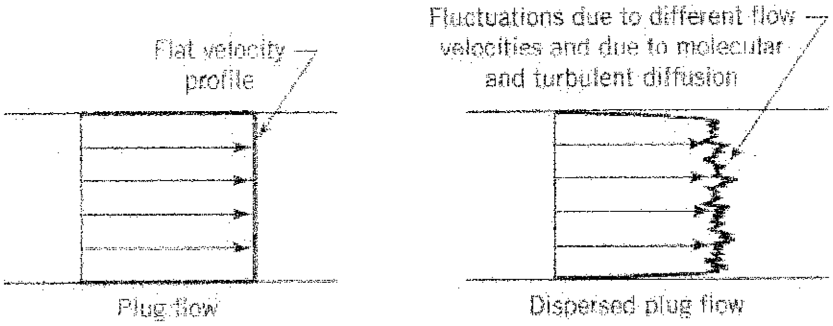

Figure 2.6 Representation of the dispersion (dispersed plug flow) model

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;

$$\frac{\partial C}{\partial t} = D \ \frac{\partial^2 C}{\partial x^2} \tag{20}$$

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

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

Where the parameter *D*, which is called the longitudinal or axial dispersion coefficient, uniquely characterizes the degree of back-mixing 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

$$\frac{\partial C}{\partial \theta} = \left(\frac{D}{uL}\right)\frac{\partial^2 C}{\partial z^2} - \frac{\partial C}{\partial z} \tag{22}$$

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,

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

$\left(\frac{D}{uL}\right) \to \infty$ 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.

$$\mathrm{Pe} = \frac{1}{D/uL} \tag{23}$$

Which the dispersion number is also defined by

$$\left(\frac{D}{uL}\right) = \frac{\sigma_\theta^2}{2} \tag{24}$$

And

$$\sigma_\theta^2 = \left(\frac{u}{L}\right)^2 \sigma^2 \tag{25}$$

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

Peclet number value increased with the increase in liquid space velocity indicating the reduction of back-mixing with higher liquid flow rate. The use of fine size of diluent also reduced the dependency of Peclet number on space velocity. Increasing trend in Peclet number with increasing liquid velocity for the upflow mode of operation when a larger size of particles was used.

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. The gas velocity affects the axial dispersion coefficient for upflow operation.

2.8 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. Thus, 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 2.7***). Arrangement of packing and adequate distribution of liquid can disrupt the ideal behavior of plug flow due to the channeling of liquid.

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.

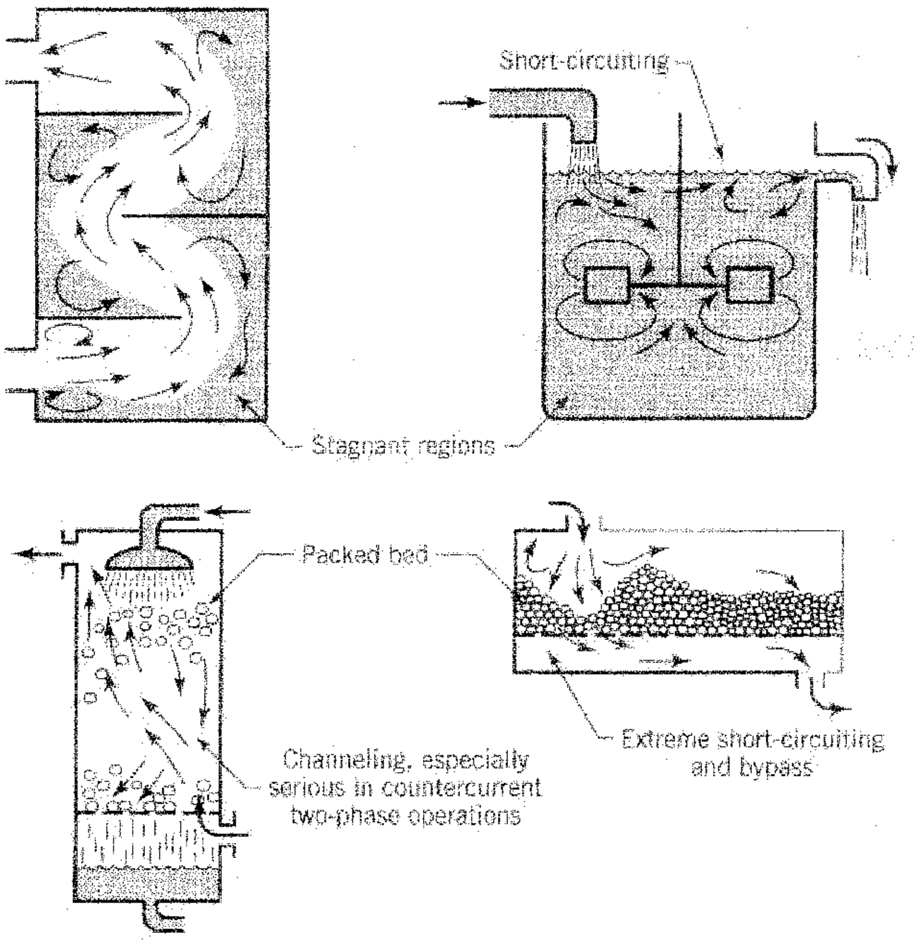

Figure 2. 7 Non-ideal flow patterns, which may exist in process equipment

The mean residence time can be determined by the equation:

$$t_m = \frac{\sum (t_i C_i \Delta t_i)}{\sum C_i \Delta t_i} \qquad (26)$$

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

$$T_d = \frac{V}{Q} \qquad (27)$$

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

$$V_{stagnant} = V\left(1 - \frac{t_m}{T_d}\right) \qquad (28)$$

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.

2.9 Late Mixing

In a 'perfectly mixed' of a reactor, the entering fluid is assumed to be distributed immediately and evenly throughout the reacting mixture. This mixing is assumed to take place even on the micro scale, and elements of different ages mix together thoroughly to form a completely micromixed fluid. If fluid elements of different ages do not mix at all, the elements remain segregated from each other and the fluid is termed completely segregated or late mixing. The extremes of complete micromixing and complete segregation are the limits of the micromixing of a reacting mixture.

For a continuous-flow system, each exit stream corresponds to a specific residence time in the reactor. Batches of molecules are removed from the reactor at different locations along the reactor in such a manner so as to duplicate the RTD function, $E(t)$. The molecules removed near the entrance to the reactor correspond to those molecules having short residence times in the reactor. Physically, this effluent would correspond to the molecules that channel rapidly through the reactor. The farther the molecules travel along the reactor before being removed, the longer their residence time. The points at which the various groups or batches of molecules are removed correspond to the RTD function for the reactor.

Because there is no molecular interchange between globules, each acts essentially as its own batch reactor. The reaction time in any one of these tiny batch reactors is equal to the time that the particular globule spends in the reaction environment. The distribution of residence times among the globules is given by the RTD of the particular reactor.

2.10 Early Mixing

In a reactor with segregated fluid, mixing between particles of fluid does not occur until the fluid leaves the reactor. The reactor exit is, of course, the latest possible point that mixing can occur, and any effect of mixing is postponed until after all reaction has taken place. As soon as the fluid enters the reactor, it is completely mixed radically with the other fluid already in the reactor. The entering fluid is fed into the reactor through the side entrances in such a manner that the RTD the reactor with side entrance is identical to the RTD of the real reactor. In the reactor with side entrances, mixing occurs at the earliest possible moment consistent with the RTD. Thus, the effect of the mixing occurs as much as possible throughout the reactor, and this situation is termed the condition of maximum mixedness or early mixing. Maximum mixedness is consistent with a given residence time distribution.

CHAPTER 3

PROJECT OVERVIEW

3.1 Research Project Work

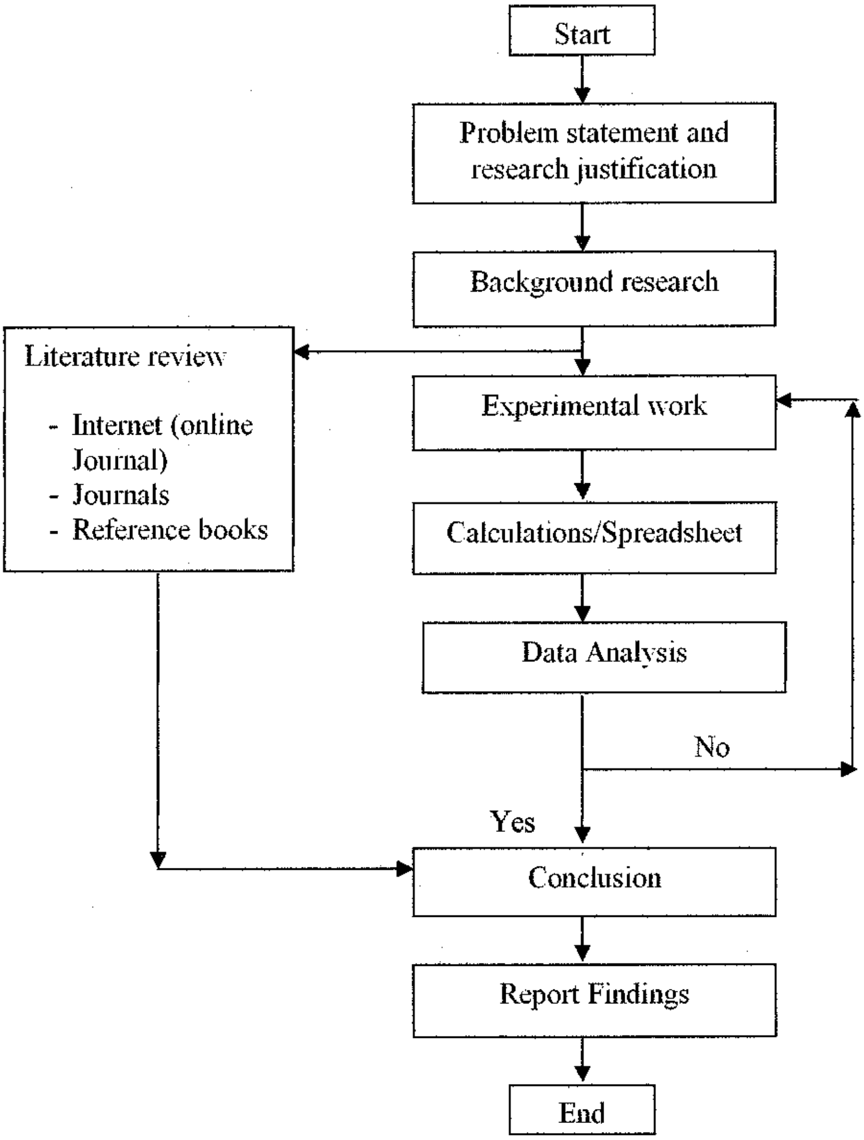

Figure 3.1 Research methodology

The research was started with the problem statement identification and justification. Once confirmed, further background research was done. This was accompanied by a series of literature review, which is a continuous and on-going process throughout the research. The literature review was done from several resources, mainly the internet and the Information and Resource Center, Universiti Teknologi PETRONAS. This includes journals, reference books and also online services such as online journals.

Once all the theoretical and findings from previous researches were firmed, sets of experiments were conducted. The results of the experiments were later analysed and compare with the theory and findings from precious researches. If the results were not satisfactory, the experiment was repeated. After analysed, the results were then compiled and conclusions were made.

3.2 Experimental Set-up

The simplified schematic diagram for the experimental set-up used is shown in ***Figure 3.2***. This set-up consists of a packed bed reactor, a system for feeding controlled amount of gas and liquid, tracer injection system and conductivity measurement instrument (conductivity probe) for detecting the concentration of tracer. The data obtained from the experiment is recorded through data Acquisition System (DAS).

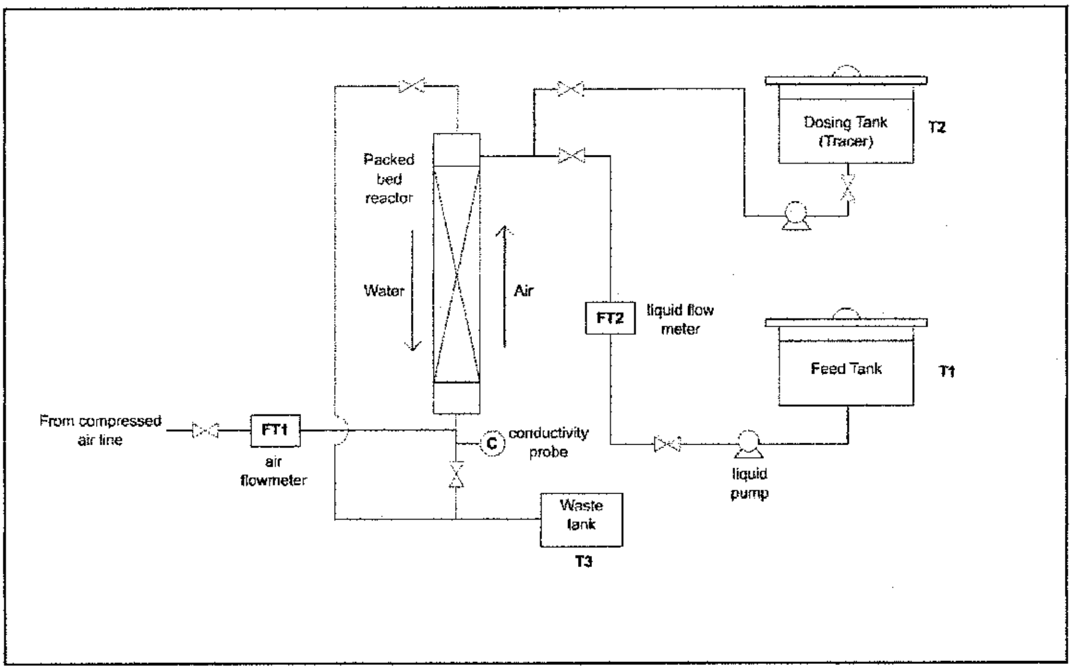

Figure 3.2 Process Diagram for RTD Studied in Tubular Reactor (BP 112)

The reactor is essentially a column made from borosilicate glass with column internal diameter of 82 mm and a height of 1500 mm. The reactor is packed with 8 x 8 mm Raschig rings. Both top and bottom of the reactor is fitted with stainless steel cap with fitted ports for differential pressure tapping.

In this study, air and water used as the gas and liquid phase due to the inexpensive and non-hazardous nature of these two materials. Sodium chloride acts as the tracer medium. The flow rate of air and water into the system is controlled using mass flow meter and the exit concentration of the tracer medium was measured using a conductivity probe. ***Table 3.1*** summarise the range of gas and liquid flow rate considered for the experiment.

Before starting the experiment, the bed was fully wetted with water. Then, the air and water flow rate is slowly adjusted to the desired values. The outlet valve of water is adjusted until water level is at the same height as the bed. The tracer was introduced through the reactor as a pulse through the dosing pump. Extra care was taken in deciding the length of time to introduce the tracer. The tracer injection period must take place over a short period of time in comparison with the residence time to ensure negligible dispersion between the point of injection and entrance to the reactor.

Operating liquid hold-up of a system is defined as the ratio of the amount of liquid in the reactor during operation to the total volume of the packed bed reactor. The operating liquid hold-up is determined as follows. The bed was fully wetted and the liquid and gas flow rate was then adjusted to the set value for the experiment. The system was left to attain a steady state condition, usually takes about 30 minutes. After reaching steady state, the gas and liquid flow is stopped simultaneously. The total free liquid in the reactor was drained in a liquid collector and measured.

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

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

CHAPTER 4

RESULTS AND DISCUSSION

4.1 Effect of Gas/Liquid Ratio on Pressure Drop and Operating Liquid Hold-up

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. The pressure drop throughout the experiment is recorded and the result is shown in ***Figure 4.1***. For all constant liquid flow rates, the pressure drop increases with increasing gas/liquid ratio.

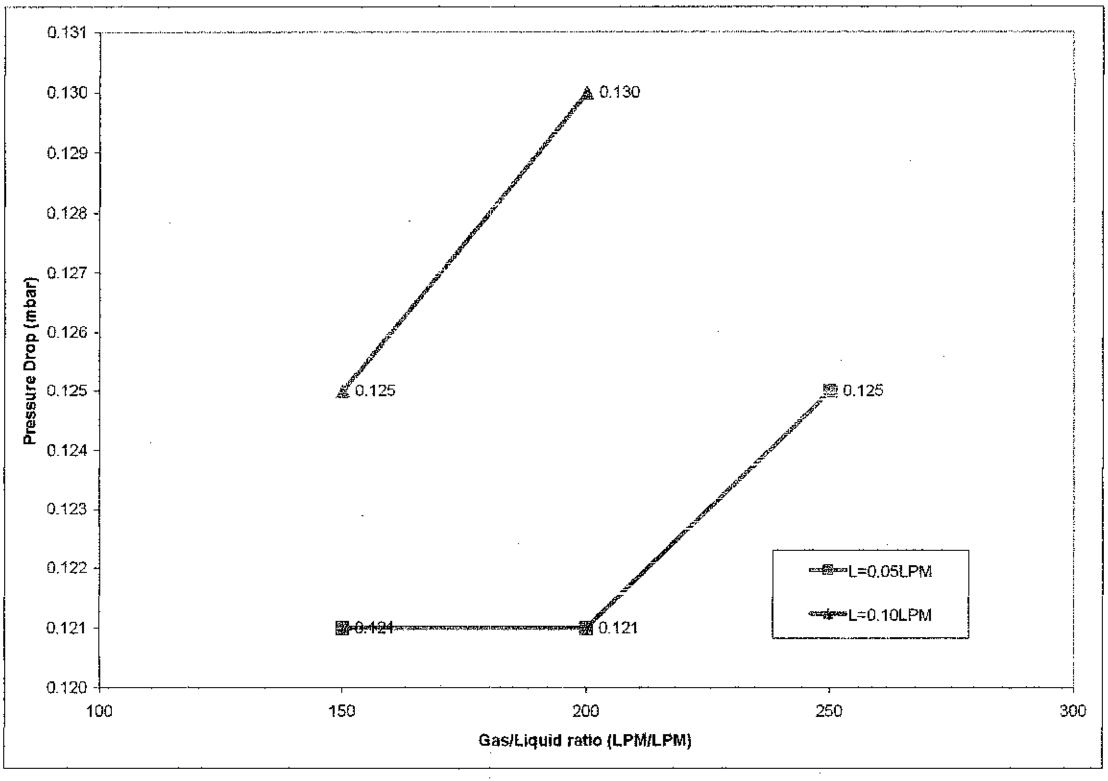

Figure 4.1 Effect of gas/liquid ratio on pressure drop at constant liquid flow rate of 0.05 and 0.10 LPM

The results obtained in this experiment is in strong agreement with results obtained by Varma et al. [4], who shows that the pressure drop increased with the gas and liquid flow rates in all the regimes.

4.1.2 Effect of Gas/Liquid Ratio on Operating Liquid Hold-up

Liquid hold-up has a direct influence on the liquid side mass transfer, loading behaviour and gas phase pressure gradients. Experimentally, operating liquid hold-up is the portion of liquid that is drained out of a catalyst bed when both gas and liquid are stopped after reaching steady state.

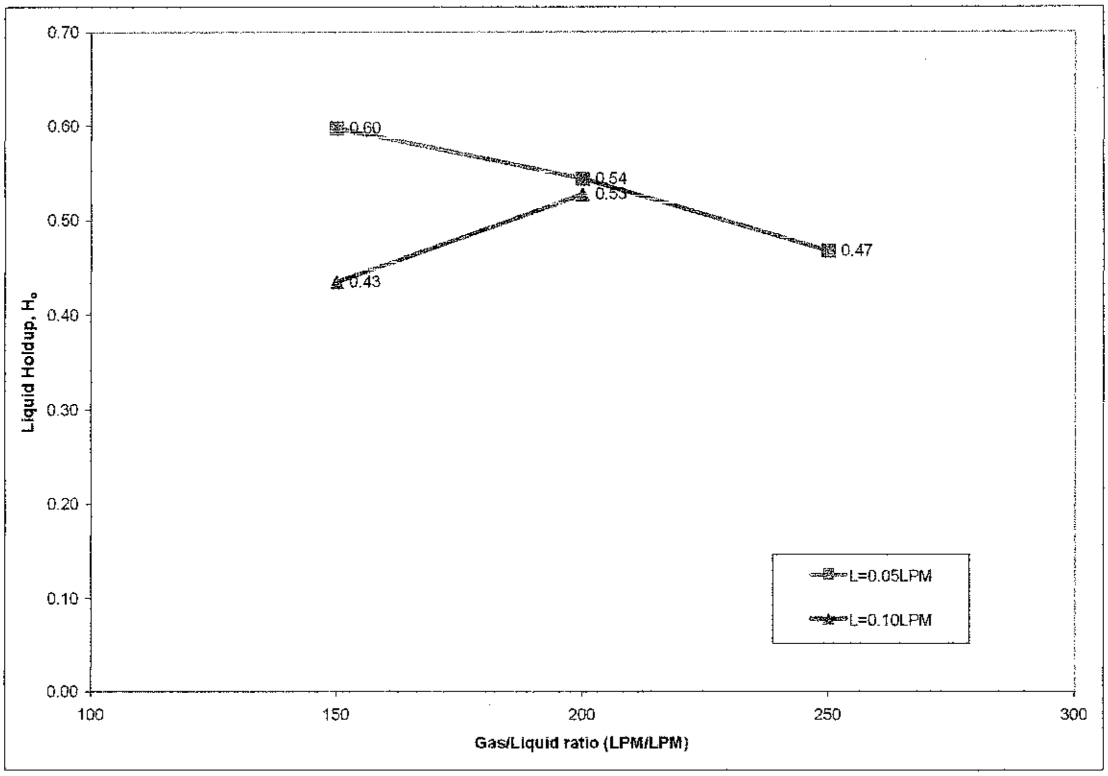

Figure 4.2 Effect of gas/liquid ratio on operating liquid hold-up at constant liquid flow rates of 0.05 and 0.10 LPM

Figure 4.2 shows that the operating liquid hold-up decreases with increasing gas/liquid ratio. This is similar to the findings by Chander et al. [5]. Also, as the liquid flow rates increase, the operating liquid holdup decreases. Thus, higher liquid flow rates could increase the reaction rate in the three phase of packed bed reactor.

Study by Chander et al. [5] showed that for the downflow mode of operation, the liquid hold-up was independent of gas flow rate for large and small sizes of diluent.

4.2 Effect of Gas/Liquid Ratio on Moments of RTD

The variation in concentration data obtained from the experiments is shown as an E curve. The values of Peclet number and mean residence time of the liquid were calculated and analysed from the RTD data obtained from the experiment.

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 Effect of Gas/Liquid ratio on *E* curve

The E curve is the distribution needed to account for non ideal flow. All three E curves (***Figure 4.3, Figure 4.4*** and ***Figure 4.5***) showed that at any gas/liquid ratio, with low liquid flow rate, which in this case is 0.05 LPM, the E curve exhibited deviation from ideal plug flow reactor and approached mixed flow behaviour. However, as the liquid flow rates are increased, the E curves approach the behaviour of a plug flow. From these three graphs, it is clearly showed that the concentration profile of tracer is unaffected by the gas/liquid ratio but is a strong function of liquid flow rates of the system.

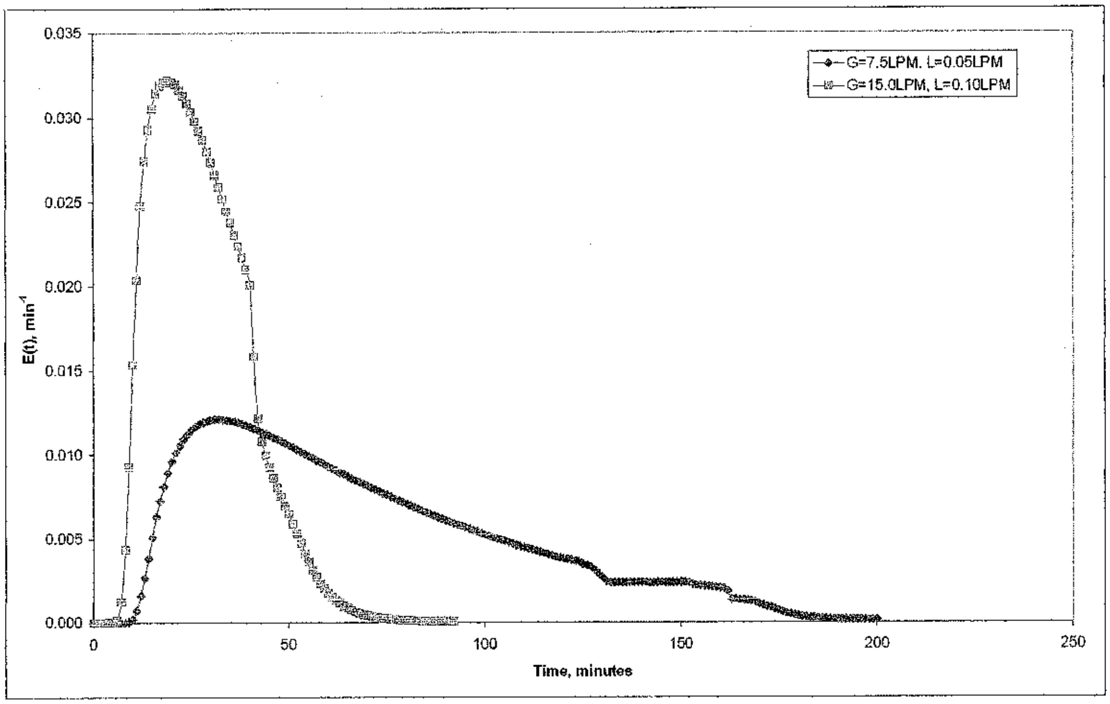

Figure 4.3 E curve for gas/liquid ratio of 150

There are some fluctuations at concentration-time tail in ***Figure 4.3*** and ***Figure 4.4*** due to equipment failure in mass flow meter. The mass flow meter needs maintenance for better results during the experiment in the future.

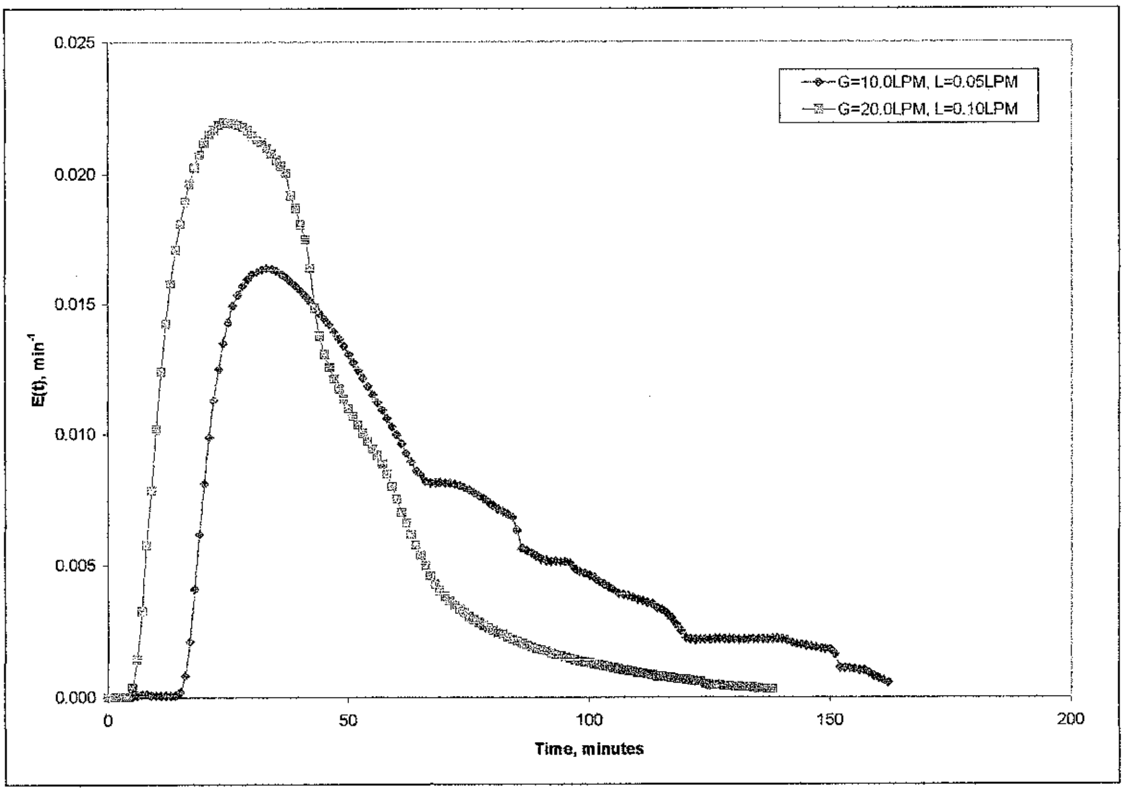

Figure 4.4 E curve for gas/liquid ratio of 200

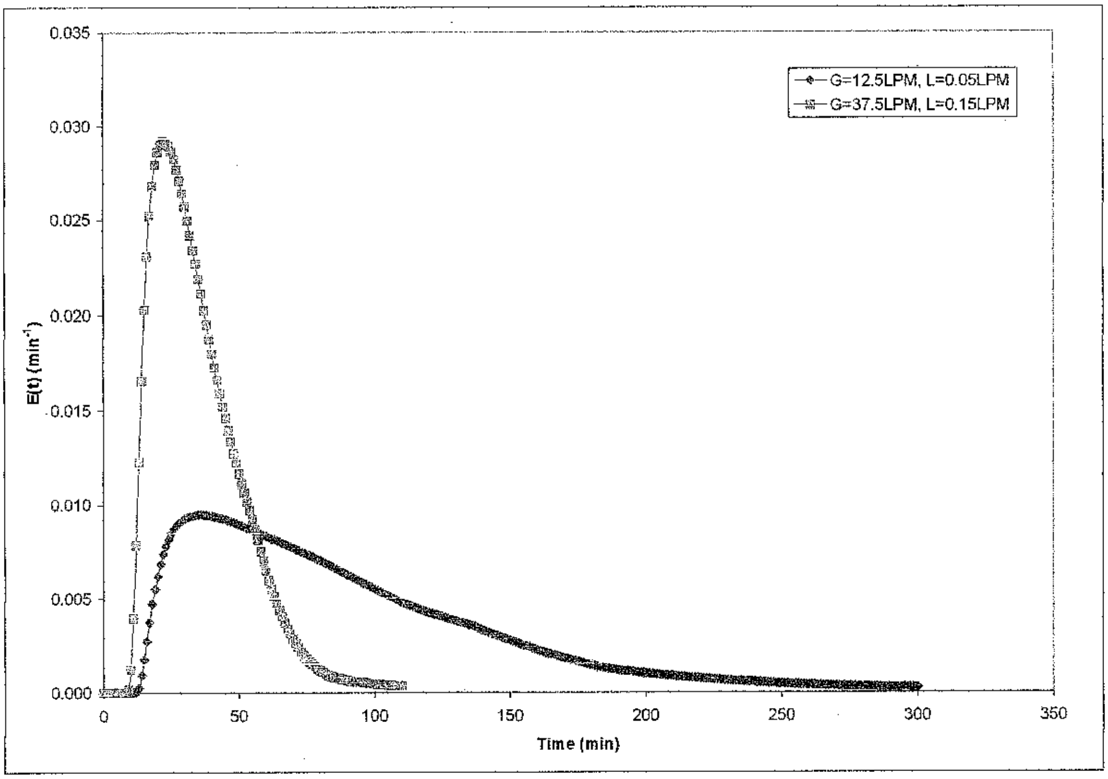

Figure 4.5 *E* curve for gas/liquid ratio of 250

4.2.2 Effect of Gas/Liquid Ratio on Mean Residence Time

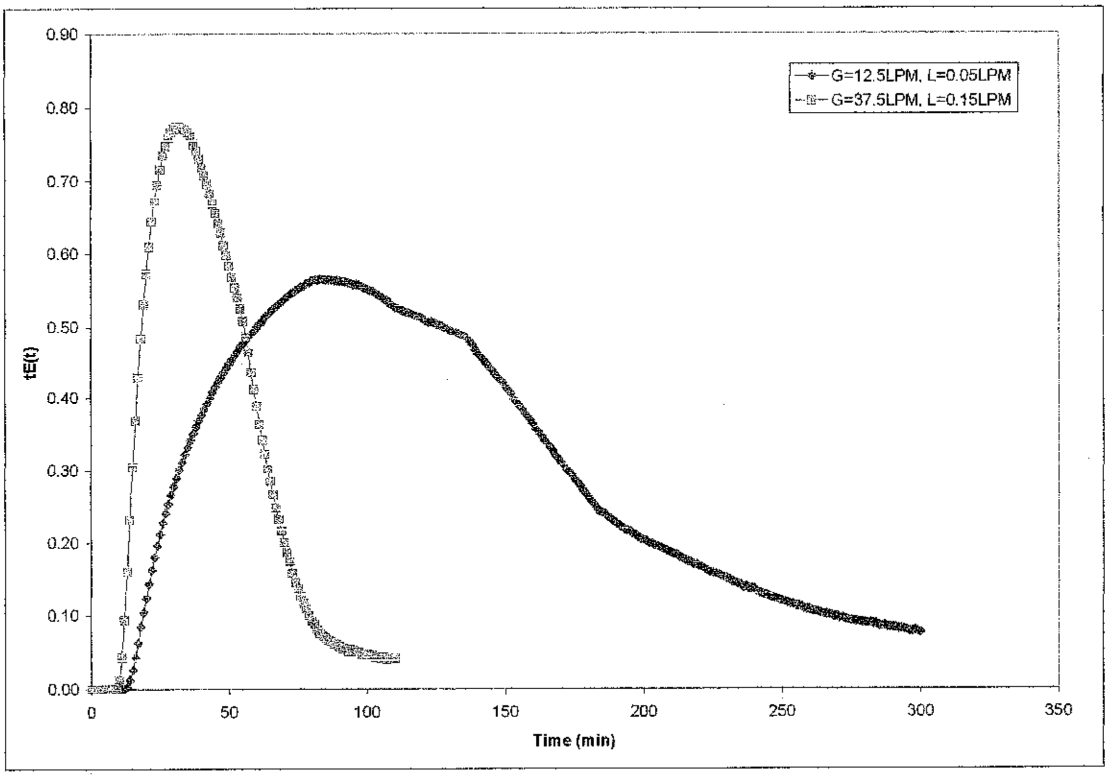

Figure 4.6 Mean residence time for gas/liquid ratio of 250

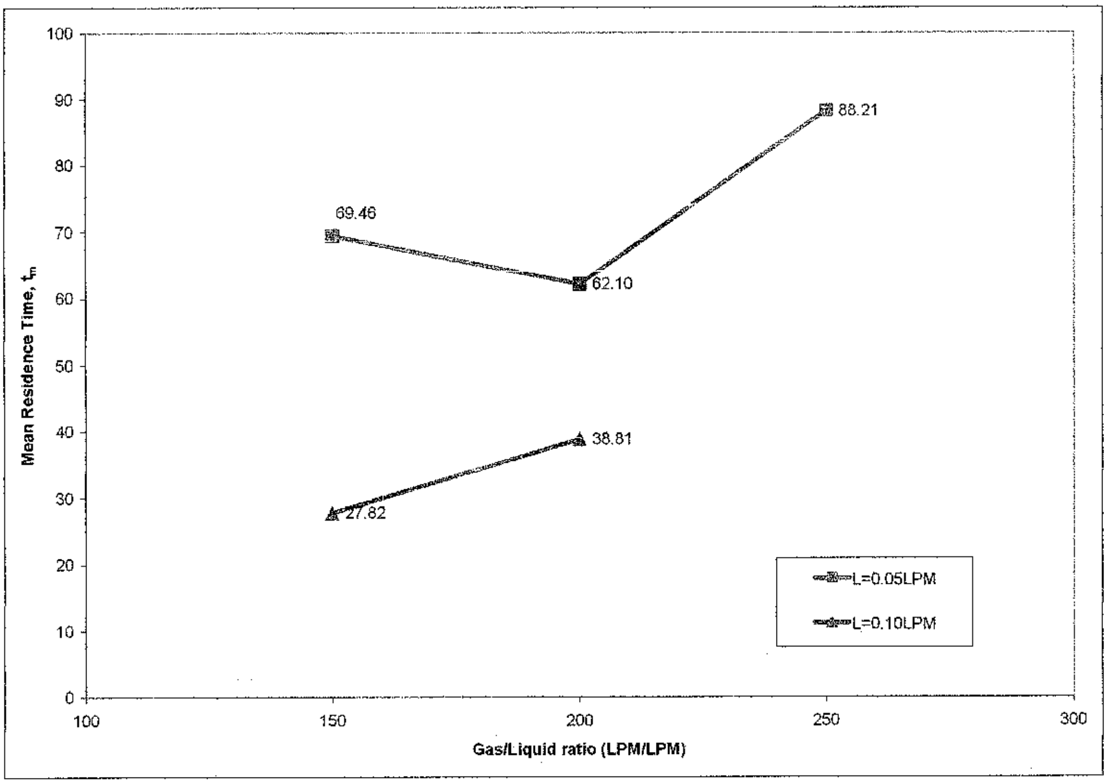

Figure 4.7 Effect of gas/liquid ratio on mean residence time of liquid at constant liquid flow rates of 0.05 and 0.10 LPM

The nominal holding time, τ, or the space time describes the time the material spends in an ideal reactor. For a closed non ideal system with no dispersion, this should equals to the mean residence time, τ_m, of the reactor.

Example of mean residence time curve is shown in ***Figure 4.6*** for gas/liquid ratio of 250. Experimentally, it is observed that there is a considerable increase in mean residence time as the gas/liquid ratio is increased, illustrated in ***Figure 4.7***. Furthermore, the mean residences time is increased with decreasing variation of liquid flow rates. These results are also agreed with Chander et al. [5], which proved that the mean residence time is increased with gas/liquid ratio for both, downflow and upflow, mode of operation.

The analysis shows that the mean residence time for this reaction is high. Furthermore, the peak of the *E* curves (***Figure 4.3***, ***Figure 4.4*** and ***Figure 4.5***) occurs earlier than the mean residence time. This is an indication of the possibility of stagnant zone in the reactor. This 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 or 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, excessive thermal reaction would not likely to occur.

4.2.3 Effect of Gas/Liquid Ratio on Variance

The next most important descriptive quantity is the spread of the distribution which is represented by variance.

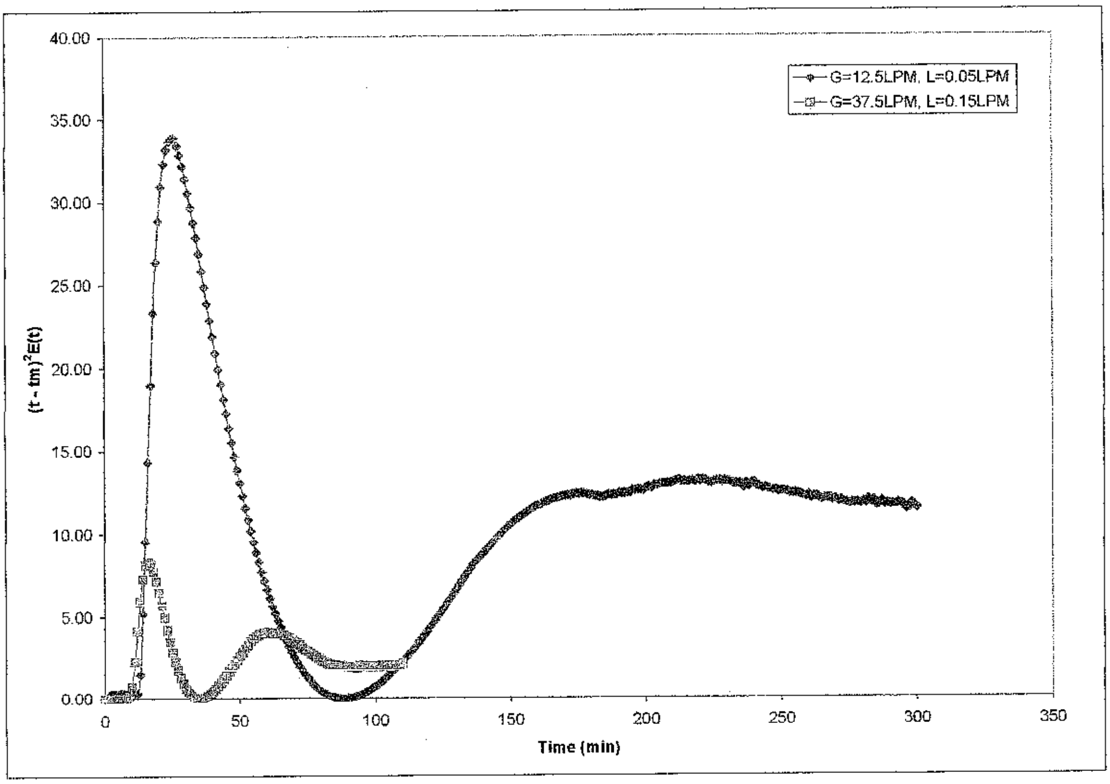

Figure 4.8 Variance curve for gas/liquid ratio of 250

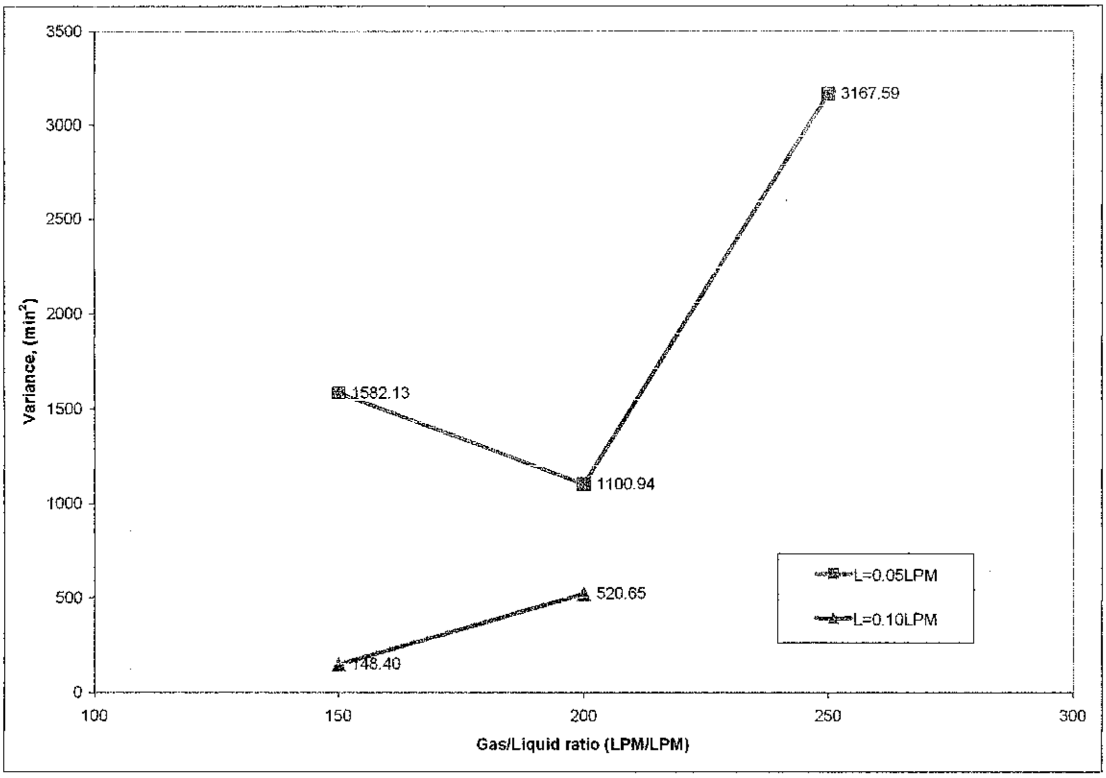

Figure 4.9 Effect of gas/liquid ratio on variance at constant flow rates of 0.05 and 0.10 LPM

Example of variance curve is illustrated in ***Figure 4.8***, which is for gas/liquid ratio of 250. Experimentally, from ***Figure 4.9***, the variance increases as the gas/liquid ratio increases. The variance is also increased with decreasing variation of liquid flow rates. These results are also consistent with the *E* curve in ***Figure 4.3***, ***Figure 4.4*** and ***Figure 4.5***; the *E* curve for highest constant liquid flow rate has the highest peak among the other constant liquid flow rates for all gas/liquid ratio, which indicates smaller variance, and vice versa. Thus, it is proved that the smaller the variance, the smaller the distribution's spread.

In order to achieve plug flow characteristic, smaller variance is required. Thus, from this experiment, it is proved that operating packed bed reactor at high gas/liquid ratio with lower gas and liquid flow rate, would result smaller variance, which is approaching plug flow behavior.

4.2.4 Effect of Gas/Liquid Ratio on Skewness

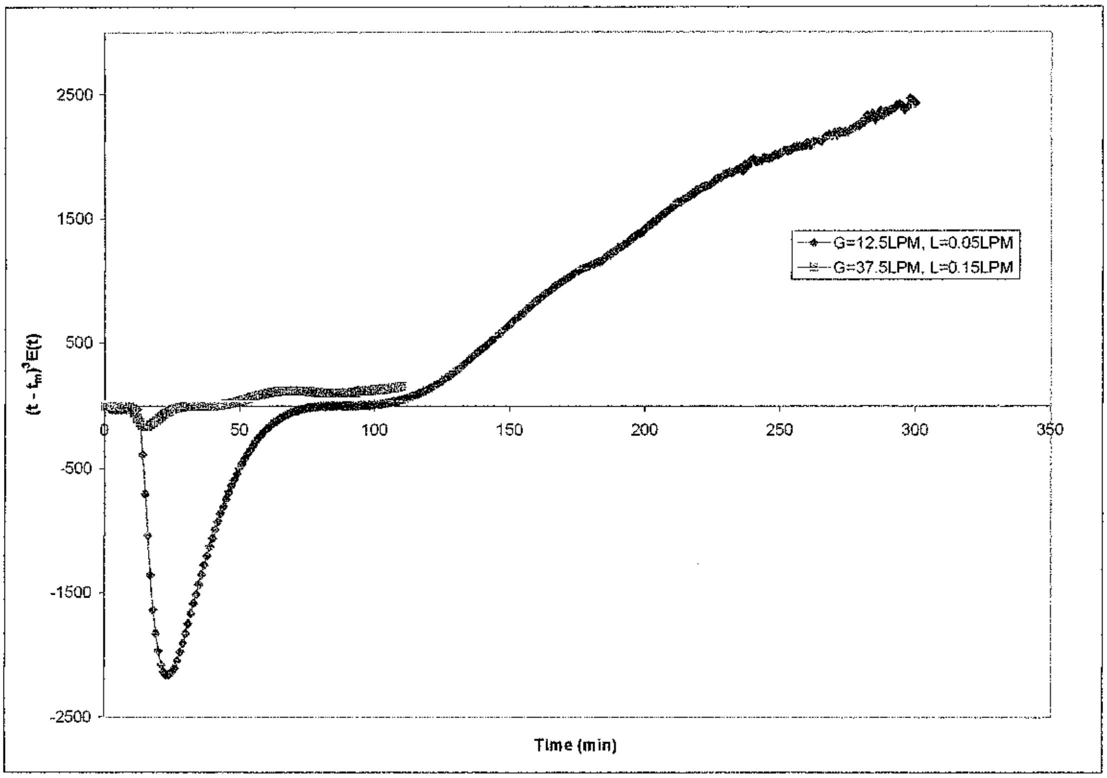

Figure 4.10 Skewness curve for gas/liquid ratio of 250

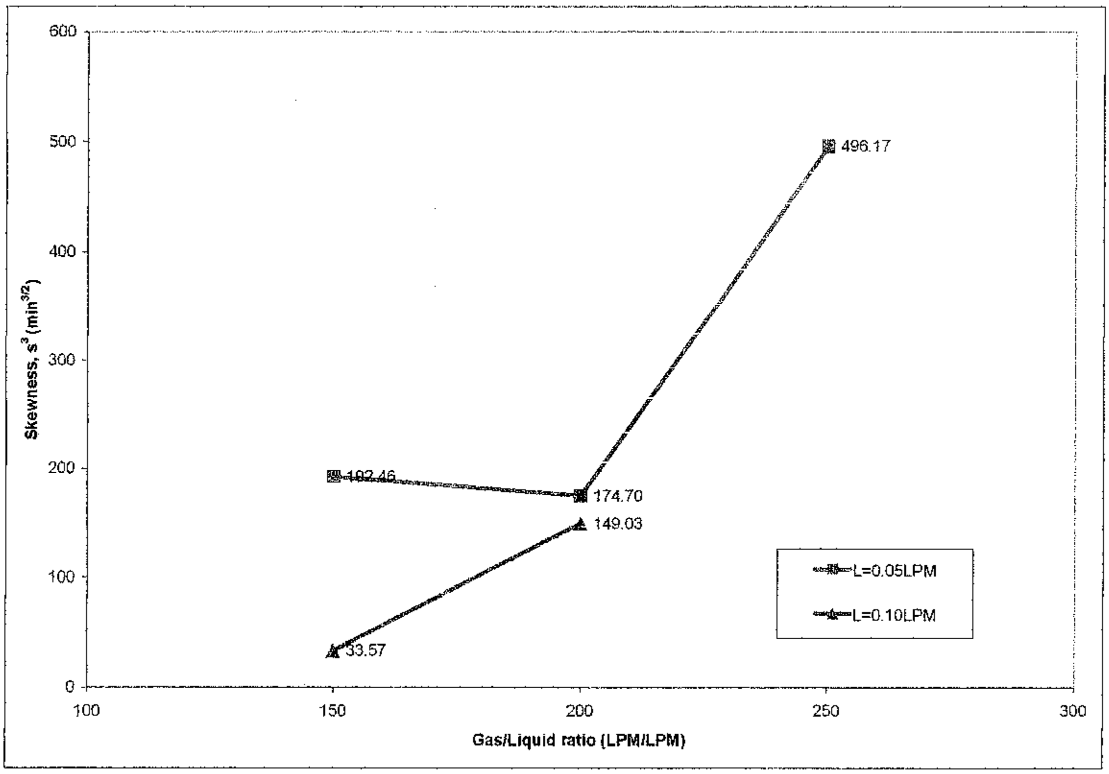

Figure 4.11 Effect of gas/liquid ratio on skewness at constant liquid flow rates of 0.05 and 0.10LPM

The skewness curve for gas/liquid ratio of 250 is shown in ***Figure 4.10***. Analysis of experiments have resulted that the skewness increases as the gas/liquid ratio increases, refer to ***Figure 4.11***, and the skewness also increases as variation of liquid flow rates decreases. Again, these results are reflected by the *E* curve in ***Figure 4.3***, ***Figure 4.4*** and ***Figure 4.5***; the *E* curve for high constant liquid flow rate is not skewed far from the reference of mean compared to other constant liquid flow rates for all gas/liquid ratio, which indicates low skewness, and vice versa. Thus, it is proved that the lower the skewness, the less skewed the distribution is, from its mean.

In order to achieve plug flow behavior, it is desirable to have smaller value of skewness, which also meant that the distribution is skewed less from the reference of the mean. Unlike *E* curve for lower flow rates which has high value of skewness and skewed more towards the left, it is preferred to operate packed bed reactor at higher gas/liquid ratio with lower gas and liquid flow rates, which will result lesser skewness and will approach the plug flow behavior.

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

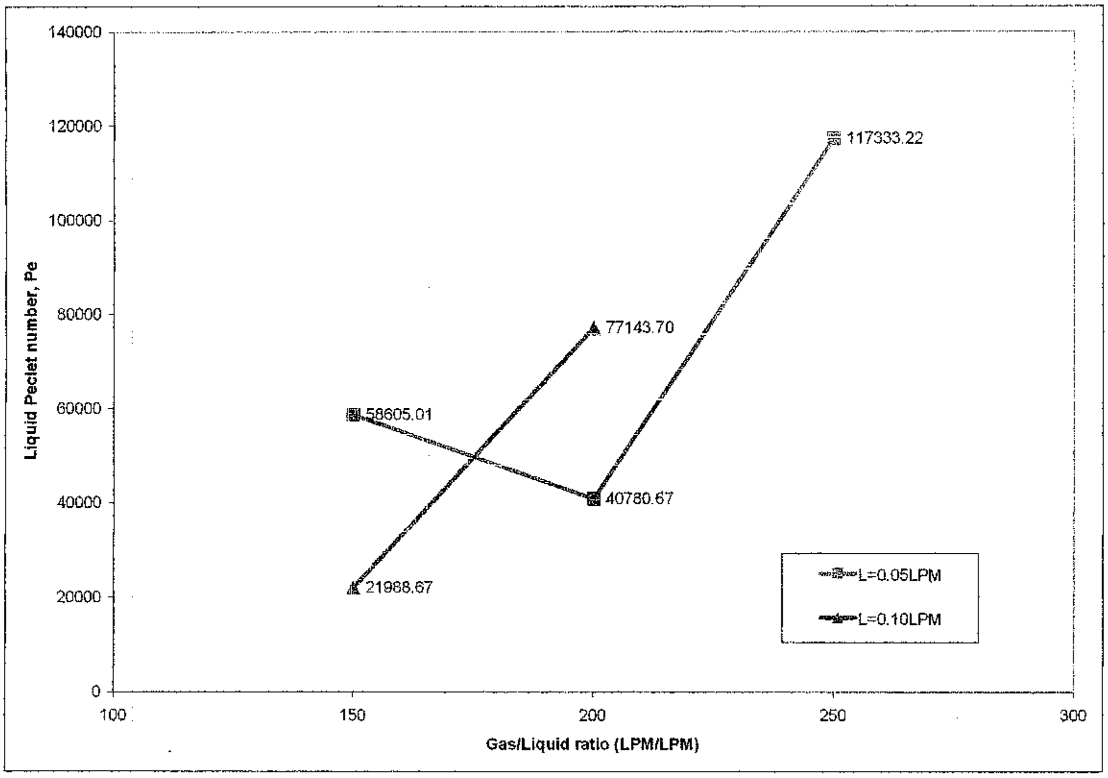

Figure 4.12 Effect of gas/liquid on liquid Peclet number at constant liquid flow rates of 0.05 and 0.10 LPM

Axial dispersion of a system gives an indication to the degree of mixing inside a reactor and is represented by the Peclet number. Lower value of Peclet number indicates higher degree of dispersion in the system.

Experimentally, the Peclet number relatively increases with increasing gas/liquid ratio, as well as decreasing variation of liquid flow rates, as illustrated in ***Figure 4.12***. However, these results do not agree with studies done by Chander et al. [5], who reported that the Peclet number value increased with the increase in LHSV for modes, downflow and upflow, of operation indicating the reduction of back-mixing with higher liquid flow rate.

Besides, ***Figure 4.13*** shows that Peclet number is decreasing as gas/liquid ratio increasing. This result was obtained from previous student who did the same research for counter-current flow. Further study is currently on-going to investigate this difference.

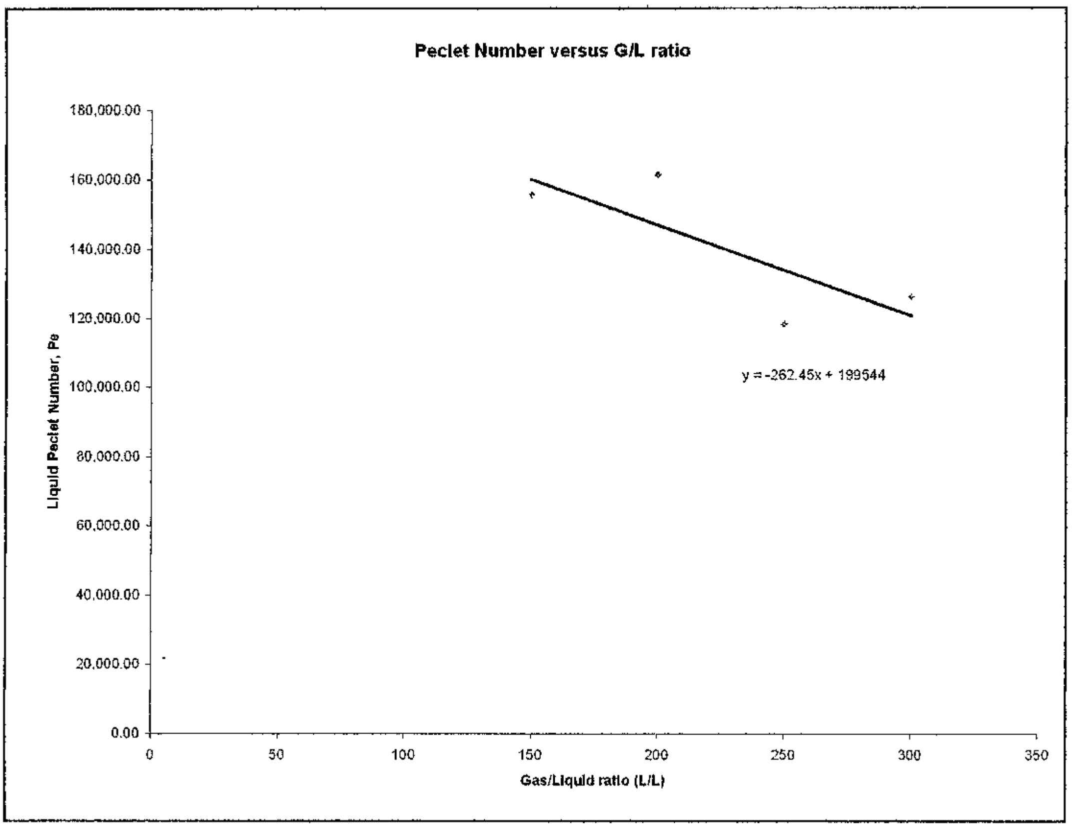

Figure 4.13 Effect of gas/liquid on liquid Peclet number from other research

4.3.2 Effect of Gas/Liquid Ratio on Stagnant Zone Volume

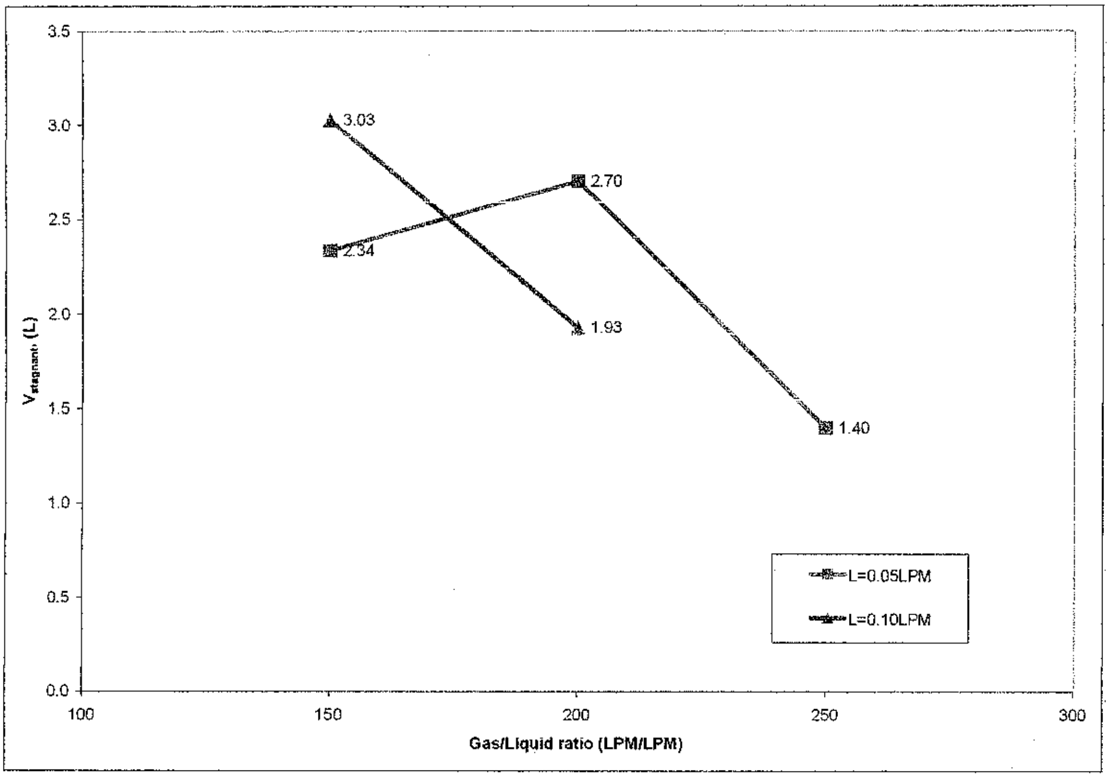

Figure 4.14 Effect of gas/liquid ratio on stagnant zone volume at constant liquid flow rates of 0.05 and 0.10 LPM

The existence of stagnant zone inside a reactor is undesirable as it reduces the active volume of the reactor, hence reducing the efficiency and the overall performance of the reactor. The variation in stagnant zone volume at different liquid flow rates and gas/liquid ratio is shown in ***Figure 4.14***. It is observed that operating at higher gas/liquid ratio or higher liquid flow rate can reduce the stagnant zone volume. This result is in agreement with Sata et.al [7].

At low gas/liquid ratio with 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

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. The theories introduced in this research are considered parallel with the experimental results.

Operating liquid hold-up, H_o, is decreasing with increasing gas/liquid ratio in experiment. Also, the liquid hold-up is decreased with increasing liquid flow rates. Thus, for a desirable process, liquid hold-up must be minimized, which can be achieved at high gas/liquid ratio with high gas and liquid flow rates.

Mean residence time, t_m is increased with increasing gas/liquid ratio. This can be explained by the induced circulatory motion of liquid with higher flow rates. 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 decreasing variation of liquid flow rates. Thus, the desirable operation can be done at high gas/liquid ratio with lower gas and liquid flow rates.

Variance, σ^2 is decreased if the gas/liquid ratio is decreased. Also, the variance is decreased with increasing variation 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, which can be achieved at low gas/liquid ratio, but with higher gas and liquid flow rates. This is also reflected in the E curves of all gas/liquid ratios.

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 increasing variation 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. This can be done at lower gas/liquid ratio, with higher gas and liquid flow rates.

Axial dispersion of liquid, expressed by Peclet number, Pe, is increases with increasing gas/liquid ratio as well as decreasing variation of liquid flow rates. This condition is contradicted with the theory. Further study is on going to find out the difference between experimental and theoretical.

Stagnant zone volume, $V_{stagnant}$, can be reduced with higher gas/liquid ratio, as well as higher liquid flow rates. 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 flow rates with higher gas and liquid flow rates.

The information obtained from this residence time distribution study may therefore may help in improving the performance of a reactor as well as finding an optimum operating condition for a particular process.

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 Effect of Different Tracer Concentration

The RTD is determined experimentally by injecting 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. Coloured and radioactive materials are the two most common types of tracers.

So that, hydrodynamic characteristic can be affected by type of tracer. For further investigation in RTD analysis, it is necessary if the research could include the different types of tracer or different tracer concentration, used in the experiment, to observe and analyse the hydrodynamics characteristic in packed bed or fixed bed reactor.

5.2.2 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.

REFERENCES

1. **Iliuta, I.; Bildea, S.C.; Iliuta, M.C.; Larachi, F.:** *Analysis of Trickle Bed and Packed Bubble Column Bioreactors for Combined Carbon Oxidation and Nitrification*, Braz. J. Chem. Eng. Vol. 19 no. 1 São Paulo Jan./Mar. 2002
2. **Ergun, S.:** Fluid Flow Through Packed Columns, Chemical Engineering Program, 48, 89-94 (1952)
3. **Turpin, J.L.; Huntington, R.L.:** *Prediction of pressure drop for two phase two component concurrent flow in packed beds*, AIChE J., 13 (1967) 1196
4. **Varma, Y.B.G.; Khan, A.A.; Khan, A.:** *Flow regime identification and pressure drop in co-current gas-liquid upflow through packed beds*, Bioprocess Engineering 16 (1997) 355–360 Ó Springer-Verlag (1997)
5. **Chander, A.; Kundu, A.; Bej, S.K.; Dalai, A.K.; Vohra, D.K.;** *Hydrodynamic characteristics of cocurrent upflow and donwflow of gas and liquid in a fixed bed reactoe*, Fuel 80(2001) 1043 – 1053
6. **Stiegel, G.J.; Shah, Y.T.:** *Industrial Engineering Chemical Process Design Dev.*, 16 (1), 37 (1977)
7. **Sata, S.A.; Mohamed, A.R.; Kamaruddin, A.;** *Hydrodynamics experiment in the packed bed reactor*, 15th Symposium of Malaysian Chemical Engineers SOMChE 2001 (A3-2)
8. **Fogler, H. Scott**: Elements of Chemical Reaction Engineering, 2nd Edition, Prentice Hall International Series, New Jersey, 711 – 725 (1992).
9. **Perry's Chemical Engineers' Handbook,** 7th Edition, Mac Graw Hill.
10. **Nikolic, L.B.; Nikolic, V.D.; Veljkovic, V.B.; Lazic, M.L.; Skala. D.U.;** *Axial dispersion of the liquid phase in a three-phase Karr reciprocating plate column,* J. Serb. Chem. Soc. 69 (7) 581 – 599 (2004)
11. **Šerbula, M.S.; Stanković.;** *Hydrodynamic characteristics of a two-phase gas liquid flow upward through a fixed bed of spherical particles*, J. Serb. Chem. Soc. 66(1)53-64(2001)
12. **Levenspiel, O.:** Chemical Reaction Engineering, 2nd Edition, John Wiley & Sons, 253 – 330 (1972).

APPENDICES

APPENDIX A RTD Analysis for Gas/Liquid Ratio of 150

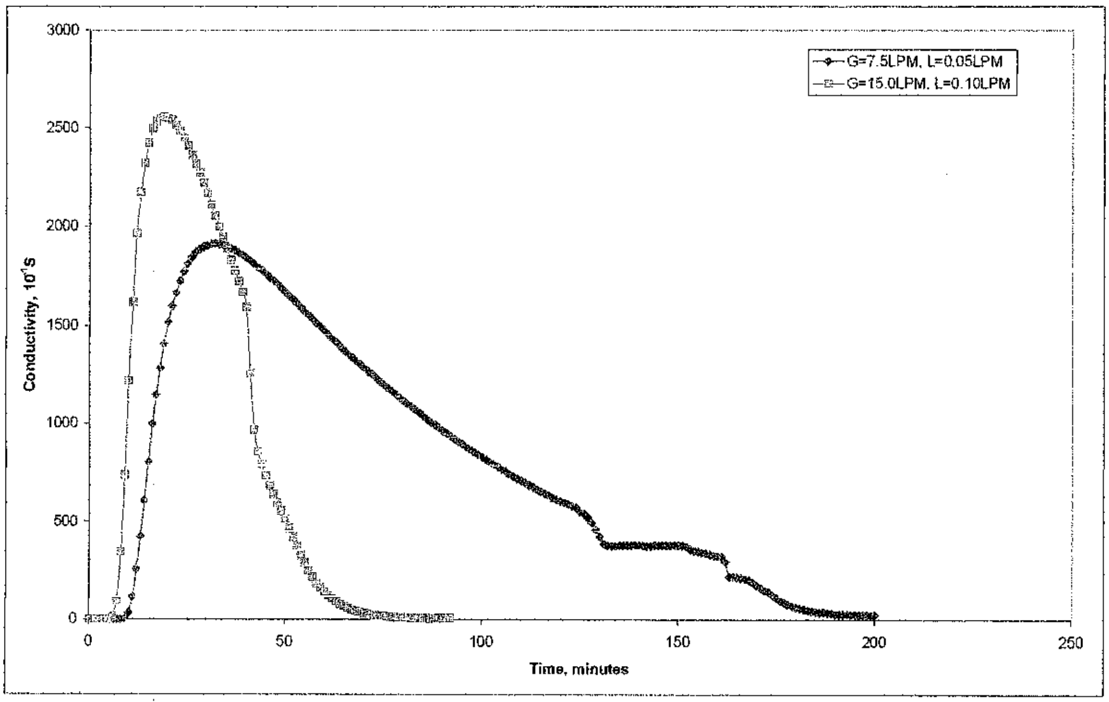

Figure A-1 Conductivity against time for gas/liquid ratio of 150

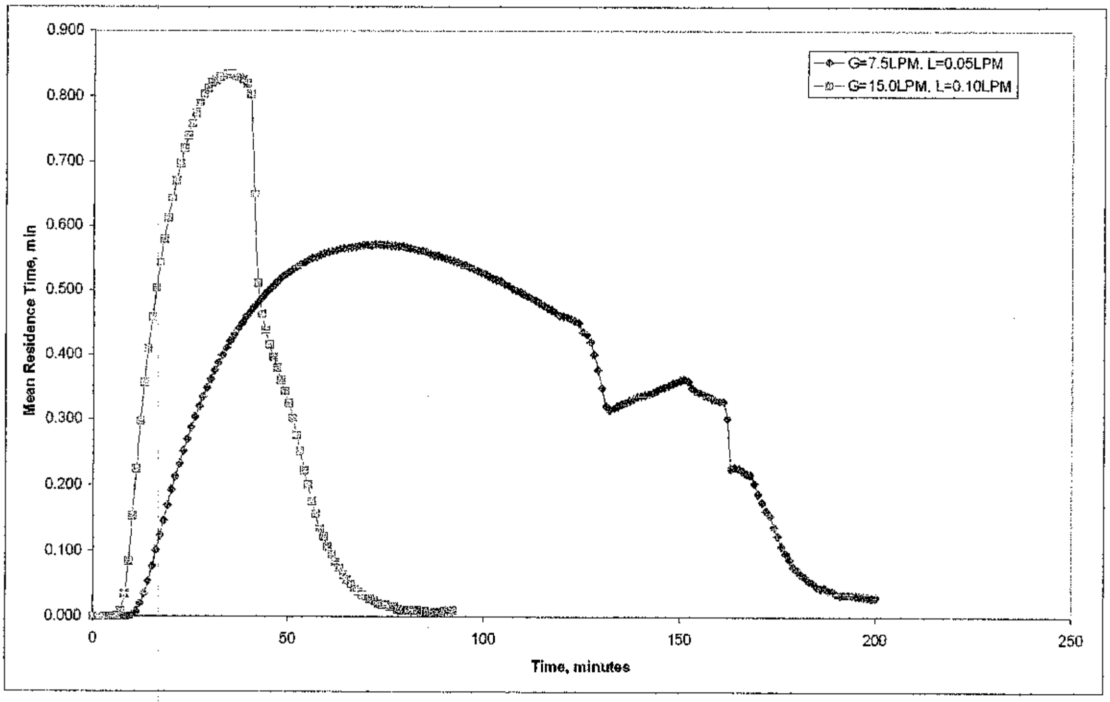

Figure A-2 Mean residence time curve for gas/liquid ratio of 150

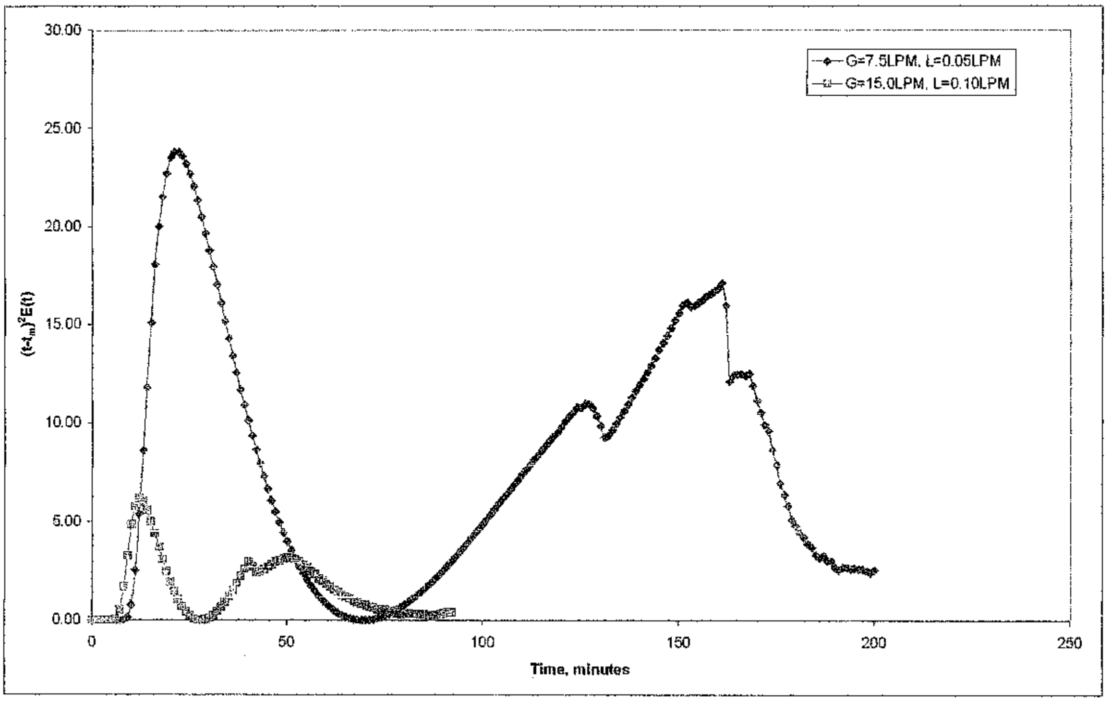

Figure A-3 Variance curve for gas/liquid ratio of 150

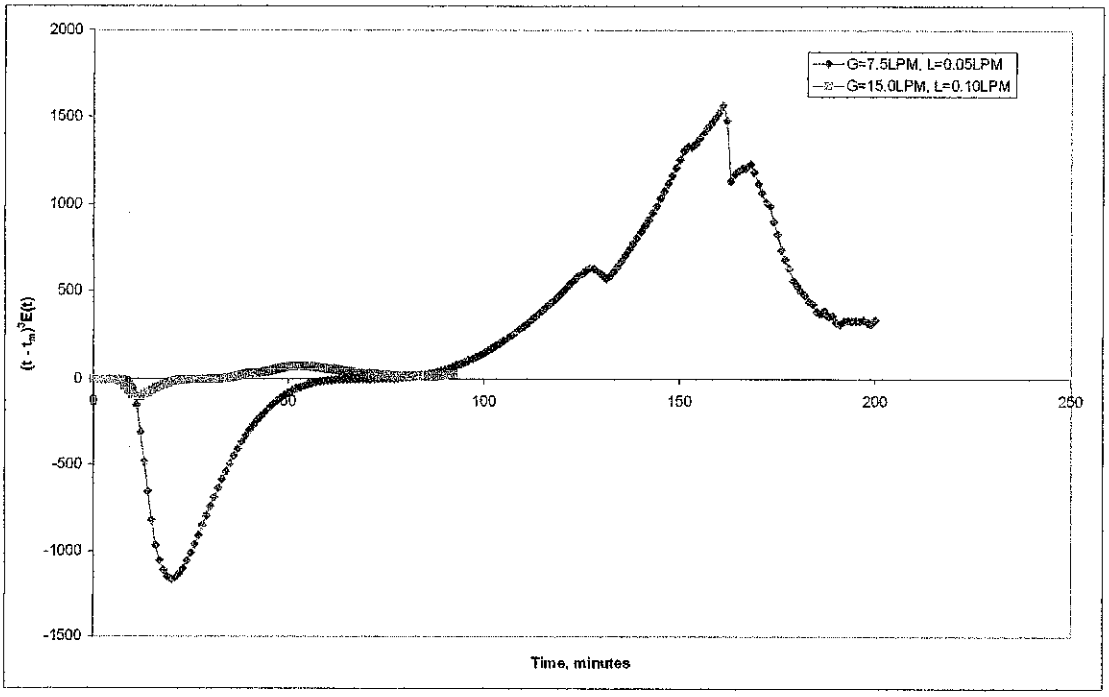

Figure A-4 Skewness curve for gas/liquid ratio of 150

APPENDIX B RTD Analysis for Gas/Liquid Ratio of 200

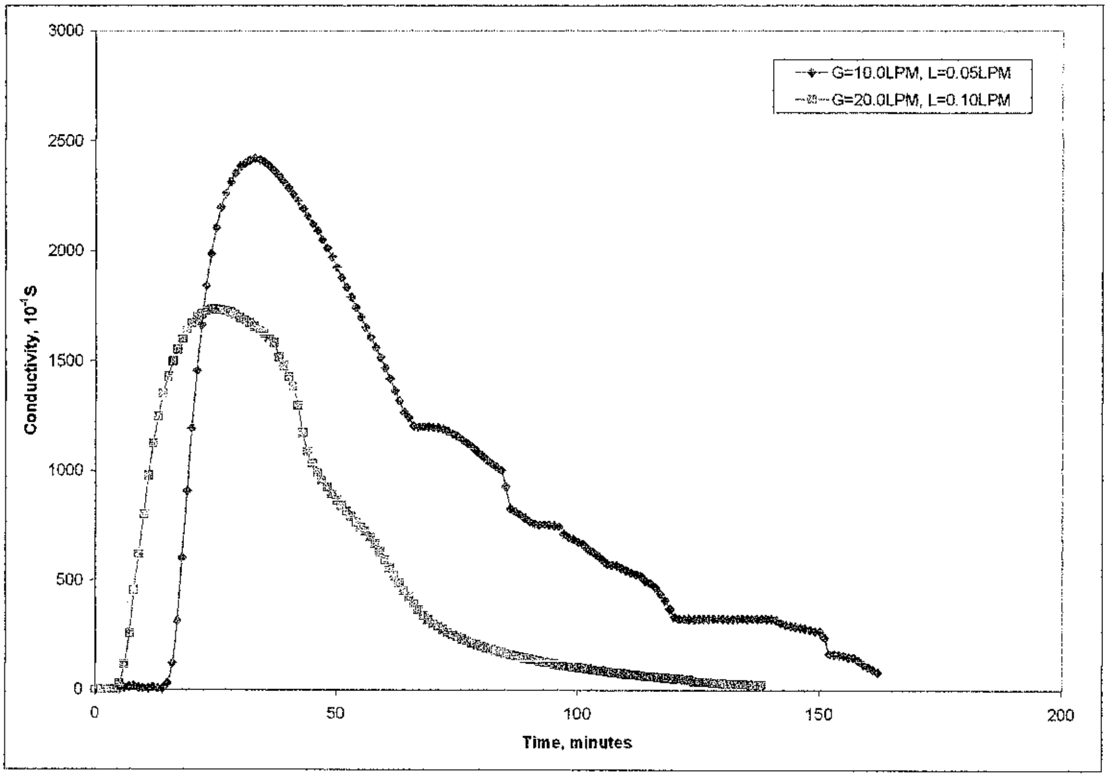

Figure B-1 Conductivity against time for gas/liquid ratio of 200

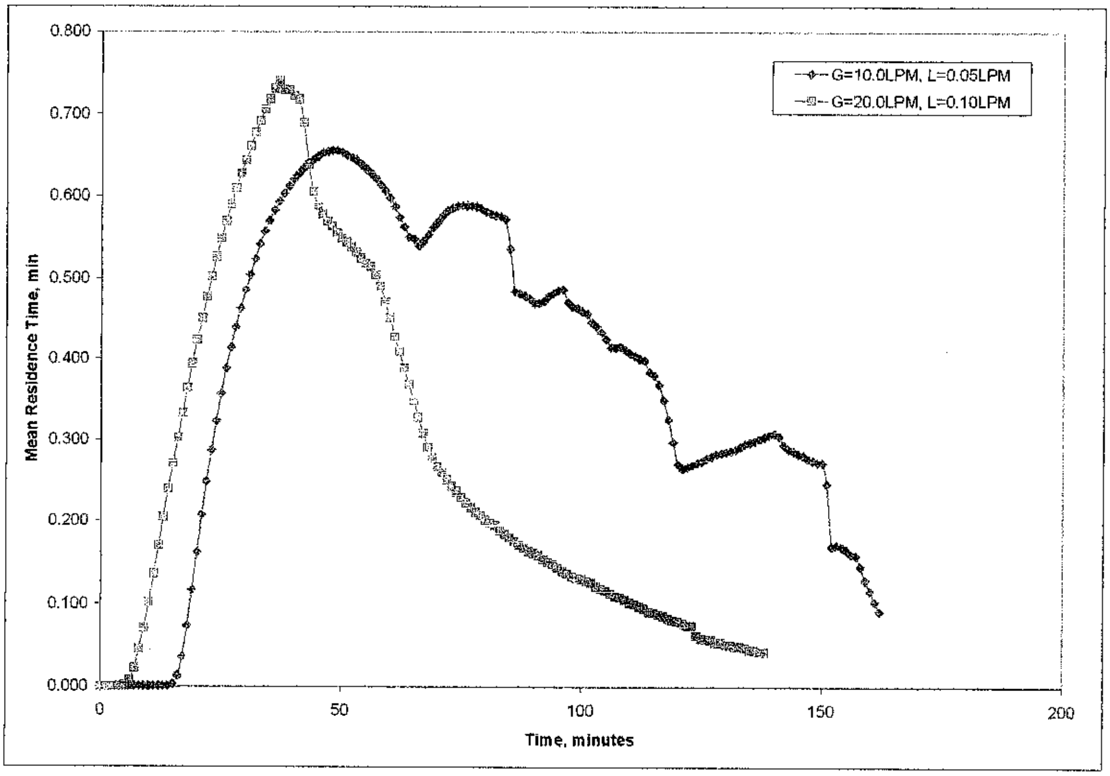

Figure B-2 Mean residence time curve for gas/liquid ratio of 200

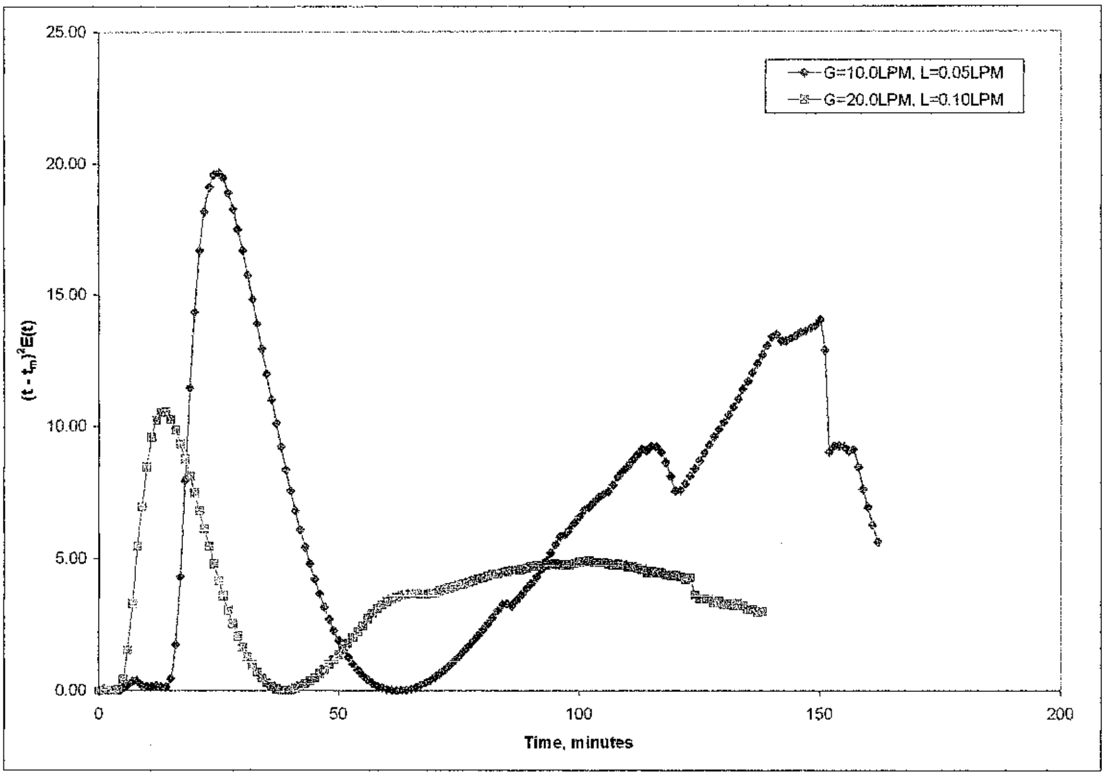

Figure B-3 Variance curve for gas/liquid ratio of 200

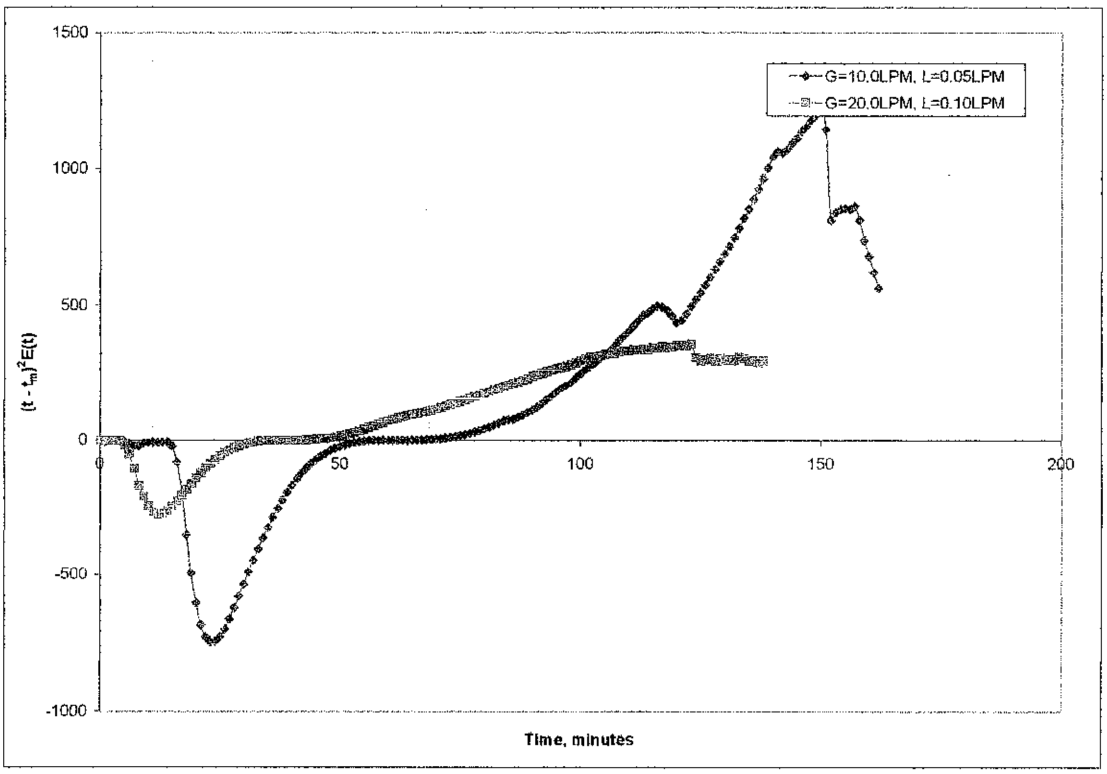

Figure B-4 Skewness curve for gas/liquid ratio of 200

APPENDIX C RTD Analysis for Gas/Liquid Ratio of 250

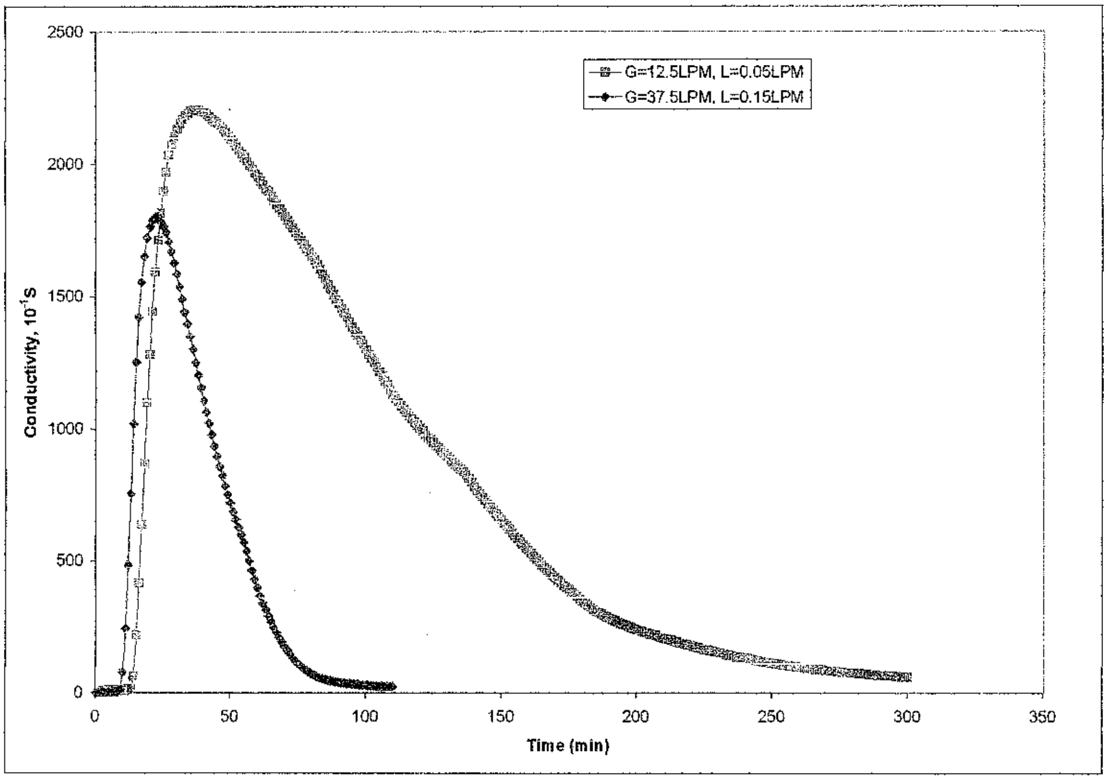

Figure C-1 Conductivity against time for gas/liquid ratio of 250