Cold model study of top blowing effect on dispersion in a channel reactor

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COLD MODEL STUDY
OF TOP BLOWING EFFECT ON DISPERSION
IN A CHANNEL REACTOR

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CANDIDATE'S DECLARATION

This is to certify that the material presented in this thesis is original. The work has been carried out in the laboratory of the Department of Materials Engineering, University of Wollongong, Australia and has not been previously submitted to any other institution or university for a higher degree.

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# TABLE OF CONTENTS

**CHAPTER I : INTRODUCTION**

**CHAPTER II : COUNTER CURRENT REACTOR THEORY**

2.1. Continuous Flow Reactors  
2.1.1. Tracer Technique  
2.1.2. Dispersion Coefficient  
2.2. Countercurrent Reactors  
2.2.1. Mass Transfer in a Metal-Slag System  
2.2.2. Mass Transfer and Dispersion in a Countercurrent Reactor  
2.3. Reactor Performance  
2.3.1. Flow Models  
2.3.2. RTD Evaluation  

**CHAPTER III : INDUSTRIAL COUNTERCURRENT FURNACES**

3.1. WORCRA  
3.1.1. Copper — U-Shape  
3.1.2. Steel-making — L-Shape  
3.1.3. Electric Steelmaking — Straight Shape  
3.2. QSL  
3.3. Vanyukov
3.4. Summary of the Main Channel Processes

3.5. Other Channel Processes

CHAPTER IV: PREVIOUS COLD MODEL STUDIES

CHAPTER V: EXPERIMENTAL

5.1. Experimental Design
  5.1.1. Materials
  5.1.2. Apparatus
    5.1.2.1. Channel Reactor
    5.1.2.2. Tracer Devices
    5.1.2.3. Gas Blowing Devices
    5.1.2.4. Ancillaries
  5.1.3. Equipment Set-up
  5.1.4. Procedure

5.2. Results
  5.2.1. Single Phase Experiments
  5.2.2. Counter Current Experiments

CHAPTER VI: DISCUSSION

6.1. Reproducibility of Results

6.2. Single Phase Experiments
  6.2.1. Mean Time
  6.2.2. Dispersion

6.3. Counter Current Experiments
  6.3.1. Mean Time
  6.3.2. Dispersion
6.3.3. Effect of Gas Flowrate on D/\mu L  
6.3.4. Effect of Bubbler Position on D/\mu L

6.4. Concluding Remarks

CHAPTER VII : CONCLUSIONS

REFERENCES
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ABSTRACT

The effect of flow regime, gas bubbling rate, and bubbler position on vessel dispersion number, $D/uL$, was studied in a room temperature, laboratory model of a channel reactor. The model reactor was a perspex channel, 200 cm long and 20 cm wide, in which tetrachloroethylene and water were used to simulate the metal and slag, respectively.

Two systems were investigated, namely single phase and two phase counter current systems. The liquid phase Reynolds numbers were in the range of 200 - 1700, gas bubbling rates were from 800 to 4000 cm$^3$/min., and the bubbler position was varied from 0 - 2 cm from the 'slag-metal' interface (for counter current case), and 0 - 4 cm from the water surface (for single phase case). The vessel dispersion number, $D/uL$, was obtained by an RTD technique using a thermal tracer and measuring the response by means of a thermistor of high sensitivity, wired to a data logger via a bridge circuit.

The results of the study have shown that:

1) For counter current experiments, as $Re$ increases, $D/uL$ increases and reaches a maximum at $Re$ around 900, then decreases. By statistical regression analysis, the average effect of bubbling rate and bubbler position was found as follows:
   
   i) slow slag, low $Re$ : $D/uL \propto Q^{0.22} P^{0.06}$
   
   ii) slow slag, high $Re$ : $D/uL \propto Q^{0.04} P^{0.11}$

   iii) fast slag, low $Re$ : $D/uL \propto Q^{0.27} P^{0.39}$

   iv) fast slag, high $Re$ : $D/uL \propto Q^{0.04} P^{0.11}$

2) For single phase experiments, there is no significant trend and effect appears for $D/uL$, as in counter current experiments, but the measured mean residence times of these experiments were in better agreement with the calculated values than those of other counter current experiments.
CHAPTER I
INTRODUCTION

Metal refining is one of the important stages in an extractive metallurgy process in producing high quality metal from ore. After the smelting process, the unwanted impurities that still remain in the liquid metal need to be removed and the metal refined. Refining processes usually involve chemical reactions and the transfer of impurities across the interface of molten metal and liquid slag phase. The impurities are oxidised from the melt forming an oxide, which is then transferred to the slag phase.

The interest in continuous refining operations was sparked from the consideration that continuous operations have many potential economical benefits over batch operations [Womer et.al, 1969; Kellogg and Diaz, 1992; Sohn, 1992]. Continuous processes have the potential to reduce capital and production costs by reducing handling and the site area required for the plant. Continuous processes also have the potential to achieve better performance because the liquid metal does not have to be preheated before the next stage as in many batch processes. This means that higher efficiency in the use of energy can be achieved and less fuel or power resources are used.

There have been a number of continuous channel type smelting processes, suggested and/or operated, in which the slag and metal phases move countercurrent to each other. The most generally known of these processes
are:

1) WORCRA (WORner, CRA) process which was operated in a pilot plant scale for both copper and iron [Baker and Worner, 1971; Worner and Andrews, 1973; Worner, 1992].

2) QSL (Queneau, Schumann, Lurgi) process, specially designed for non-ferrous metals, has been commercially operated to produce lead bullion from galena [Mazcek et.al., 1981; Queneau, 1989; Arthur et.al., 1992].

3) Vanyukov process, which has been commercially operated for non-ferrous metals in USSR since 1974 [Bystrov et.al., 1992; Hunt et.al., 1992; Bystrov, 1994].

4) Romelt (Romanets, melting) process, which is essentially an adaptation of the Vanyukov process for ironmaking, reportedly operated in USSR on a pilot plant scale until 1988 producing 30,000 tonnes of blast furnace grade pig iron [Weston and Thomson, 1996].

5) Ausmelt-SASE (South Australian Steel & Energy) project, a joint venture project which, as an Ausmelt reactor, has operated in commercial scale for non-ferrous metals, is being set up as an “Ausiron” reactor, in Whyalla, South Australia, to produce pig iron in a demonstration plant [Cui, 1996].

In addition to the above continuous processes as such, channel type furnaces have been employed as a part of continuous steelmaking process development starting with the iron ore smelting in the blast furnace to produce hot metal which is then fed to a channel type continuous refining furnace to produce steel. A recent example of this application of the channel furnace has been reported...
by Iwasaki et.al. [1989].

In continuous refining processes, in general, there are two ideal reactors with different flow characteristics, plug flow and mixed flow. In an ideal plug flow system, the flow is developed in one direction of flow, which results in a concentration gradient parallel to the direction of flow. In an ideal mixed flow system, components in the phase are dispersed in every direction resulting in a constant concentration throughout the system.

From the process design point of view, applying chemical performance theory, it is concluded that a plug flow continuous reactor behaves in an equivalent manner to a batch reactor in carrying out metallurgical reactions [Guthrie, 1993, p. 196; Szekely, 1967]. Guthrie [1993] also showed that a plug flow reactor is theoretically more efficient than a mixed flow reactor, especially in achieving over 90% refining efficiency, referring to removal percentage of the impurity.

In continuous metal slag reactors, there are also two main ways of contacting the metal and slag phases, cocurrent and countercurrent. Countercurrent refining is theoretically very promising for metal-slag systems, if plug flow characteristics can be achieved [Szekely, 1967]. He also showed that a plug flow reactor involving countercurrent slag-metal flow arrangement is considered to be preferable to a batch reactor and cocurrent pattern continuous reactor, from a purely reactor design point of view. However, there is some concern that back-mixing or dispersion caused by gas injection (i.e., oxygen blown into steel for refining) would prevent the continuous countercurrent plug flow ideal from being approached.

According to Nelson et.al. [1995], continuous countercurrent smelting processes
can be an effective substitute for the traditional methods if their dispersion characteristics are within acceptable limits. The purpose of the present study is, therefore, to examine the extent of dispersion with various top blowing conditions and liquid flow rates in a laboratory cold model continuous countercurrent reactor. It is considered that such a study might provide some useful insight into the dispersion phenomena occurring in the industrial prototype, quite apart from its possible application in the dispersion theory, in general.
CHAPTER II
COUNTER CURRENT REACTOR

THEORY

2.1. CONTINUOUS FLOW REACTORS

There are two main types of reactors used in the metallurgical industry, that is, batch and continuous. If there is no flow from the reactor during reaction and the reactants are held in the reactor while they react, the reactor is described as a 'batch' reactor (Figure 2.1a). The second type of the reactor is called a 'continuous' reactor, where the process takes place while the reactants enter continuously and the products formed by the reaction continuously flow out (Figure 2.1b).

Under incompressible steady state flow condition, the mean residence time of fluid or the average time spent by the fluid to flow through a continuous reactor is given by the following equation [Guthrie, 1993]:

$$ t = \frac{V}{Q} \text{ \..........(2.1)} $$

where $V$ = volume of the reactor (m$^3$), and
Q = volumetric flow rate of the fluid (m³/s).

(a) Batch reactors

(b) Continuous reactors

Figure 2.1. Some industrial metallurgical reactors [Guthrie, 1993].
There are two ideal steady-state flows that may happen in a continuous reactor, plug flow and mixed flow. These two kinds of flows are the ideal conditions that can be reached in a continuous reactor. These two ideal flows result in very different behaviour regarding the way how the material in the reactor is distributed. In reality, these ideal conditions are rare. Real processes always deviate from these two ideals.

The main characteristics of a plug flow reactor are that there are changes of composition along the length of the reactor, the fluid flows with a uniform velocity profile, and that no part of the fluid mixes with other parts ahead or behind [Levenspiel, 1993]. Another important characteristic is that the residence time in the reactor is the same for all elements of fluid. This is implied by the fact that the fluid flows in a uniform velocity profile, which means that every part of the entering fluid flows through the reactor with the same velocity and spends the same time inside the reactor before it flows out [Guthrie, 1993].

As a contrast, the ideal mixed flow is characterised with its uniform composition in every part of the reactor [Levenspiel, 1993]. When the reaction takes place, the composition within the reactor will change with time, while the composition in the plug flow reactor will change according to its position along the reactor. In a well-mixed flow reactor, all of the entering fluid will be instantaneously mixed with the fluid that is already inside. Some parts of the entering fluid flow out quicker while others remain longer inside the reactor [Guthrie, 1993]. It means that the residence time of some parts of the fluid will be widely different from the mean time.
Mixing or diffusion transverse to the direction of flow can happen in an ideal plug flow reactor, but not in the direction of the flow [Levenspiel, 1972]. When any mixing or diffusion exists in the direction of the flow, or if there is any fluid velocity gradient, the velocity profile will be fluctuated. Figure 2.2 shows the schematic diagram of this difference.

Figure 2.2. Velocity profile fluctuation of dispersed plug flow [Levenspiel, 1972].

This fluctuations affect the residence time, which will vary for each part of the fluid. This is a condition of flow dispersion, which is a condition that lies between plug flow and mixed flow. It is called longitudinal or axial dispersion, which can be considered as the flow characteristics of a certain reactor. Higher dispersion makes the flow characteristics get closer to mixed flow.
2.1.1. TRACER TECHNIQUE

The flow characteristics of a reactor can be determined using tracer measurements [Levenspiel, 1972], also called the stimulus-response technique. The tracer functions as a stimulus that disturbs the flow system, then measurement is done to provide information on how the flow responds to the stimulus and how the tracer is distributed in the flow system. The flow characteristics can be ascertained by analysing the information obtained from experiments.

There are numerous tracers that can be used, i.e. dye, salt solution, radioactive material and so forth. Dye tracer, especially, can also be used for visual observation purposes. Table 2.1 lists various kinds of tracers, property that is measured and the equipment that is used.

Table 2.1. Some tracers used in liquid systems [Wen and Fan, 1975].

<table>
<thead>
<tr>
<th>Tracer</th>
<th>Classification</th>
<th>Equipment</th>
</tr>
</thead>
<tbody>
<tr>
<td>NaCl</td>
<td>conductivity</td>
<td>thermal or electrical conductivity cells, recorder or potentiometer</td>
</tr>
<tr>
<td></td>
<td>salt solution</td>
<td></td>
</tr>
<tr>
<td>KCl</td>
<td>radioactivity</td>
<td>Geiger counter</td>
</tr>
<tr>
<td>Ba$^{140}$, Sb$^{124}$</td>
<td>radioactivity</td>
<td></td>
</tr>
<tr>
<td>HCl</td>
<td>concentration analysis</td>
<td>titration with AgNO$_3$</td>
</tr>
<tr>
<td>benzoic acid</td>
<td></td>
<td>titration with NaOH</td>
</tr>
<tr>
<td>KMnO$_4$</td>
<td>colour and light sensitive dyes</td>
<td>spectrophotometer</td>
</tr>
<tr>
<td>dye</td>
<td></td>
<td>colorimeter</td>
</tr>
</tbody>
</table>
Generally, any material can be chosen as a tracer as long as it can be detected or observed and it does not disturb the flow pattern [Levenspiel, 1972]. However, it is very important to carefully select an appropriate tracer for a specific system to give an accurate measurement [Wen and Fan, 1975].

One type of experimental technique that is commonly used is the pulse experiment. In this experiment, a tracer is introduced at the inlet over a very short time, to approximate an instant pulse. An “instant tracer” in an ideal plug flow reactor will provide a recorded concentration curve vs time as shown in Figure 2.3.

![Figure 2.3. Tracer concentration curve of an instant tracer input in an ideal reactor [Hill, 1977].](image)

In the outlet or along the reactor, the tracer is recorded again. For an ideal plug flow reactor, the curve will always give the same shape of response as the input
shape as shown in Figure 2.3. In reality, this is impossible, and the recorded concentration changes vs time curves, as shown in Figure 2.4, are far more realistic.

Figure 2.4. Shapes of tracer concentration curves that are probably formed along the reactor [Levenspiel, 1972].

The spreading of the curve is a sign that there was a dispersion of the tracer whilst it followed the liquid flow. The extent of dispersion represents the mixing that has occurred. In other words, it represents the deviation of the flow from the ideal plug flow.

2.1.2. DISPERSION COEFFICIENT

The extent of dispersion can be quantitatively measured by a dispersion coefficient, \( D \) (m\(^2\)/s) [Levenspiel, 1993]. In metallurgical systems, involving molten metals, its use has been summarised by Szekely and Themelis [1968].
Dispersion coefficient was also used recently by Sohn et.al. [1992] to correlate the effect of variables on dispersion in room temperature modelling of channel reactors.

For comparative studies, it is preferable to use a dimensionless value, $D/uL$ (where $u =$ fluid velocity and $L =$ length of the reactor) as a means to characterise the spreading. Figure 2.5 shows this comparison in graphical form. For comparison purposes, the concentration vs dimensionless time curve is used, with $\theta$ as the dimensionless time value.

If $D/uL$ is less than 0.01, the deviation from plug flow is considered to be small (curve 2.5b), while a value of $D/uL$ more than 0.01 is considered to be in the intermediate range to a large dispersion from plug flow (curve 2.5c and 2.5d). Notice that the $D/uL$ value for curve 2.5a in Figure 2.5 is zero, which indicates that there is no dispersion at all and the flow is an ideal plug flow.

As a comparison, the outlet curve of the tracer concentration for mixed flow is also given in Figure 2.5 (curve e). This curve, with a value of $D/uL = \infty$, points out an ideal mixed reactor, and it is said, qualitatively, that in completely mixed flow, the dispersion is maximum.
Figure 2.5. Deviation of tracer curve from ideal plug flow due to different extents of dispersion [Levenspiel, 1972].
2.2. COUNTERCURRENT REACTORS

In many metal refining processes, slag is used to remove impurities from molten metal. Ideally, these two liquids should be immiscible so they can be physically separated. When there is more than one phase involved in the processes, then the phase contacting patterns become one important factor to maximise the reaction to take place. There are several kinds of contacting patterns possible as shown in Figure 2.6.

The contacting patterns (f), (g) and ((h) in Figure 2.6 can only be applied for the mixed flow reactor, and the two patterns in (d) and (e) are for reactors with combination of plug flow and mixed flow. Only three kinds of contacting patterns in Figure 2.6 can happen in a plug flow reactor, namely countercurrent (a), cocurrent (b), and crosscurrent (c) [Levenspiel, 1972].

Countercurrent contacting pattern is present when two plug flows move along a vessel in opposite directions, while cocurrent pattern happens when these two plug flows move in the same direction. The crosscurrent pattern exists when the two flows cross each other. In regards to the flow patterns between slag and metal, there are two contacting patterns that must be considered for a continuous refining reactor, that is, countercurrent and cocurrent.

Figure 2.7 demonstrates temperature changes, that can also indicate composition changes one could expect along a reactor for an ideal cocurrent (Fig. 2.7a) and countercurrent (Fig. 2.7b) system [Levenspiel, 1972].
Figure 2.6. Ideal contacting patterns for two flowing phases [Levenspiel, 1972].
This phenomenon can be adapted to impurities transfer in metal-slag refining since the mass transfer characteristics are analogous to the heat transfer characteristics. It creates an analogous curve of composition changes, in terms of impurities concentration changes, in a continuous countercurrent metal-slag refining system, as shown in Figure 2.8.

The composition difference of the impurities between the two phases can be considered as the “driving force” of the reaction. The driving force in the cocurrent flow (Fig. 2.8a) becomes smaller along the reactor, while in the countercurrent flow (Fig. 2.8b) it is kept high enough to maintain the refining process to take place along the reactor which is the advantage that countercurrent contacting pattern gives compared with the cocurrent pattern.
2.2.1. **MASS TRANSFER IN A METAL-SLAG SYSTEM**

There are two important operations needed to take place in a metal refining process, which are [Deo and Boom, 1993]:

1) **Chemical reaction** in the metal phase with the help of oxygen that transform the impurity to its oxide form, for example the oxidation reaction of silicon in steel refining:

   \[
   [Si] + O_2(g) \rightarrow (SiO_2) \quad \text{.................(2.2)}
   \]

2) **Mass transfer**, a process of material transfer from one phase to another, which can be numerous for one chemical reaction. In relation with slag-metal

![Figure 2.8. Composition changes along the reactor length for metal-slag refining systems.](image)
case, an example would be the transfer of an impurity-oxide from the metal to the slag phase across an interface.

At the temperature of steelmaking, including the metal refining process, the chemical reactions are extremely fast [Deo and Boom, 1993]. Assuming that thermodynamic equilibrium immediately exists at the interface by this rapidly occurring chemical reaction, it can be expected that:

$$\frac{C_{m,eq}}{C_{s,eq}} = \frac{C_{m,i}}{C_{s,i}} = m \quad \ldots \ldots \ldots \ldots (2.3)$$

where $C_{m,eq} =$ concentration of impurity in metal at equilibrium (mol/m$^3$),

$C_{s,eq} =$ concentration of impurity in slag at equilibrium (mol/m$^3$),

$C_{m,i} =$ concentration of impurity in metal at the interface (mol/m$^3$),

$C_{s,i} =$ concentration of impurity in slag at the interface (mol/m$^3$), and

$m =$ equilibrium partition constant.

The equilibrium partition constant, $m$, is sometimes called partition or distribution coefficient, which is the ratio of impurities concentration between metal and slag phases.

According to Whitman's two film theory [Deo and Boom, 1993; Guthrie, 1992], the rate of interphase mass transfer can be expressed by:

$$N = k_{ov} (C_m - C_s) \quad \ldots \ldots \ldots \ldots (2.4)$$

where $N =$ mass flux of the impurity (mol/m$^2$.s)
**2.2.2. MASS TRANSFER AND DISPERSION IN A COUNTERCURRENT REACTOR**

The relationship between mass transfer and dispersion can be described in one equation each, for metal and for slag, as follows [Engh et. al., 1971]:

\[
- \frac{1}{P_m} \frac{d^2 C_m}{dx^2} + \frac{dC_m}{dx} + k \left( C_m - \frac{C_s}{m} \right) = 0
\]

\[
\left(2.6\right)
\]

\[
\frac{1}{P_s} \frac{d^2 C_s}{dx^2} + \frac{dC_s}{dx} + kfm \left( C_m - \frac{C_s}{m} \right) = 0
\]

\[
\left(2.7\right)
\]

where \( P \) = inverse Peclet number in metal phase  = \( \left( \frac{D}{uL} \right)_m \)
\[
\frac{1}{P_s} = \text{inverse Peclet number in slag phase} = \left( \frac{D}{uL} \right)_s
\]
\[x = \text{distance along the reactor}\]
\[k = \text{mass transfer coefficient}\]
\[f = \frac{Q_m}{Q_s m}\]

\[Q_m = \text{volumetric flowrate of liquid metal}\]
\[Q_s = \text{volumetric flowrate of liquid slag}\]

The first term of equations (2.6) and (2.7) represents the dispersion value, where \(1/P\) (inverse Peclet number) = \(D/uL\) (dimensionless value that characterises the spreading/dispersion rate). The second term of equations (2.6) and (2.7) represents the concentration gradient, \(dC/dx\), and the third term of the equations (2.6) and (2.7), i.e. \(K(C_i - C_{i,eq})\), represents the mass transfer term.

2.3. REACTOR PERFORMANCE

One basic property of a continuous process is the Residence Time Distribution (RTD). Examination of this property with pulse experiments can allow the performance of the reactor and its flow behaviour to be determined. The actual mean time of any RTD measurement can be calculated from appropriate
equations, whilst the theoretical mean residence time ($t_{\text{theory}}$) is obtained from a simple equation, i.e.:

$$t_{\text{theory}} = \frac{V}{v} \quad (2.9)$$

where $V$ = volume of fluid in the reactor

$v$ = volumetric rate of fluid flow

Using qualitative comparison between an RTD$_{\text{actual}}$ curve and the RTD$_{\text{theory}}$ curves, the reactor performance can be improved by correcting the problem that was diagnosed. The RTD$_{\text{actual}}$ measurement itself is also important to determine the flow behaviour by calculating various parameters, such as stagnant volume, turnover time and the like, as well as the dispersion value ($D/udL$).

The utility of the $D/udL$ value is that it characterises the dispersion of a reactor and hence its departure from the ideality of heat and mass transfer and the chemical reaction, or reactions, taking place there. For all manner of packed beds and full flow tubular reactors, as well as flow in coiled pipes and flow in channels, theory shows that [Levenspiel, 1972]:

$$\frac{D}{uL} = \frac{D}{ud} \times \frac{d}{L} \quad (2.10)$$

where $D/ud$ = intensity of dispersion and $d/L$ is the characteristic geometric factor [Levenspiel, 1972].
The intensity of dispersion depends on [Levenspiel, 1972]:

1) fluid properties as given by Schmidt Number $(\mu/\rho D)$, and

2) flow regime as characterised by Reynolds Number $(du/\nu)$.

In the above the characteristic dimension for pipes and channels is the tube diameter or its equivalent, such as $4 \, R_H$, and the length $L$. For packed beds, particle size is the characteristic dimension together with the bed length, $L$. Additionally, bed porosity is also usually involved since generally superficial velocity is used in engineering calculations and design, instead of the actual or pore velocity. Figures 2.9 and 2.10 show the relationship between intensity of dispersion and Reynolds Number for both laminar and turbulent flow in packed beds and pipes, respectively.

![Figure 2.9. Relationship between intensity of dispersion and Reynolds Number in packed beds for both laminar and turbulent flow [Levenspiel, 1972].](image)
Figures 2.10. Relationship between intensity of dispersion and Reynolds Number in pipes for both laminar and turbulent flow [Levenspiel, 1972].
2.3.1. FLOW MODELS

Models can be made to represent the flow pattern in a particular reactor, as well as for diagnostic purposes, or even for scale up. These models view the process, such as for example a continuous refining furnace, as consisting of a number of regions or volumes each of which being otherwise described by a single parameter.

Examples, include models consisting of [Levenspiel, 1972]:

1) Active and stagnant regions in which the former may consist of plug and/or mixed flow volumes.

2) Parallel flow regions in which one or the other, or both, may be plug or mixed flow volumes. The by-pass flow model may be considered as an extreme example of a parallel flow model in which the by-pass volume is very much smaller than the other volumes.

3) Recirculating flow regions in which part of the exit flow is recycled back to the inlet. Although this model, with its recycle ratio parameter, may appear to be a single parameter model and not a multi-parameter model, it should be borne in mind that the other regions, including the recycle, may behave as plug flow or mixed flow or dispersed flow volumes.

Apart from the dispersion model, based on the value of D/uL, and, as can be seen from equation (2.11),
\[ \frac{2}{\theta} = 2\left( \frac{D}{uL} \right) = \frac{1}{N} \] ..........................(2.11),

its equivalent single parameter tanks-in-series model, there are also the so-called multi-parameter models which have been extensively used in the modelling of metallurgical reactors [Szekely and Themelis, 1968].

In equation (2.11), N is the number of tanks in series and it is obvious that the larger is N the more closely the flow approaches the ideals of plug flow. Thus, for a single mixed flow volume the RTD is exponential and its D/uL = \infty, as shown in Figure 2.5. The curve for D/uL = 0.2 in Figure 2.5, in which mixed flow behaviour is still evident by the exponential tail, corresponds to N = 2.5, and so forth.

Metallurgical reactors have been modelled by the tanks-in-series models [Szekely and Themelis, 1968], as well as by one or the other multiparameter model, depicted in Figure 2.11, as such, or as a combination. The latter method was adopted by Sohn et.al. [1992] to model their cold, laboratory scale channel reactor.
Figure 2.11. Some variations of multiparameter model [Levenspiel, 1993].
2.3.2. RTD EVALUATION

Mathematical evaluation of RTD from an experimental response curve is generally straightforward. The first parameter usually calculated is $t$ and this is done according to equation (2.12) as such, or its incremental version [Levenspiel, 1972].

$$
\bar{t} = \frac{\int_0^\infty t \, C \, dt}{\int_0^\infty C \, dt}
$$

(2.12)

Next, $E_\theta$ or $E_t$ is calculated in a manner depending on the nature of the input, i.e. pulse, step etc., and the shape of the response curve. For the often encountered "dispersion type" response, i.e. a single peaked, more or less symmetrical curve, the variance is calculated by equation (2.13):

$$
\sigma^2 = \frac{\int_0^\infty t^2 \, C \, dt}{\int_0^\infty C \, dt} - \bar{t}^2
$$

(2.13)

Another method of calculating $\sigma^2$ that may sometimes also be used, is the so-called 61% height method [Levenspiel, 1972], and is based on measurement of the width of the curve at its 61% height ($W_{61}$) and the distance of the peak of the curve from the origin ($d$). Then $\sigma^2$ is calculated by equation (2.14):

$$
\sigma^2 = \frac{1}{4} \left( \frac{W_{61}}{d} \right)^2
$$

(2.14)
The value of $2$, obtained from equation (2.13) or (2.14), can then be used to obtain $\frac{D}{uL}$ from equation (2.15):

$$\frac{2}{\theta} = \frac{\sigma^2}{t^2} = 2\left(\frac{D}{uL}\right) \quad \text{...............................................}(2.15)$$

Evaluation of $\frac{D}{uL}$ from $2$ obtained from equation (2.13) is the usual method used. Its evaluation from $\frac{2}{\theta}$ obtained from equation (2.14) is generally considered not as accurate. The use of the 61% method is, however, useful as a check if the response curve is excessively skewed or has a long tail, because it only measures the principal properties of the response curve.

Generally, evaluation of a proper RTD from an experimental response curve depends on the boundary conditions, i.e. the way the tracer is introduced and measured. Additionally, the properties of the tracer and those of the flowing fluid, as well as the intricacies of the flow pattern itself, especially if more than one fluid is involved, make it impossible to rely on just the response curve for $\frac{D}{uL}$ without also considering the input curve. If the latter is available, then theory shows that for a "sloppy" input of tracer [Levenspiel, 1993]:

$$\Delta \sigma^2 = \sigma^2_{\text{output}} - \sigma^2_{\text{input}} \quad \text{..........................}(2.16)$$

where $\Delta \sigma^2$ is the difference of the variances of the output and input curves and is the proper method to evaluate $\frac{D}{uL}$ when the input is not ideal.

Quite often the response curve exhibits a very long tail which may be impossible to be sure of whether it is part of the flow in the reactor as such, namely,
existence of a relatively stagnant volume, or whether it is due to the tracer being adsorbed on the reactor walls/internals, or partly settling at the bottom or rising to the surface due to density differences and so on. The only way to be sure is to employ a radioactive isotope of the actual flowing fluid as a tracer, but this is either impossible if no radioactive isotope exists, or impractical if it has a very short half life and the RTD measurements are being done far away from the atomic reactor where the isotope can be produced [Levenspiel, 1993].

Often, some clues of what is going on in the reactor can be obtained by observing the shape of the response curve and comparing \( \tau_{\text{theory}} \), as calculated by equation (2.9), with \( \tau_{\text{actual}} \) as a result of an experiment. Some possibilities can be found [Levenspiel, 1993], as it is seen in Figure 2.12 and Figure 2.13.

![Figure 2.12. Some possible RTD curves for plug flow [Levenspiel, 1993].](image)
Figure 2.12 illustrates some possibilities that can occur for a plug flow reactor. If $t_{\text{theory}}$ is approximately the same as $t_{\text{actual}}$ (Figure 2.12a), it means that a reasonably good flow, in the sense that the full volume is active, was achieved. If $t_{\text{actual}}$ is less than $t_{\text{theory}}$ (Figure 2.12b), this indicates the presence of stagnant region in the reactor. Another type of curve is that depicted in Figure 2.12c, showing multiple decaying peaks at regular intervals which indicates strong internal recirculation. If double peaks exist (Figure 2.12d), parallel paths or flow channelling in the reactor are indicated, parts of the flow being faster than others. The last curve (Figure 2.11e) shows a late curve ($t_{\text{actual}}$ is more than $t_{\text{theory}}$), which is theoretically not possible because the material balance around the system is contravened, indicating some error in the experimental measurements or non-steady state characteristics, or tracer is disappearing by reaction or transfer [Levenspiel, 1993].

Figure 2.13 illustrates some possible curves for a mixed flow reactor. A reasonably good flow, in the sense of its approach to ideality, is indicated when the curve is like that in Figure 2.13a, it is exponentially decayed and the $t_{\text{actual}}$ is more or less the same with $t_{\text{theory}}$. Figure 2.13b shows a time lag of the peak which is a characteristic of the existence of a plug flow in series with mixed flow. Figure 2.13c shows a slow internal recirculation which can be caused by sluggish or slow turnover of fluid, or inadequate mixing. Figure 2.13d, e, f, respectively, show an early curve scenario (means that there is stagnant fluid), late curve example, and sharp early peak (indicating shortcircuit from inlet to outlet).
In addition to the clues provided by the shape of the response curves as discussed above, the knowledge of the process itself and the physical equipment involved is essential in making judgement as to reality and in interpreting the response curves with more confidence.

Figure 2.13. Some possible RTD curves for mixed flow [Levenspiel, 1993].
CHAPTER III

INDUSTRIAL COUNTERCURRENT FURNACES

There are three well-known processes that have been developed to be used in a continuous metallurgical process to produce metal from its ore with countercurrent flow. The first process was the WORCRA, which was invented by Prof. H. K. Worner whilst at Conzinc Riotinto of Australia Ltd. (WORner and CRA). The second one is called QSL, the initials standing for the inventors’ names, Paul E. Queneau and R. Schuhmann and LURGI AG, the company that has been acting as the process developer. The third process is the Vanyukov process, named after the inventor.

Even though these three furnaces have many differences, there are four main principles which apply to all three, namely:

1. metal and slag flow countercurrently,
2. there is gas injection through tuyeres or lances,
3. they are continuous processes,
4. metal and slag flow under gravity force.

The WORCRA process was initially conceived in the early 1960's for steelmaking, although it was trialed extensively for non-ferrous bath smelting. The QSL process was developed during the 1980's for non-ferrous smelting,
especially in processing galena to make lead bullion. The Vanyukov process was developed in the late 1960's for producing copper matte, but since 1984 the process has also been developed for iron ore.

3.1. WORCRA

The WORCRA process was designed for the first time to treat copper concentrate and produce Cu-metal instead of Cu-matte like conventional furnaces. A pilot plant was built in Cockle Creek, NSW, Australia, in 1963 and the results [Worner, 1968] were used to build a semi-commercial plant in Port Kembla, NSW, Australia, in 1968. Both plants had a U-shaped horizontal geometry.

Together with the development of the copper smelting-converting process, WORCRA continuous principle was applied to the steelmaking process in a pilot plant in Cockle Creek, NSW, Australia using an L-shaped horizontal furnace. The plant was only designed to do the refining step of the steelmaking process as it used a molten pig iron feed. The furnace was further developed in Lulea, Sweden, where a straight shape horizontal furnace was used. This time, the process combined two conventional steelmaking batch processes, smelting and refining, in one continuous furnace.

Although different shapes were being used for particular furnace, the process itself had the same zonalised concept [Worner, 1992]: smelting took place in a more or less central zone; oxidation of the 'internal fuels' was achieved in the
directly-connected converting (or refining) zone where the majority of the flux was added. By the use of an appropriately cooled slag dam or end wall, the slag was forced to flow generally countercurrent to the matte or metal and be 'cleaned' by the incoming sulphide concentrates (or coal in the application to oxidic ores such as iron, manganese and chromium) in the vigorously mixed smelting zone before flowing on through the relatively quiescent slag settling zone for continuous tapping. The metal was tapped continuously from a deeper taphole at the opposite end.

3.1.1. COPPER — U-SHAPE

This U-shape furnace was used in the semi-commercial plant at Port Kembla, Australia, for continuous copper-making trials. The main results were [Worner et.al., 1972]: blister copper (98.75 % Cu) could be directly and continuously produced from concentrates for refining to anodes; low copper content slag (0.56 % Cu in average); and high-$\text{SO}_2$ (5-8 % $\text{SO}_2$) gas, and a disposable slag. The slag produced did not need to be recycled to the reactor, as in the conventional copper converting processes.

Figure 3.1 illustrates the basic zonalised concept that was utilised in the copper smelting-converting process. The process started in the smelting zone (area B) where concentrate entered with certain feedrate. Here, the concentrate was smelted forming fine droplets that had greater weight than the oxide compound which allowed separation between matte and slag by settling.
The smelting took place in a 3.05 m diameter bowl, connected to the downward-sloping converting zone, on one side, and to an upward-sloping slag-converting zone, on the other side. The converting zone was 1.83 m wide at the liquid line and 8.53 m long. The slag-cleaning zone was 2.44 m wide and 9.45 m long.

Fig. 3.1. View of U-shaped WORCRA furnace built at Port Kembla, Australia [Womer et.al., 1970].

Because of the downslope along B-A (converting zone), the matte flowed through this area. During its journey in this zone, the matte reacted with the oxygen from air lancing to produce blister copper that flowed through an underpass in the end riser wall to the Cu well. Bath depth at this riser wall was 1.22 m. The slag, being retained by the riser wall, flowed countercurrently with
matte through A-B-C-D (from converting zone to the slag-cleaning zone) to the slag well. At the start side of the slag-cleaning zone (area C), there was concentrate input to help slag washing. This concentrate reacted with slag from the smelting zone to form matte and an impurities enriched slag. Matte that was generated along C-D (slag-cleaning zone) flowed back by gravity, caused by the downward-sloping to the smelting zone.

At the upper side of the furnace there was a combustion-free space zone which allowed SO$_2$ produced from the processing stage to flow through and go to the out-gas pipe. It was 2.74 m above liquid level in the converting zone and 1.37 m high in the slag-cleaning zone. This size was considered enough for gas injection intensity in converting zone of 0.10 m$^3$ of air per m$^2$ surface area per second.

There were two kinds of matte generated at different zones in this furnace. One was formed in the smelting zone, where the process “started” and the other was formed in the slag-cleaning zone in the “end” of the process, but the second matte would flow to the smelting zone and be mixed with the smelting zone matte. This mixing matte flowed through the converting zone to be converted to blister copper.

The furnace was designed to smelt more than 70,000 kg dry concentrates per day. Data were collected from three campaigns in 23 months of continuous operation. The concentrates that have been used in this semi-commercial plant came from two different places, Cobar in New South Wales and Mt. Lyell in Tasmania. The compositions are shown in Table 3.1.
Table 3.1. Composition of concentrates used in WORCRA semi commercial plant at Port Kembla, Australia [Womer et.al., 1972].

<table>
<thead>
<tr>
<th>Element</th>
<th>Cobar, %</th>
<th>Mt Lyell, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cu</td>
<td>23.7</td>
<td>26.0</td>
</tr>
<tr>
<td>Fe</td>
<td>30.6</td>
<td>30.5</td>
</tr>
<tr>
<td>S</td>
<td>32.9</td>
<td>35.0</td>
</tr>
<tr>
<td>SiO₂</td>
<td>3.7</td>
<td>3.6</td>
</tr>
<tr>
<td>Al₂O₃</td>
<td>0.7</td>
<td>0.1</td>
</tr>
<tr>
<td>CaO</td>
<td>0.1</td>
<td>0.1</td>
</tr>
<tr>
<td>Zn</td>
<td>5.0</td>
<td>0.1</td>
</tr>
<tr>
<td>Pb</td>
<td>2.2</td>
<td>0.1</td>
</tr>
</tbody>
</table>

From a series of data [Womer et.al., 1972] over a certain 30-day period that was considered as a representative period of relatively steady operation, the following results were found. A total concentrate feed of 2174.3 tonnes contained 514.1 tonnes of copper, of which the total copper tapped was 509.6 tonnes. This corresponds to a copper recovery of 97.8 %. The average copper analysis for campaign no. 3 when Cobar concentrates were being smelted was: Cu 98.75 %, S 0.74 %, Fe 0.15 %, Pb 0.06 %, and Zn 0.01 %. The average metal flowrate was approximately 0.19 kg/s.

During campaign no. 3, slag tapped from the furnace for a continuous 30-day period showed that the slag contained an average value of 0.56 % Cu. Analysis was done of samples taken from granulated slag. It was also found
that low slag losses were achieved with silica in the range of 28 - 32 %. During this campaign, the average slag composition was 29.3 % SiO₂ and 7.5 % CaO [Worner et.al., 1972].

### 3.1.2. STEEL-MAKING ---- L-SHAPE

After the experience with non-ferrous metals, the possibility of using WORCRA furnace in the area of steelmaking was examined. It was visualised that iron could be made into steel using continuous oxygen lancing with metal and slag flowing countercurrently.

The WORCRA basic zonalised concept was utilised in the L-shaped furnace built at Cockle Creek in 1965, the difference being in the shape of the furnace. In the copper smelting-converting process, a U-shaped furnace had been used, but an L-shaped furnace, as shown in Figure 3.2, was chosen for the first WORCRA steelmaking.

The furnace had three zones. At the central zone, there was a bowl-shaped area called the feed zone. This bowl had a refining zone branch and a slag clean-up branch connected at right angles to each other. In spite of huge differences between steelmaking and copper-making process, the route of metal and slag flow in this furnace is almost the same with the WORCRA furnace that was used for copper smelting-converting. The feed was received in the bowl area, whilst in the U-shaped Copper furnace there were two feed
points, one in the central zone and the other before the slag-cleaning zone. There was also another material inlet in this furnace, just for lime addition.

Fig. 3.2. View of L-shaped WORCRA Steel-Making (W.S.M.) [Worner, 1992].

Hot metal feed, coming from a cupola, entered the bowl, an area in which the fluid slag and metal generally flowed cocurrently. Because of the downward sloping along the refining zone, the metal flowed through it and during this
journey, steel was formed as a result of reaction between metal with the oxygen which was sequentially jetted from the top. At the end of the refining zone, there was a dam which forced the slag to flow back along the refining zone to the bowl, while steel flowed through the underpass at the bottom of the dam to the steel outlet. The slag generated in the refining zone flowed countercurrently to the metal, through the bowl and the slag clean-up zone to the slag outlet.

The furnace was designed to handle up to 5 tonnes per hour feed, but the cupola furnace did not allow a rate greater than 3.5 tonnes per hour. The trial runs were performed 29 times, giving some promising results that was claimed as a proof of the potential of this technology [Baker and Womer, 1971].

The results from trials 21 and 23 were considered by Womer et.al. [1969] as indicative of the process. The hot metal feed composition was: 4.1 % C, 1.2 % Si, 0.76 % Mn, 0.095 % P, and 0.065 % S. This was fed into the bowl at an average rate of around 3.3 tonnes per hour. With a refining time of approximately 186 minutes, the steel had an average composition of 0.52 % C, 0.04% Si, 0.16 % Mn, 0.014 % P, 0.017 % S. This corresponds to 87 % C removal, 97 % Si, 79 % Mn, 86 % P, and 74%S.
3.1.3. ELECTRIC STEELMAKING ---- STRAIGHT SHAPE

Integrated continuous melting-refining steelmaking with countercurrent flow was tested by Worner and co-workers in Lulea, Sweden. The project ran ten trials with up to 30 hours of continuous process, and the result confirmed its potential for achieving high productivity, high refining efficiency, and high Fe recovery [Baker and Worner, 1971]. As it can be seen in Figure 3.3, the whole WORCRA melting-refining process is mostly the same with the WORCRA steel-refining process. The only exception is the presence of smelting reaction in the bowl zone to melt the cold concentrate feed that was used, which did not exist in the Cockle Creek furnace that used hot metal as the feed.

Fig. 3.3. View of straight-shaped Electric WORCRA Steel-Making [Worner et.al., 1969].
Two kinds of feed were used. Most of the trials used granulated pig iron, the two last trials used Wyberg sponge iron, and some trials blended those two. The average analysis of the pig iron is shown in Table 3.2. The Wyberg sponge iron has composition: total Fe 87.5 %, metallic Fe 76.6 %, oxygen 4.7 %, C 0.67 %, Mn 0.95 %, P and S 0.01 % each, SiO₂ 3.68 %, MgO 1.12 %, Al₂O₃ 1 %, and CaO 0.35 %.

<table>
<thead>
<tr>
<th>wt. %</th>
<th>C</th>
<th>Si</th>
<th>Mn</th>
<th>P</th>
<th>S</th>
</tr>
</thead>
<tbody>
<tr>
<td>Low-Si</td>
<td>4.23</td>
<td>0.48</td>
<td>0.67</td>
<td>0.08</td>
<td>0.025</td>
</tr>
<tr>
<td>High-Si</td>
<td>3.45</td>
<td>1.87</td>
<td>0.32</td>
<td>0.16</td>
<td>0.102</td>
</tr>
</tbody>
</table>

3.2. QSL

The QSL process (Kellogg and Diaz, 1992; Arthur et. al., 1992) was developed as a continuous lead smelting process to produce crude lead bullion directly. The process is similar to the WORCRA but has its own unique features. A schematic of the process, as operated in Stolberg-Germany, is shown in
Figure 3.4. The 32 m long cylindrical vessel furnace has two major zones, oxidation zone and slag-reduction zone. The process started in the oxidation zone. In this first zone, the feed was charged from the top, oxygen was jetted through the 3 bottom tuyeres, and the SO\textsubscript{2} rich gas outlet was placed. This zone has a diameter of 2.74 m and is 11.6 m long.

![Diagram of QSL Furnace](image)

Fig. 3.4. View of vertical section of QSL Furnace [Queneau, 1989].

Inside the furnace, a molten bath of lead oxide dissolves the charge. With the presence of oxygen that was jetted through the bottom, chemical partial oxidation reaction happened to produce lead bullion. As for copper in the WORCRA process, molten metal, being the heaviest fluid in this furnace, flows to the right end under the effect of gravity.
Lead rich slag from this zone, called primary slag, goes left through the bottom of the divider wall set up between two main zones of the furnace and enters the second zone. In this slag-reduction zone, which has a diameter of 2.25 m and is 20.3 m long, the primary slag meets pulverized coal and oxygen injected by 5 tuyeres from the bottom of the furnace, reducing lead oxide to form lead bullion.

The slag from this zone is very low in lead content because the lead oxide is gradually reduced while it flows through the second zone of the furnace. This slag flows from the outlet set up on the left of the furnace and positioned about the surface of the slag to ensure that the outflow slag is the lowest lead-content slag. The lead bullion formed in the slag-reduction zone will flow back to the oxidation zone directly towards the lead bullion outlet.

Figure 3.4 shows a schematic picture of the furnace. In reality, there are multiple slag-reduction zones to ensure that the tapped slag does not contain significant amounts of lead oxide. The Berzelius Stolberg QSL furnace for example, has three slag-reduction zones, and the whole furnace is slightly sloped about 0.5 % which can be rotated through 90 degrees when operation is interrupted [Mazcek, et.al., 1981]. The slag in the first and second reduction zone is not tapped, but continuously flows to the next zone, because it still contains a considerable amount of lead oxide. The final slag is tapped at the end of the third slag-reduction zone [Kellogg and Diaz, 1992].

Green pellet charges have been used in this type of furnace. These pellets are an agglomeration of mixed lead concentrate, fluxes, and recycled flue dust.
The lead concentrate and fluxes blend are fed at 20.75 tonnes per hour, combine with 5.5 tonnes per hour recycled dust, both with 50 % Pb-content. Other material is lignite coke, which is fed into the oxidation zone at a feedrate of 0.75 tonnes per hour and in the reduction zone at 0.5 tonnes per hour.

The total bath volume of 31 m³, of which approximately 18 m³ is in the oxidation zone and 13 m³ in the reduction zone. This kind of charge yields three outputs. First output is the lead bullion that is tapped at the end of the oxidation zone at a flowrate of 9 tonnes per hour and contains more than 98 % Pb. At the end of reduction process slag is tapped at 5.5 tonnes per hour flowrate and contains 2 % of lead. The third output is a high SO₂ concentration off-gas that contains flue dust. This output passes through a waste heat boiler for heat recovery and a precipitator for de-dusting before flowing to the sulfur recovery plant. The precipitated flue dust, containing 50 % Pb, is recirculated to the furnace [Queneau, 1989].

3.3. VANYUKOV

This furnace was first proposed and tested in bench scale experiments in 1949, in Russia, and has been improved since that time, in laboratories, pilot plants, and industrial plants. There are three pilot plants at Ryazan, Ust-Kamenogorsk, and Lipetsk. There are also six currently production scale furnaces in operation, four at Norilsk and two at Balkhash which treat more than 1,500,000 tonnes of Copper and Copper-Nickel concentrates per year in
total amount and produce more than 350,000 tonnes of Copper [Bystrov et. al., 1992].

Some advantages have been proved in practice: good sulphide ores and minerals smelting result, high efficiency of fuel, and low impact on environment. It has been extended in application to oxidised ores and non-metallic materials and to applications where low quality coal may be burned cleanly for the purpose of power generation.

The furnace has a rectangular geometry, with tuyeres located along the side wall. Cold solid particles of material are charged into a bath of melted agitated hot slag. These charges are wetted and heated by melted slag, but there is no reaction until they approach the temperature of the slag.

Sulphide compounds are melted, meet the oxygen that is jetted through tuyeres, forming drops of matte and metal in the slag. All the oxides, coal and other material, because of the intense agitation, are dissolved in the slag or burned with oxygen in the slag. The denser matte liquid flows towards the bottom to the metal outlet, while the lighter fluid is retained by a wall that forces the fluid to flow back to another tapping hole. Indeed, this is the essence of how the countercurrent flow is created.

The first industrial furnace was constructed in 1974 at Balkhash Copper Smelter. It had a cross section at the tuyere level of 4.2 sqm and was designed as a rectangular shaft about 7 m in height. The central part of the shaft at a level of 1 to 5 m was constructed with a water cooled copper jacket to replace refractory. At that time more than 10,000 tonnes of charge had been
processed and the specific productivity of 50 to 70 tonnes of charge per 1 m² per day of bath area was achieved [Bystrov et al., 1992].

This furnace has been tried several times for various kind of concentrates, generally for copper sulphide concentrates, and Bystrov [1994] reports that the process has given satisfying results. In particular, there has been low dust escape from the process, with only 0.5 - 1.5 % of the feed being lost as dust. Low specific fuel consumption has also been claimed, about 3 %. Comparatively low metal content in slag (0.4 - 0.6 %) has been obtained in the furnace without slag cleaning. Also, the process has reportedly achieved high content of SO₂, between 25 - 65 %, in the exhaust gases allowing lower capital and operating cost during their further processing.

3.4. SUMMARY OF THE MAIN CHANNEL PROCESSES

As noted in the introduction to this chapter, the three best known channel type furnaces, namely, WORCRA, QSL and Vanyukov, although different in detail, have the same underlying principles. They are: continuous gravity driven, counter current flow of metal and slag processes with gas injection. Therefore, in principle, they may be regarded as being the same unit process in the same way as, for example, smelting-reduction is the same unit process, but of course, different in execution. For the sake of completeness, the differences of detail among the three channel type furnaces i.e. WORCRA, QSL, and Vanyukov are listed in Tables 3.3 and 3.4.
Table 3.3. Operating Parameters of WORCRA Furnace.

<table>
<thead>
<tr>
<th>Furnace</th>
<th>U-Shape Port Kembla</th>
<th>L-Shape Cockle Creek</th>
<th>Straight Shape Lulea, Sweden</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed Flowrate</td>
<td>Concentrate Cu 3 tonnes per hour</td>
<td>Iron Hot Metal 2.75 - 3.33 tph</td>
<td>Granulated Pig Iron</td>
</tr>
<tr>
<td>Product Flowrate</td>
<td>Blister Copper 0.7 tph*</td>
<td>Steel</td>
<td>Steel</td>
</tr>
<tr>
<td>Metal impurities</td>
<td>Fe, S, SiO₂, Zn, Pb, Al₂O₃, CaO 98</td>
<td>C, Si, Mn, P, S 88 - 96</td>
<td>C, Si, Mn, P, S</td>
</tr>
<tr>
<td>% Removed</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Shape</td>
<td>U-shaped Horizontal</td>
<td>L-shaped Horizontal</td>
<td>Straight Horizontal</td>
</tr>
<tr>
<td>Dimension# (m)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Smeltingbowl</td>
<td>3.05</td>
<td>15 m long</td>
<td>4.5 m long</td>
</tr>
<tr>
<td>diameter</td>
<td>8.53 x 1.83</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Converting/Refining zone</td>
<td>9.45 x 2.44</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Combustion-free Space</td>
<td>1.22 (at the riser wall in converting zone)</td>
<td>1.37 (slag-cleaning zone)</td>
<td></td>
</tr>
<tr>
<td>Bath depth</td>
<td>2.74</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Volume, m³</td>
<td>97*</td>
<td>not available</td>
<td>not available</td>
</tr>
<tr>
<td>Hearth Area, m²</td>
<td>Converting zone = 15.61*</td>
<td>not available</td>
<td>not available</td>
</tr>
<tr>
<td>(Horizontal S.A.)</td>
<td>Slag-refining zone = 23.06*</td>
<td>not available</td>
<td></td>
</tr>
<tr>
<td>Bath Volume, m³</td>
<td>21.9*</td>
<td>not available</td>
<td>not available</td>
</tr>
<tr>
<td>Lance/Tuyere</td>
<td>Lance</td>
<td>Lance</td>
<td>Lance</td>
</tr>
<tr>
<td>No. of Lance/Tuyere</td>
<td>15</td>
<td>4</td>
<td>not available</td>
</tr>
<tr>
<td>Gas Flowrate, Nm³/h</td>
<td>oxygen 2600#</td>
<td>oxygen</td>
<td>oxygen</td>
</tr>
<tr>
<td>Diameter of injector</td>
<td>85 mm</td>
<td>not available</td>
<td>not available</td>
</tr>
</tbody>
</table>

*Attained by calculation according to available data
#Converted to m from ft in the original data
Table 3.4. Operating Parameters of Vanukov and QSL Furnaces.

<table>
<thead>
<tr>
<th>Furnace</th>
<th>Vanyukov (Norilsk 1)</th>
<th>Vanyukov (Norilsk 2)</th>
<th>QSL</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed Flowrate</td>
<td>Cu concentrate 63 tph</td>
<td>Cu concentrate 195 tph</td>
<td>Pb concentrate &amp; Recycle Dust 20.75 &amp; 5.5 tph</td>
</tr>
<tr>
<td>Product Flowrate</td>
<td>Cu matte 18 tph</td>
<td>Cu matte 50 tph</td>
<td>Pb bullion 9 tph</td>
</tr>
<tr>
<td>Slag Flowrate</td>
<td>35 tph</td>
<td>80 tph</td>
<td>3.5 tph</td>
</tr>
<tr>
<td>Other Output Flowrate</td>
<td>not available</td>
<td>not available</td>
<td>Dust (recycle) 5.5 tph</td>
</tr>
<tr>
<td>Shape</td>
<td>Rectangular stationary</td>
<td>Rectangular stationary</td>
<td>Tilting cylindrical vessel</td>
</tr>
<tr>
<td>Dimension (m)</td>
<td>not available</td>
<td>not available</td>
<td>Length = 32</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>Oxidation zone diameter = 2.74</td>
</tr>
<tr>
<td></td>
<td></td>
<td></td>
<td>Reduction zone diameter = 2.25</td>
</tr>
<tr>
<td>Volume, m³</td>
<td>130</td>
<td>230</td>
<td>149</td>
</tr>
<tr>
<td>Hearth Area, m² (Horizontal S.A.)</td>
<td>20</td>
<td>36</td>
<td>not available</td>
</tr>
<tr>
<td>Bath Volume, m³</td>
<td>42</td>
<td>72</td>
<td>31</td>
</tr>
<tr>
<td>Lance/tuyere</td>
<td>Tuyere</td>
<td>Tuyere</td>
<td>Tuyere</td>
</tr>
<tr>
<td>No. of tuyeres</td>
<td>16</td>
<td>24</td>
<td>8</td>
</tr>
<tr>
<td>Gas Flowrate, Nm³/h</td>
<td>O₂ 68 %</td>
<td>O₂ 75 %</td>
<td>O₂ 96 %</td>
</tr>
<tr>
<td>Dim. of injector</td>
<td>not available</td>
<td>not available</td>
<td>not available</td>
</tr>
</tbody>
</table>
3.5. OTHER CHANNEL PROCESSES

As mentioned in Chapter 1, there are two other processes that have also started to implement the counter current pattern, Romelt and Ausmelt. The Romelt process [Weston and Thomson, 1996] adopted Vanyukov technology to produce pig iron. The project started with the pilot plant in 1985 that was closed in 1988, then restarted in 1994 with intention to market and commercialise this technology. Figure 3.5 is a schematic the diagram of the Romelt process.

Charged with coal, iron oxides and fluxes, this technology converts the iron ore or iron bearing waste materials to pig iron using non-coking coal. Oxygen is introduced from tuyeres on each side of the furnace to agitate the bath and gasify some coal, and blown from the upper side to enable post combustion. Weston and Thomson [1996] claimed that this process, using thermal coal of 16 - 20 % volatile matter and approximately 8 % ash, has been satisfactorily tested and also claimed that the product is favoured as scrap substitute as it contains no gangue, has low residuals and has a high carbon content.

Figure 3.6 shows a diagram of the Ausmelt process. The process [Cui, 1996] will use top submerged lance system to produce pig iron. A pilot plant of one tonne/day feed capacity has been run for over five years. Consideration of prototype plant of 2 tonnes per hour capacity is proceeding, with intention to establish design and operating parameters for a commercial scale plant. In commercial scale, the process has been used so far only to extract nickel and zinc from iron rich slags [Cui, 1996].
Figure 3.4. Simplified Ausmelt process flowsheet [Cui, 1996].

Figure 3.5. Simplified Romelt flowsheet [Weston and Thomson, 1996].
Yamamoto et al. (1983) adopted WORCRA concept of continuous refining to the multistage steelmaking process. In the first stage, desiliconization process with oxygen and iron ore was carried out, continued by dephosphorization and desulphurization hot metal refining with soda ash (Na$_2$CO$_3$). Countercurrent flow pattern was applied at the second stage. They claimed to have achieved around 60 to 75% of refining efficiency. The estimated dimensionless dispersion value were 0.15 and 0.5 for hot metal and slag flow respectively.
CHAPTER IV
PREVIOUS COLD MODEL STUDIES

The pilot and semi commercial plants described in Chapter 3 showed the success of continuous refining processes applying countercurrent flow of slag and metal. So far, this success is only accounted for in term of productivity, refining efficiency, and metal recovery of these processes. Each plant has a certain shape and dimension of reactor, metal and slag flowrates. It also applies different way of blowing procedure such as position and the number of injectors, and the flowrates of the injectant gas, usually oxygen.

Intrigued by this success, many studies have been done to investigate the flow characteristics in a continuous countercurrent reactor. It is interesting to recognize the behaviour of the flow in these reactors by studying how each parameter affects mass transfer and dispersion. One common technique to do this investigation is to use tracer measurement study and analyse the resultant residence time distribution. The tracer measurement can be done directly with the hot metal to obtain the quantitative data, or by making a cold model simulation, which is much more convenient and less costly, to study qualitative effects and trends.

Szekely [1967] conducted some early cold model studies to investigate longitudinal dispersion in a channel reactor using a dye tracer and colorimetric
analysis. A rectangular shape reactor was used, with dimensions of 1.2 m long by 0.1 m wide by 0.15 m high. The experiment simulated a continuous steelmaking furnace using water to represent liquid metal in single phase with bottom blowing procedure using air bubbled at the rates corresponding to the CO evolution, expected in the prototype.

Unfortunately, the information about this cold model is incomplete since other parameters, like liquid flowrates, were not reported, and other pertinent technical information was also not supplied. It was only reported that from sample analysis calculation at the outlet, inverse Peclet number of around 0.1 was attained. Szekely [1967] claimed that the result confirmed the deductions made from previous Open Hearth pilot plant trials.

A Cold model study for bottom gas injection in a single and two phase channel reactor, using thermal tracer analysis, was carried out by Agrawal [Sohn et.al., 1992]. The Major interest of this study was to approximate how the system geometry and other parameters like liquid velocity, gas flow rate, and injector separation affect the longitudinal mixing. Water and tetrachloroethylene (TCE) were used to simulate slag and metal, respectively. For single phase experiments, only one liquid flowed through the channel; for two phase experiment, both liquids flowed through the reactor in countercurrent flow.

The model reactor was rectangular in shape with basic dimensions of 2 m long by 0.2 m wide by 0.4 m high. The width could be altered by installing a temporary wall. Two other widths were chosen, namely 0.1 m and 0.128 m, while still maintaining the same dimensionless groups and geometry (represented by
width/height ratio). This alteration was meant to examine the effect of changes in scale. Two widths, 0.2 m and 0.1 m, were used in two sets of single phase experiments with water. The 0.128 m width was chosen to do one set of single phase experiment with TCE. Each set of experiments was conducted at three injector separations, three liquid flowrates, and two gas flowrates. Nitrogen gas was injected using 1 mm bore diameter bubblers to simulate bottom-blown oxygen [Sohn et.al., 1992].

The results, including typical response curves reproduced here in Figures 4.1 and 4.2, were reported in a paper by Sohn et.al. [1992]. The main results were

![Temperature Reading vs. Time](image)

**Figure 4.1.** Typical response curve results at the inlet and outlet from the single phase system of cold model experiment reported by Sohn et.al. [1992].
given in terms of the inverse Peclet number defined as $D/uW$, with $W$ being the width of the channel. For the dimension of their system given above, their $D/uW$ values are ten times those of $D/uL$, the common definition of inverse Peclet number. Three other dimensionless groups that Sohn et.al. [1992] considered important were dimensionless gas flow rate, Froude number, and Reynolds number.

Figure 4.2. Typical response curve results at the TCE and water outlet from the countercurrent system of cold model experiment reported by Sohn et.al. [1992].
According to equation (2.10), the term $D/u_w$, defined by Sohn et.al. (1992) as the inverse Peclet number, is actually intensity of dispersion, which, as noted earlier, is correlated with Schmidt and Reynolds number (Figures 2.9 and 2.10). As regards the $D/u_L$ (and $D/ud$) term in equation (2.10), Levenspiel [1993] states: “This is a new and different type of dimensionless group introduced by workers in chemical reaction engineering. Unfortunately somebody started calling the reciprocal of this group the Peclet number. This is wrong, it is neither the Peclet number nor its mass transfer analog, which is widely called the Bodenstein number in Europe. The difference rests in the use of $D$ in place of $\bar{D}$, hence these groups have completely different meaning.”

One result presented by Sohn et.al. [1992] is the plot of $D/u_W$ versus Reynolds number, reproduced here in Figure 4.3 and showing the effect of two bubbling rates. Comparing Figure 4.3 with Figure 2.10, it is clear that $D/u_W$ of Sohn et.al. [1992] follows the turbulent part of Figure 2.10 well, including the absolute values, the difference between which being accounted for by the different characteristic dimension employed ($W$ vs $d$).

The fact that there is no initial increase in $D/u_W$, even though the Reynolds numbers are low, and are in the laminar region in Figure 2.10, could be due to:
1) the documented laminar/turbulent transition in channel flow occurring at a different Reynolds numbers from that in pipe flow [Murphy, 1950], and
2) an effect of a bubbling gas on the flow regime.

The latter is supported by noting that quadrupling the bubbling gas flow rate in Figure 4.3 only shifts the curve slightly up and not move it proportionally to the gas flow rate increase.
One significant conclusion of Sohn et.al. [1992] is that the results of one system can be predicted using the results from other system with the same geometry and the same dimensionless groups value. This finding is essential for designing systems with entirely different parameters such as scales, kind of liquids, or liquid and gas flowrate.

The above conclusion was obtained by conducting two sets of single phase experiments using two liquids with different kinematic viscosities (water and TCE)
but complete dynamic similarity. The results of these experiments are represented by Figure 4.4 in which the \( \frac{D}{uW} \) values are plotted vs dimensionless injector spacing (\( S/W \), where \( S \) is the distance between two injectors) for both systems with matching parameters.

![Figure 4.4. Comparison of \( \frac{D}{uW} \) between dynamically similar systems [Sohn et.al., 1992].](image)

There are two pairs of corresponding flowrates from complete dynamic similarity systems plotted in Figure 4.4. The water system was 0.2 m wide and 0.2 m high with \( 0.000075 \) \( m^3/s \) gas flowrate; the TCE system was 0.128 m wide and 0.128 m high with \( 0.000025 \) \( m^3/s \) gas flowrate. The first pair of liquid flowrates that shows dynamic similarity is \( 0.0035 \) m/s water flowrate and \( 0.0028 \) m/s TCE flowrate.
Water flowrate of 0.0142 m/s corresponding to TCE flowrate of 0.0113 m/s creates the second pair. Figure 4.4 clearly shows that the D/uW values of each corresponding pair are approximately equivalent [Sohn et.al., 1992].

A similar experiment comparing two single phase systems using water with different parameters has been formerly carried out by Agrawal (Sohn et.al., 1992]. Two water systems with different scale and gas flowrates were tested. The first system was 0.1 m wide and 0.1 m high with 0.0000033 m$^3$/s gas flowrate; the second system was 0.2 m wide by 0.2 m high with 0.000019 m$^3$/s. The results are presented in Figure 4.5.

For these results, the condition of complete dynamic similarity could not be obtained between the two systems with the same kinematic viscosity, at least one dimensionless group had different value. Figure 4.5 shows that the water flowrate of 0.0025 m/s in the 0.1 m-wide system was dynamically similar with water flowrate of 0.0035 m/s in the 0.2 m-wide system in terms of having the same Froude number and dimensionless gas flowrate, but the D/uW values are considerably different.
Reynolds numbers of the 0.2 m-wide system are 2.8 times the 0.1 m-wide system for a certain corresponding water flowrates. This difference of Reynolds numbers is significant in lower liquid flowrate cases for causing the difference of D/uW in the two systems which has limited dynamic similarity. The system with higher Reynolds numbers will have lower D/uW values. At higher liquid flowrate, i.e. in the case of 0.01 m/s and 0.0142 m/s water flowrate pair, the influence of the Reynolds number in decreasing the D/uW value becomes less [Sohn et.al., 1992].

It is also important to observe that the curves in Figures 4.4 and 4.5 demonstrate similar trend lines which identify injector separation effect. These trend lines
suggest that longitudinal mixing can be minimised by controlling the injection spacing. In the foregoing work, the value of $D/uW$ was maximum if the injectors were apart at the distance twice the width for most conditions. This characteristic of injector spacing was also recognized by Sahai [1979].

With regard to the effect of the gas flow rate and liquid velocity on $D/uW$, Sohn et.al. [1992] chose to express the effect on $D$ alone, namely as $D \propto Q^a u^b$. Again, these authors refer to $D$ as "the eddy diffusivity", which according to Levenspiel [1993], as noted earlier, is wrong. The effect of $Q$ and $u$ on $D$ of Sohn et.al. [1992] was found not to be too precise and to also vary from one set of experiments to the other. Thus, Sohn et.al. [1992] state that "a regression analysis of the data showed that the eddy diffusivity in lower phase was proportional to $Q^{0.11}$ and it was also proportional to $u^{0.09}$.

On the other hand, for the single phase their reported regression analysis showed $D \propto Q^{0.25} u^{0.45}$, in the upper phase of the two phase system as $D \propto Q^{0.50} u^{0.48}$, and in the case of the half width experiments $D \propto Q^{0.1} u^{0.23}$. Sohn et.al. [1992] have attempted to explain these differences, and others, between their results and those of Engh et.al. [1971] and Jenkins et.al. [1971] obtained on a WORCRA reactor as being due to a "very different flow behaviour".

In their paper, Sohn et.al. [1992] developed a flow model of the channel reactor by combining one ideal mixed regions and two ideal plug flow recycle regions
from Figure 2.11, noting that the experimental and predicted responses were reasonably comparable. In this regard it may be of interest to remark that a plug flow model incorporating two ideal mixed flow regions in parallel seems also to agree with their response curve. This can be demonstrated by reploting the TCE curve in Figure 4.2 on semi-logarithmic coordinates as shown in Figure 4.6, and comparing the result with that of the flow model given by Levenspiel [1993] in Figure 4.7.

![Figure 4.6. Semi-logarithmic plot of the TCE outlet curve of Figure 4.2.](image-url)
What the foregoing results suggest is that, perhaps, the response curves obtained in this kind of work are imprecise and therefore they can fit any reasonable flow model. This is, of course, not to say that flow modelling is of no value. On the contrary, what is implied by the above is that more work should be done to perfect the system and narrow the flow model range. Brooks et.al. [1996] suggested that significant study is required to resolve many issues of channel design, i.e. the relationship between mass transfer and dispersion, geometry and dispersion, and optimum depths of slag and metal phases.
CHAPTER V
EXPERIMENTAL

5.1. EXPERIMENTAL DESIGN

A basic cold model was developed to simulate industrial countercurrent refining reactors at room temperature with top lance gas injection. The model was designed to allow control of liquid and gas injection flowrates, as well as manipulation of lance spacing and height. The model was purposely based on the bottom injection model of Sohn et.al [1992] to allow direct comparison of results.

5.1.1. MATERIALS

Two fluids were used in these experiments to model the molten metal and slag at room temperature. Ideally, these fluids would have similar flow characteristics as the real molten metal and slag. The fluids needed to be immiscible to each other and have a density ratio, ideally above 2.0, to simulate molten metal and slag. They should also have a relatively high boiling point, because a thermal tracer technique was applied.

The selected liquids were tetrachloroethylene (TCE) and water. These acted as molten metal and slag, respectively. Water was first considered as one of
the fluids because it is easy to obtain, low cost, non flammable and non toxic. Tetrachloroethylene was chosen because it is considered as the most compatible liquid with the water, according to the criteria set above.

This pair of liquids (TCE and water) is not an ideal representative of molten metal and slag in a cold model, especially its density difference that is not as high as the real molten liquid (density ratio of TCE and water is only 1.62). This is chosen partly based on its availability, flammability and toxicity reasons. Sohn et.al [1992] and Nelson et.al [1995] chose the same liquids for their cold model studies with bottom gas injection. Some of these properties are listed in Table 5.1.

Table 5.1. Physical Properties of Tetrachloroethylene and Water (Nelson et.al., 1995).

<table>
<thead>
<tr>
<th>Property</th>
<th>Water</th>
<th>Tetrachloroethylene</th>
</tr>
</thead>
<tbody>
<tr>
<td>Density (kg/m³) 298 K</td>
<td>997</td>
<td>1615</td>
</tr>
<tr>
<td>Dynamic Viscosity (kg/ms) 298 K</td>
<td>8.9 × 10⁻⁴</td>
<td>8.4 × 10⁻⁴</td>
</tr>
<tr>
<td>Kinematic Viscosity (m²/s) 298 K</td>
<td>8.9 × 10⁻⁷</td>
<td>5.2 × 10⁻⁷</td>
</tr>
<tr>
<td>Melting Point (K)</td>
<td>273.2</td>
<td>250.5</td>
</tr>
<tr>
<td>Boiling Point (K)</td>
<td>373.2</td>
<td>394.3</td>
</tr>
<tr>
<td>Thermal Conductivity (W/mK)</td>
<td>0.606</td>
<td>0.127</td>
</tr>
<tr>
<td>Interfacial Tension (N/m) 293 K</td>
<td></td>
<td>0.04748</td>
</tr>
</tbody>
</table>
5.1.2. APPARATUS

5.1.2.1. Channel Reactor

The reactor was made of perspex with dimensions, corresponding to those used by Sohn et.al. [1992], of 0.2 m wide by 2 m long by 0.2 m high (Figure 5.1). Perspex was chosen due to its compatibility to the liquids. Perspex was also used by Sohn and co-workers [Sohn et.al, 1992; Iyer and Sohn, 1994], so the resistance of perspex especially to tetrachloroethylene has been proven in previous work.

Figure 5.1. Schematic diagram of experimental channel/reactor.
Baffles were installed at both ends to allow counter current flow to occur and to control the fluid heights. Baffle A (12 cm high), called the TCE height regulator baffle, kept the height of TCE inside the tank at a certain level, which was 10.5 cm. Baffle B, called the water dam baffle, was placed to function as a dam for water so it flowed in an opposite direction to TCE. The height of water, at 2 cm high from the interface, for countercurrent experiments, was determined by Baffle C (12.5 cm high), the height regulator baffle for water. In single phase experiments, the water height was maintained by Baffle A, at 12 cm.

5.1.2.2. TRACER DEVICES

Following Sohn et.al [1992], heat was chosen as the tracer material, provided by an electric hand held heater, for obtaining the residence time distribution curve. Thermistors, which are temperature variant resistors, were used to measure temperature changes with time by detecting the changes of voltage across the thermistor. Figure 5.2 shows the electrical circuit for each thermistor.

Stainless steel tubes were used to cover the connecting wires between the thermistor and an IBM laptop computer AT-386 which acted as data logging device. The data were recorded and saved on the disk in the form of comma delimited spreadsheet file. This allowed to perform calculations using Excel software later.
Figure 5.2. Diagram of electrical circuit for each thermistor.

The thermistors, made in the laboratory, were very sensitive and capable of detecting approximately 0.06 degree change of temperature on one reading point. Figure 5.3 shows calibration curve between the thermistor reading and the actual change of temperature, in degrees Celcius.

The calibration was done by measuring temperature changes using a thermistor and a thermometer in parallel. A thermal pulse was introduced to a bucket of water, containing both the thermistor and the thermometer placed in the same spot. The thermistor recorded the peak value and the thermometer read the real temperature in degrees Celcius, before the pulse was introduced as $T_0$ and the highest temperature reached as $T_1$. This procedure was repeated several times, varying the heating time. The recorded
thermometer read the real temperature in degrees Celsius, before the pulse was introduced as $T_0$ and the highest temperature reached as $T_1$. This procedure was repeated several times, varying the heating time. The recorded peak value from the thermistor reading was plotted against the temperature difference $(T_1 - T_0)$ to obtain the calibration curve.

![Calibration curve of thermistor reading](image)

Figure 5.3. Calibration curve of thermistor reading.

5.1.2.3. GAS BLOWING DEVICES

The experiments in the present study used three gas injectors, 1 mm bore-diameter, to simulate the condition of oxygen injection in the real refining process while enhancing the mass transfer effect at the interphase boundary. Each injector was connected to a flowmeter that could be varied from 0.00001 m$^3$/s to 0.000083 m$^3$/s of gas flowrate. A gas bottle containing industrial nitrogen was used to provide the gas needed. The injection gas was
introduced from the top of the reactor and the position of the injectors could be adjusted horizontally and vertically.

5.1.2.4. Ancillaries

Flow of the water was regulated by a rotameter with a maximum flowrate of 0.000091 m$^3$/s. Water leaving the reactor went directly to a drain. The TCE required a storage tank, because the TCE outflow could not be just drained out for economic and safety reasons. The TCE was collected in a storage tank and circulated back into the reactor. The TCE was pumped out, using a variable drive magnetic pump, from the storage tank to the reactor, via a rotameter with maximum capacity of 0.000093 m$^3$/s.

Other supporting equipment were some pipes, fittings, and flexible hose to make connections and to set up the equipment. Standard laboratory retort stands and steel plates were also used to hold the thermistors and injectors in place and to enable the adjustments to be made as required.

5.1.3. EQUIPMENT SET-UP

Generally, the equipment used was set up as shown in Figure 5.4. The reactor was put on a supporting stand to allow the installation of TCE and water outlets at the bottom of the reactor. The TCE recirculating system was formed by fitting the storage tank, pump, and the TCE flowmeter together, then, using
the flexible hose, connecting the flowmeter to the TCE inlet and connecting the TCE outlet to the storage tank. The water circulation was installed by connecting the tap, water flowmeter, and water inlet, then connecting the water outlet to the drain.

Figure 5.4. Equipment setup for the countercurrent experiment.
The above configuration was applied for the countercurrent experiment. For single phase experiments, this configuration was simplified. The connection starting from the water supply to the water drain was not used. Water, as the fluid used in the single phase experiment, was placed in the recirculation tanks then pumped to the reactor and so forth, following the TCE flow path in countercurrent experiment.

The gas blowing system was formed by distributing the gas from the gas bottle to the three gas flowmeters using a manifold and connecting the flowmeters to the injectors. The injectors were positioned at the top of the reactor and to achieve top blowing condition, they were separated 0.3 m from each other. The depth of injectors would be adjusted during the experiments by retort stands.

Thermistors were placed near the tracer inlet to measure the input curve of the tracer, and also near the water dam baffle to measure the output curve. The thermistors were held by a steel plate to allow immersion depth adjustment. Figure 5.5 illustrates this injectors and tracer measurement arrangements.

For countercurrent experiments, the tracer was introduced by quickly immersing the hand held heater (Fig. 5.5a). In single phase experiments, the tracer was introduced by pouring boiling water into the inlet (Fig. 5.5b). In both cases, the heater was placed in a bucket filled with water. This allowed the heater to be heated outside the reactor before immersion in the countercurrent case, or to boil the water before being poured in the single phase case.
Figure 5.5. Schematic diagram of bubblers and thermistors configuration;
(a) two phase system,
(b) single phase system.
5.1.4. PROCEDURE

The main point of these experiments was to examine the effect of the top blowing, on the vessel dispersion value. Conditions that were assumed to have influence on this value were the depth of immersion of the gas injector, and the gas flowrate. Therefore, in one series of experimental runs, one condition was set up as a constant while the other was varied. As an addition, the influence of Reynolds number was studied by varying the fluid flowrate in each run of a series.

The first experimental stage was carried out at a gas flowrate of 1500 cc/minute per injector to study the effect of different immersion depths on the vessel dispersion number at various Reynolds numbers. Three variations of immersion of the injector were chosen, namely at the interface, 2 cm below the interface, and 4 cm below the interface, for countercurrent experiments. For single phase experiments, using water only, the injectors were positioned at the surface, 2 cm below the surface, and 4 cm below the surface.

The next stage was done by keeping the injectors steady at 2 cm below the interface to investigate the variation of gas flowrate at 800 cc/minute, 1500 cc/minute, 2000 cc/minute, and 4000 cc/minute. These variations of gas flowrates were chosen to provide a comparison of a wide range of gas flowrates and to study how these affect the flow dispersion.
The experiments were divided into three sets. Each set applied the conditions and procedure outlined above. The first set was the single phase experiment. In this experiment, water was used as the only liquid flow through the reactor, in the TCE direction (refer to Fig. 5.4). The water velocity was varied from 0.1 cm/s to 0.52 cm/s, equivalent to Reynolds numbers of 234 – 1170.

Two other sets of experiments were the countercurrent experiments. Here, the TCE velocities were varied between 0.09 cm/s to 0.44 cm/s, associated with Reynolds number of 341 - 1703. Each set of experiment had the same TCE flow rate variations, with a certain constant water flow rate. The water flow rate was kept constant for each set, 0.000018 m³/s (slow slag moving experiment) and 0.000037 m³/s (fast slag moving experiment), corresponding to Reynolds number of 1011 and 2079, respectively. Table 5.2 gives the parameters applied for each experimental series.
Table 5.2. Experimental series parameters.

<table>
<thead>
<tr>
<th>Seri Code</th>
<th>type of experiment</th>
<th>gas flow rate per injector (l/minute)</th>
<th>injector depth (cm)</th>
<th>Flow rates (cm/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Series 1</td>
<td>one phase</td>
<td>1.5</td>
<td>0</td>
<td>0.52 0.47 0.42 0.37 0.31 0.26 0.21 0.16 0.10</td>
</tr>
<tr>
<td>Series 2</td>
<td>one phase</td>
<td>1.5</td>
<td>2</td>
<td>0.52 0.47 0.42 0.37 0.31 0.26 0.21 0.16 0.10</td>
</tr>
<tr>
<td>Series 3</td>
<td>one phase</td>
<td>1.5</td>
<td>4</td>
<td>0.52 0.47 0.42 0.37 0.31 0.26 0.21 0.16 0.10</td>
</tr>
<tr>
<td>Series 4</td>
<td>one phase</td>
<td>0.8</td>
<td>2</td>
<td>0.52 0.47 0.42 0.37 0.31 0.26 0.21 0.16 0.10</td>
</tr>
<tr>
<td>Series 5</td>
<td>one phase</td>
<td>2.0</td>
<td>2</td>
<td>0.52 0.47 0.42 0.37 0.31 0.26 0.21 0.16 0.10</td>
</tr>
<tr>
<td>Series 6</td>
<td>one phase</td>
<td>4.0</td>
<td>2</td>
<td>0.52 0.47 0.42 0.37 0.31 0.26 0.21 0.16 0.10</td>
</tr>
<tr>
<td>Series 7</td>
<td>slow slag moving</td>
<td>1.5</td>
<td>0</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 8</td>
<td>slow slag moving</td>
<td>1.5</td>
<td>2</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 9</td>
<td>slow slag moving</td>
<td>1.5</td>
<td>4</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 10</td>
<td>slow slag moving</td>
<td>0.8</td>
<td>2</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 11</td>
<td>slow slag moving</td>
<td>2.0</td>
<td>2</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 12</td>
<td>slow slag moving</td>
<td>4.0</td>
<td>2</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 13</td>
<td>fast slag moving</td>
<td>1.5</td>
<td>0</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 14</td>
<td>fast slag moving</td>
<td>1.5</td>
<td>2</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 15</td>
<td>fast slag moving</td>
<td>1.5</td>
<td>4</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 16</td>
<td>fast slag moving</td>
<td>0.8</td>
<td>2</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 17</td>
<td>fast slag moving</td>
<td>2.0</td>
<td>2</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
<tr>
<td>Series 18</td>
<td>fast slag moving</td>
<td>4.0</td>
<td>2</td>
<td>0.41 0.37 0.33 0.28 0.24 0.20 0.16 0.12 0.08</td>
</tr>
</tbody>
</table>
5.2. RESULTS

The first six experimental series were carried out in single phase, followed by six series of 0.023 m/s constant slag flowrate experiments, then six series of fast moving slag, successively. In this section, some raw data will be reported, continued with some calculated results. The data will be shown in graphical forms, as a result of temperature measurements that were recorded by thermistors at the inlet and outlet. The calculated results will be presented either in graphs or in tables.

In those tables, the bubbling condition is represented by “x/y” notation, means that the bubblers were immersed at “x” cm deep with bubbling flowrate of “y”. Therefore, 0/1500 is interpreted as 1500 cc/minute gas flowrate positioned at the surface (in single phase case) or at the interface (in countercurrent case), and 2/800 is 800 cc/minute gas flowrate positioned at 2 cm below the surface or the interface.

5.2.1 SINGLE PHASE EXPERIMENTS

Figure 5.6 shows a typical tracer response curve at the inlet and outlet that well represents the fifty four single phase experimental runs. The inlet curve is characterised by a distinct peak, formed almost instantly after the tracer was injected, followed by a long tail while approaching the base line condition. The
Figure 5.6. Inlet and outlet tracer response curve for single phase experiments. Water flow rate = 0.52 cm/s, gas flow rate = 1500 cc/min., injector depth = 2 cm below the surface.

time that was needed to reach base line or initial condition increased as the flow rate decreased. This can be seen in Figure 5.7.

The outlet curve is characterised by a bell-shaped curve with lower peak value than the inlet response curve. The width of the bell part of the curve varied, with slower liquid flowrate giving a wider bell-shape. Figure 5.6 shows the spread of the outlet response curve for an experimental series with gas flow rate of 1.5 l/minute and injector depth of 2 cm. It should be noted that the response curves do not appear smooth because they were printed on a coarse setting. The actual sampling rate was, of course, very much higher.
Figure 5.7. Changes of shapes of tracer response curve with the decreasing of water flowrate. Gas flowrate = 1500 cc/min., injector depth = 2 cm below the surface.
Tables 5.3 and 5.4 show the calculation results of mean time and $D/uL$, respectively, for different bubbling condition. Experimental mean times were calculated by equation (2.12), and the theoretical mean times, calculated by equation (2.9), are also shown in this table. Table 5.4 shows the results of $D/uL$ calculation, with first $\Delta \sigma^2$ obtained by equation (2.12) and (2.16), and then $D/uL$ calculated from equation (2.15). The results of the $D/uL$ calculation in Table 5.4 are also plotted against their corresponding Reynolds number in Figures 5.8 to 5.13 for each bubbling condition.

It was realized that wider range of Reynolds number than 1200 for single phase experiment and 1700 for countercurrent was needed for better result and the present work would have done that if possible. But, the maximum value of Reynolds number for both cases was limited due to the limitation of the equipment; in this case, the water flowmeter and TCE flowmeter.

Table 5.3. Experimental mean times results for single phase experiments.

<table>
<thead>
<tr>
<th>Flowrate (cm/s)</th>
<th>$t_{mean}$ (minutes)</th>
<th>Calculation</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0/ 1500</td>
<td>2/ 1500</td>
</tr>
<tr>
<td>0.10</td>
<td>17.6</td>
<td>16.8</td>
</tr>
<tr>
<td>0.16</td>
<td>9.8</td>
<td>11.5</td>
</tr>
<tr>
<td>0.21</td>
<td>7.5</td>
<td>8.2</td>
</tr>
<tr>
<td>0.26</td>
<td>5.4</td>
<td>5.8</td>
</tr>
<tr>
<td>0.31</td>
<td>5.7</td>
<td>5</td>
</tr>
<tr>
<td>0.37</td>
<td>4.1</td>
<td>4.1</td>
</tr>
<tr>
<td>0.42</td>
<td>3.4</td>
<td>3.6</td>
</tr>
<tr>
<td>0.47</td>
<td>2.8</td>
<td>3.2</td>
</tr>
<tr>
<td>0.52</td>
<td>2.1</td>
<td>2.1</td>
</tr>
</tbody>
</table>

81
Table 5.4. D/\mu L results for single phase experiments.

<table>
<thead>
<tr>
<th>Reynolds Number</th>
<th>0/1500</th>
<th>2/1500</th>
<th>4/1500</th>
<th>2/800</th>
<th>2/2000</th>
<th>2/4000</th>
</tr>
</thead>
<tbody>
<tr>
<td>234</td>
<td>0.04</td>
<td>0.05</td>
<td>0.06</td>
<td>0.04</td>
<td>0.05</td>
<td>0.07</td>
</tr>
<tr>
<td>351</td>
<td>0.06</td>
<td>0.06</td>
<td>0.07</td>
<td>0.05</td>
<td>0.06</td>
<td>0.07</td>
</tr>
<tr>
<td>468</td>
<td>0.04</td>
<td>0.05</td>
<td>0.08</td>
<td>0.04</td>
<td>0.07</td>
<td>0.06</td>
</tr>
<tr>
<td>585</td>
<td>0.06</td>
<td>0.06</td>
<td>0.08</td>
<td>0.06</td>
<td>0.06</td>
<td>0.08</td>
</tr>
<tr>
<td>702</td>
<td>0.04</td>
<td>0.04</td>
<td>0.05</td>
<td>0.04</td>
<td>0.04</td>
<td>0.05</td>
</tr>
<tr>
<td>819</td>
<td>0.04</td>
<td>0.04</td>
<td>0.05</td>
<td>0.03</td>
<td>0.04</td>
<td>0.05</td>
</tr>
<tr>
<td>936</td>
<td>0.09</td>
<td>0.06</td>
<td>0.07</td>
<td>0.07</td>
<td>0.05</td>
<td>0.07</td>
</tr>
<tr>
<td>1053</td>
<td>0.07</td>
<td>0.07</td>
<td>0.06</td>
<td>0.07</td>
<td>0.06</td>
<td>0.06</td>
</tr>
<tr>
<td>1170</td>
<td>0.07</td>
<td>0.03</td>
<td>0.05</td>
<td>0.04</td>
<td>0.04</td>
<td>0.05</td>
</tr>
</tbody>
</table>

Figure 5.8. Plot of D/\mu L against Reynolds number for single phase experiments with bubbling condition:
1500 cc/minute flowrate, positioned at the surface.
Figure 5.9. Plot of $D/uL$ against Reynolds number for single phase experiments with bubbling condition: 1500 cc/minute flowrate, positioned 2 cm below the surface.

Figure 5.10. Plot of $D/uL$ against Reynolds number for single phase experiments with bubbling condition: 1500 cc/minute flowrate, positioned 4 cm below the surface.
Figure 5.11. Plot of $D/uL$ against Reynolds number for single phase experiments with bubbling condition: 800 cc/minute flowrate, positioned 2 cm below the surface.

Figure 5.12. Plot of $D/uL$ against Reynolds number for single phase experiments with bubbling condition: 2000 cc/minute flowrate, positioned 2 cm below the surface.
Figure 5.13. Plot of $D/uL$ against Reynolds number for single phase experiments with bubbling condition: 4000 cc/minute flowrate, positioned 2 cm below the surface.

5.2.2. COUNTER CURRENT EXPERIMENTS

Figure 5.14 shows a typical tracer response curve at the inlet and outlet that represents thirty six counter current experimental runs with slow slag moving (Figure 5.14a) and twenty seven runs with fast slag moving (Figure 5.14b). As in single phase, the inlet curve is characterised by an instant peak after the tracer was introduced, followed by a long tail while approaching the base line condition. The difference is, in the counter current experiments, the peak is not as sharp as in the single phase experiments, and becomes more disperse at lower TCE flowrates, in that the width of the peak becomes wider, and the long tail is more pronounced.
Figure 5.14. Inlet and outlet tracer response curve for counter current experiments. TCE flow rate = 0.44 cm/s, gas flow rate = 1500 cc/min., injector depth = 2 cm below the surface.

a) Water flow rate = 0.45 cm/s.
b) Water flow rate = 0.9 cm/s.

The outlet curve is characterised by a truncated bell-shaped curve with long tail, which is different from the single phase experiments. The peak of the curve is much lower than that of the inlet curve. The width of the bell part of the curve varied, with slower liquid flowrate giving a wider bell-shape, as well as longer tail.
The experimental mean times for countercurrent experiments, calculated by equation (2.12), are shown in Tables 5.5 and 5.6 for slow slag moving and for fast slag moving, respectively. Theoretical mean times, as calculated by equation (2.9), are also given in each table.

Table 5.5. Experimental mean times results for countercurrent experiments with slow slag moving.

<table>
<thead>
<tr>
<th>Flowrate (cm/s)</th>
<th>t mean (minutes)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0/1500</td>
</tr>
<tr>
<td>0.09</td>
<td>2.7</td>
</tr>
<tr>
<td>0.13</td>
<td>1.9</td>
</tr>
<tr>
<td>0.18</td>
<td>2.6</td>
</tr>
<tr>
<td>0.22</td>
<td>2.4</td>
</tr>
<tr>
<td>0.27</td>
<td>2.5</td>
</tr>
<tr>
<td>0.31</td>
<td>2.7</td>
</tr>
<tr>
<td>0.36</td>
<td>2.6</td>
</tr>
<tr>
<td>0.40</td>
<td>2.3</td>
</tr>
<tr>
<td>0.44</td>
<td>2.5</td>
</tr>
</tbody>
</table>
Table 5.6. Experimental mean times results for countercurrent experiments with fast slag moving.

<table>
<thead>
<tr>
<th>Flowrate (cm/ s)</th>
<th>t mean (minutes)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0/ 1500</td>
</tr>
<tr>
<td>0.09</td>
<td>1.667</td>
</tr>
<tr>
<td>0.13</td>
<td>1.167</td>
</tr>
<tr>
<td>0.18</td>
<td>1.256</td>
</tr>
<tr>
<td>0.22</td>
<td>1</td>
</tr>
<tr>
<td>0.27</td>
<td>1.042</td>
</tr>
<tr>
<td>0.31</td>
<td>1.292</td>
</tr>
<tr>
<td>0.36</td>
<td>1.423</td>
</tr>
<tr>
<td>0.40</td>
<td>1.333</td>
</tr>
<tr>
<td>0.44</td>
<td>1.625</td>
</tr>
</tbody>
</table>

Table 5.7 shows the results of D/uL calculation for countercurrent experiment with slow slag moving, first applying equations (2.12) and (2.16) and then equation (2.15). The D/uL results for countercurrent experiment with fast slag moving are shown in Table 5.8. These D/uL values are also presented in graphs, Figures 5.15 to 5.21, for different bubbling and slag flowrate condition.
Table 5.7. D/ul results for countercurrent experiments with slow slag moving.

<table>
<thead>
<tr>
<th>Reynolds Number</th>
<th>D/ul 0/1500</th>
<th>D/ul 2/1500</th>
<th>D/ul 2/800</th>
<th>D/ul 2/2000</th>
</tr>
</thead>
<tbody>
<tr>
<td>342</td>
<td>0.03</td>
<td>0.04</td>
<td>0.02</td>
<td>0.04</td>
</tr>
<tr>
<td>513</td>
<td>0.04</td>
<td>0.04</td>
<td>0.05</td>
<td>0.05</td>
</tr>
<tr>
<td>684</td>
<td>0.05</td>
<td>0.06</td>
<td>0.06</td>
<td>0.07</td>
</tr>
<tr>
<td>855</td>
<td>0.06</td>
<td>0.08</td>
<td>0.08</td>
<td>0.09</td>
</tr>
<tr>
<td>1026</td>
<td>0.08</td>
<td>0.09</td>
<td>0.1</td>
<td>0.07</td>
</tr>
<tr>
<td>1197</td>
<td>0.07</td>
<td>0.07</td>
<td>0.1</td>
<td>0.09</td>
</tr>
<tr>
<td>1368</td>
<td>0.08</td>
<td>0.08</td>
<td>0.07</td>
<td>0.08</td>
</tr>
<tr>
<td>1538</td>
<td>0.04</td>
<td>0.04</td>
<td>0.07</td>
<td>0.06</td>
</tr>
<tr>
<td>1709</td>
<td>0.06</td>
<td>0.06</td>
<td>0.07</td>
<td>0.08</td>
</tr>
</tbody>
</table>

Table 5.8. D/ul results for countercurrent experiments with fast slag moving.

<table>
<thead>
<tr>
<th>Reynolds Number</th>
<th>D/ul 0/1500</th>
<th>D/ul 2/1500</th>
<th>D/ul 2/800</th>
</tr>
</thead>
<tbody>
<tr>
<td>1709</td>
<td>0.06</td>
<td>0.02</td>
<td>0.07</td>
</tr>
<tr>
<td>1538</td>
<td>0.26</td>
<td>0.08</td>
<td>0.08</td>
</tr>
<tr>
<td>1368</td>
<td>0.14</td>
<td>0.11</td>
<td>0.18</td>
</tr>
<tr>
<td>1197</td>
<td>0.13</td>
<td>0.04</td>
<td>0.06</td>
</tr>
<tr>
<td>1026</td>
<td>0.18</td>
<td>0.02</td>
<td>0.11</td>
</tr>
<tr>
<td>855</td>
<td>0.15</td>
<td>0.06</td>
<td>0.1</td>
</tr>
<tr>
<td>684</td>
<td>0.09</td>
<td>0.06</td>
<td>0.06</td>
</tr>
<tr>
<td>513</td>
<td>0.13</td>
<td>0.15</td>
<td>0.16</td>
</tr>
<tr>
<td>342</td>
<td>0.11</td>
<td>0.07</td>
<td>0.07</td>
</tr>
</tbody>
</table>
Figure 5.15. Plot of $D/uL$ against Reynolds number for slow slag moving counter current experiments with bubbling condition: 1500 cc/minute flowrate, positioned at the interface.

Figure 5.16. Plot of $D/uL$ against Reynolds number for slow slag moving counter current experiments with bubbling condition: 1500 cc/minute flowrate, positioned 2 cm below the interface.
Figure 5.17. Plot of $D/u_L$ against Reynolds number for slow slag moving counter current experiments with bubbling condition: 800 cc/minute flowrate, positioned 2 cm below the interface.

Figure 5.18. Plot of $D/u_L$ against Reynolds number for slow slag moving counter current experiments with bubbling condition: 2000 cc/minute flowrate, positioned 2 cm below the interface.
Figure 5.19. Plot of $D/u_L$ against Reynolds number for fast slag moving counter current experiments with bubbling condition: 1500 cc/minute flowrate, positioned at the interface.

Figure 5.20. Plot of $D/u_L$ against Reynolds number for fast slag moving counter current experiments with bubbling condition: 1500 cc/minute flowrate, positioned 2 cm below the interface.
Figure 5.21. Plot of $D/uL$ against Reynolds number for fast slag moving counter current experiments with bubbling condition: 800 cc/minute flowrate, positioned 2 cm below the interface.
CHAPTER VI
DISCUSSION

6.1. REPRODUCIBILITY OF RESULTS

The result of replicated runs, for both single phase and two phase experiments, are given in Tables 6.1 and 6.2, respectively.

Table 6.1. Reproducibility of results of 9 replicated runs for the single phase experiments.

<table>
<thead>
<tr>
<th>Run</th>
<th>t (minutes)</th>
<th>D/μL</th>
<th>square of deviation t</th>
<th>D/μL</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>3.8</td>
<td>0.17</td>
<td>0.3403</td>
<td>0.0021</td>
</tr>
<tr>
<td>2</td>
<td>3.4</td>
<td>0.13</td>
<td>0.0336</td>
<td>0.0000</td>
</tr>
<tr>
<td>3</td>
<td>3.05</td>
<td>0.13</td>
<td>0.0278</td>
<td>0.0000</td>
</tr>
<tr>
<td>4</td>
<td>2.86</td>
<td>0.14</td>
<td>0.1272</td>
<td>0.0002</td>
</tr>
<tr>
<td>5</td>
<td>2.68</td>
<td>0.14</td>
<td>0.2880</td>
<td>0.0002</td>
</tr>
<tr>
<td>6</td>
<td>3.5</td>
<td>0.11</td>
<td>0.0803</td>
<td>0.0002</td>
</tr>
<tr>
<td>7</td>
<td>3.64</td>
<td>0.1</td>
<td>0.1792</td>
<td>0.0006</td>
</tr>
<tr>
<td>8</td>
<td>2.37</td>
<td>0.1</td>
<td>0.7168</td>
<td>0.0006</td>
</tr>
<tr>
<td>9</td>
<td>3.65</td>
<td>0.1</td>
<td>0.1878</td>
<td>0.0006</td>
</tr>
<tr>
<td>total</td>
<td>28.9500</td>
<td>1.1200</td>
<td>1.9810</td>
<td>0.0046</td>
</tr>
<tr>
<td>mean</td>
<td>3.2167</td>
<td>0.1244</td>
<td>0.2201</td>
<td>0.0005</td>
</tr>
</tbody>
</table>

standard deviation =

0.4692  0.0227
From Table 6.1, it is seen that \( \bar{t} \), the mean of nine runs, is 3.22 minutes with the standard deviation of 0.47 minutes and for \( D/uL \), the mean value is 0.124 with the standard deviation of 0.023. These results are considered to be well within the experimental error and their reproducibility can be regarded as sufficient for comparative purposes considered.

Table 6.2. Reproducibility of results of 7 replicated runs for the two phase experiments.

<table>
<thead>
<tr>
<th>Run</th>
<th>( t ) (minutes)</th>
<th>( D/uL )</th>
<th>square of deviation</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>3.55</td>
<td>0.11</td>
<td>0.142237</td>
</tr>
<tr>
<td>2</td>
<td>3.4</td>
<td>0.12</td>
<td>0.051594</td>
</tr>
<tr>
<td>3</td>
<td>2.78</td>
<td>0.13</td>
<td>0.154337</td>
</tr>
<tr>
<td>4</td>
<td>3.12</td>
<td>0.12</td>
<td>0.002794</td>
</tr>
<tr>
<td>5</td>
<td>3.17</td>
<td>0.12</td>
<td>0.000008</td>
</tr>
<tr>
<td>6</td>
<td>2.93</td>
<td>0.15</td>
<td>0.058980</td>
</tr>
<tr>
<td>7</td>
<td>3.26</td>
<td>0.13</td>
<td>0.007594</td>
</tr>
<tr>
<td>total</td>
<td>22.2100</td>
<td>0.8800</td>
<td>0.4175</td>
</tr>
<tr>
<td>mean</td>
<td>3.1729</td>
<td>0.1257</td>
<td>0.0596</td>
</tr>
</tbody>
</table>

From Table 6.2, the results are, \( \bar{t} \) of 3.17 minutes and standard deviation of 0.24 minutes and for \( D/uL \), the corresponding values are the mean of 0.126 and standard deviation of 0.012. Likewise, these results are considered to be well within the experimental error and by comparison with the single phase flow, the reproducibility for the two phase experiments are better than they are for the single phase.
The difference between the two results can be explained by the different methods of introducing the tracer, as described in Section 5.1.3. That is, for the single phase case, the tracer was introduced by pouring the hot water into the channel, whereas for the two phase case, it was by the plunging of the heater into the channel.

6.2. SINGLE PHASE EXPERIMENTS

6.2.1. MEAN TIME

Figure 6.1 shows the plot of t against liquid flowrate, both experimental and theoretical, as a result of calculation as shown in Table 5.3. Theoretical mean times are plotted in the picture with the line that connects each value, while the experimental mean times are plotted as scattered dots. The graph clearly shows that the experimental mean times are in good agreement with the theoretical mean time.

Based on compartment model theory [Levenspiel, 1993], that:

\[ t_{\text{experiment}} = \frac{V_{\text{active}}}{\text{flowrate}} \] (6.1)

then from the calculation of the results in Table 5.3 it is found that \( V_{\text{active}} \) of these particular experiments is approximately 85 % of \( V_{\text{reactor}} \).
The above finding of the active volume being about 85 % of the total reactor volume is a reasonable result suggesting that the remainder (15 %) is a so-called dead volume. In reality, such a volume is semi-stagnant, indicating that the fluid moves through it very slowly. Of course, there is always the possibility that the above result is simply due to the measurement error of the mean time. In fact, according to the reproducibility results in Table 6.1, a 15 % error of measurement has a probability of occurrence of \( p = 0.18 \).

It may be of interest to compare the above result with that of Sohn et.al. [1992] using their response curve in Figure 4.1. Measurements show an experimental mean time of 68 seconds and a calculated mean time of 60 seconds, using the values of \( L = 120 \) cm and \( u = 2 \) cm/s given in Figure 4.1. The difference is also
about 15%. However, without any statement by Sohn et.al. [1992] as to the magnitude of error of their measurements, it is not possible to make a conclusion either way, namely, is this difference (15%) an error of measurement or is it a stagnant volume in the reactor. Therefore, it is reasonable to assume that the 15% difference in the present experiment might be the case of stagnant volume.

6.2.2. DISPERSION

Figure 6.2 shows the results of D/uL for all conditions of liquid and gas flowrates used, calculated by equation (2.15) using the variance difference obtained from equation (2.16) with experimental variances calculated by equation (2.13). The plots for each particular condition can be seen in Figures 5.8 to 5.13.

It is evident in Figure 6.2 that the values of D/uL in single phase experiments are scattered within the range of D/uL of 0.14 - 0.19 and it is very difficult to be precise in determining the trend consistency for each different experimental condition. Nevertheless, there is an upward trend of these values, both when examining some single graphs (Figures 5.8 to 5.13) or when examining the collected data (Figure 6.2). On examining the values in Table 5.4, it is interesting to find out that for low flowrates (below Reynolds number of 900), D/uL mostly increases as the bubbler position is deeper and gas flowrate is higher. For Re above 900, it is difficult to determine such a trend, because the
phenomena do not appear as consistent as in the lower flowrates and the values are scattered.

Figure 6.2. Plot of $D/u_L$ against Reynolds number for all conditions of single phase experiments.

In the light of the single phase result of Sohn et.al. [1992] in Figure 4.3, showing a clear downward trend of $D/u_W$ versus Reynolds numbers, that in Figure 6.2 is disappointing. Because of this the results were reexamined in an attempt to find an explanation. Because of the problem of mean times in the two phase flow to be discussed in the next section, the $D/u_L$ value were recalculated, but no change emerged. It is therefore considered that the results in Figure 6.2 are correct within the experimental error given in Table 6.1, namely, a standard deviation of approximately 0.023, or on average a relative difference of about 20%. Therefore, although there is a small upward trend of $D/u_L$ with Reynolds numbers in Figure 6.2, reminiscent of that in Figure 2.9, the above relative error of 20% is too large to make any such valid conclusions.
6.3. COUNTERCURRENT EXPERIMENTS

6.3.1. MEAN TIME

The experimental mean times in Tables 5.5 and 5.6, calculated by equation (2.12), are very much different from those calculated by equation (2.9). By inspection of the two values in Tables 5.5 and 5.6, it can be seen that the difference is about 2 to 5 times at high flowrates and a one order of magnitude or more, at low flowrates.

In order to understand why there is this difference, the experimental response curves were re-examined, and it was typical that for the low flowrates, the response curves exhibited a very long tail. From theory (Section 2.3.2), the long tail of the response curve is a spurious result as it is not a property of the main signal but is caused by either adsorption of the tracer and slow release thereafter or in this case, a loss of heat into the surroundings.

In view of the above result, it was considered necessary to examine the curves from peak to peak which it is known would, in these circumstances, better represent the true travelling time [Levenspiel, 1993]. When the peak to peak method is used for calculating the \( t \), the results are as shown in Tables 6.3 and 6.4. By inspection of the results in those two tables, plotted as Figures 6.3 and 6.4, it is now obvious that the theoretical and the experimental results are in very much better agreement, especially at the high flowrates.


Figure 6.3. Plot of experimental mean times (peak to peak method) and its theoretical mean times against liquid flowrate for slow slag moving counter current experiments.

Figure 6.4. Plot of experimental mean times (peak to peak method) and its theoretical mean times against liquid flowrate for fast slag moving counter current experiments.
It is of interest to see if anything can be gleaned about this dilemma from the results of Sohn et.al. [1992] as was done in the case of single phase flow in section 6.2.1. Using the above authors response curve for the two-phase flow in Figure 4.2 together with their experimental statement therein, the result is a measured mean time of 1.94 minutes and a calculated mean time of 0.75 minutes, a difference of about 2.5 times.

From the above result, it is clear that the Sohn et.al. [1992] typical response curve for their two-phase experiments, also gives a mean time value much different from the calculated one. So in this respect, it can be said that the present results and Sohn et.al. [1992] results, are in agreement. Of course, the conclusion that the two sets of results agree with respect to a discrepancy between measured and calculated mean times does not resolve the dilemma of the discrepancy. Clearly, further investigations, with preferably a different measuring technique, is strongly indicated.

6.3.2. DISPERSION

Figures 6.5 and 6.7 show the results of D/uL for all conditions of liquid and gas flowrates used, for slow slag moving and fast slag moving, respectively, calculated by equation (2.15) using the variance difference obtained from equation (2.16) with experimental variances calculated by equation (2.13).
It is obvious from Figure 6.5 that the $D/u_L$ at first increases with Reynolds number, reaches a maximum at Reynolds number of about 1100 and then decreases again. If this result is compared with that reported by Sohn et.al. [1992] in Figure 4.3, it is seen that their results only show the decreasing trend and there is no upward part as in Figure 6.5.

![Graph](image)

Figure 6.5. Plot of $D/u_L$ against Reynolds number for all conditions of counter current experiments with slow slag moving.

On the other hand, $D/ud$ in pipes in Figure 2.10 shows that for laminar flow, $D/ud$ increases, reaches a maximum at Reynolds number = 2100 and then decreases in the turbulent flow region exponentially with increasing Reynolds number. Considering the relationship between $D/ud$ and $D/u_L$ in equation (2.10) then, for otherwise constant conditions, it is valid to conclude that the same pattern will be followed for $D/u_L$ also as that for $D/ud$ in Figure 2.9.
In view of the earlier conclusion regarding $\bar{t}$, namely, $\bar{t}$ is more appropriately calculated by the peak to peak method, especially for low flow rates or low Reynolds numbers, a check was made to see if these $\bar{t}$ values (Table 6.3), if used in the calculation of $D/uL$, would change the trend shown in Figure 6.5. The so calculated $D/uL$ values are shown in Table 6.5 and plotted in Figure 6.6 as a function of Reynolds number.

Table 6.5. $D/uL$ results for counter current experiments with slow slag moving using peak to peak method of mean time calculation.

<table>
<thead>
<tr>
<th>Reynolds Number</th>
<th>0/1500</th>
<th>2/1500</th>
<th>2/800</th>
<th>2/2000</th>
</tr>
</thead>
<tbody>
<tr>
<td>342</td>
<td>0.11</td>
<td>0.11</td>
<td>0.11</td>
<td>0.1</td>
</tr>
<tr>
<td>513</td>
<td>0.09</td>
<td>0.09</td>
<td>0.12</td>
<td>0.07</td>
</tr>
<tr>
<td>684</td>
<td>0.12</td>
<td>0.13</td>
<td>0.13</td>
<td>0.13</td>
</tr>
<tr>
<td>855</td>
<td>0.16</td>
<td>0.17</td>
<td>0.18</td>
<td>0.14</td>
</tr>
<tr>
<td>1026</td>
<td>0.13</td>
<td>0.12</td>
<td>0.14</td>
<td>0.14</td>
</tr>
<tr>
<td>1197</td>
<td>0.13</td>
<td>0.15</td>
<td>0.13</td>
<td>0.12</td>
</tr>
<tr>
<td>1368</td>
<td>0.13</td>
<td>0.14</td>
<td>0.14</td>
<td>0.13</td>
</tr>
<tr>
<td>1538</td>
<td>0.11</td>
<td>0.11</td>
<td>0.12</td>
<td>0.1</td>
</tr>
<tr>
<td>1709</td>
<td>0.06</td>
<td>0.08</td>
<td>0.04</td>
<td>0.07</td>
</tr>
</tbody>
</table>

By inspection of Figure 6.6, it is quite clear that the same trend prevails as in Figure 6.5. The difference is that the values of $D/uL$ are now about twice those shown in Figure 6.5. The maximum $D/uL$ again appears at about the same Reynolds number as in Figure 6.5.
The reason why the transition Reynolds number at which the maximum $D/uL$ occurs is lower than that shown in Figure 2.10 is almost certainly caused by the difference in the flow geometry. In the case of Figure 2.10 it is a pipe, and in the case of Figure 6.6 it is a channel.

![Figure 6.6. Plot of $D/uL$ against Reynolds number for all conditions of counter current experiments with slow slag moving using peak to peak method of mean time calculation.](image)

Figure 6.6. Plot of $D/uL$ against Reynolds number for all conditions of counter current experiments with slow slag moving using peak to peak method of mean time calculation.

Figure 6.7 shows a scattered $D/uL$ plot against Reynolds number for counter current experiments with fast slag moving, which are difficult to interpret. Similarly to the steps taken in the preceding case, the response curves were re-examined first. It was apparent that, on average, the curves for the fast slag experiments were discernibly different from those of the slow slag.
experiments. This almost certainly could explain the difference between the results for the two sets of conditions in Figures 6.5 and 6.7.

![Figure 6.7. Plot of D/uL against Reynolds number for all conditions of counter current experiments with fast slag moving.](image)

Next, as in the preceding case, the peak to peak mean times in Table 6.4 were used to recalculate the D/uL values. The values so calculated are shown in Table 6.6, and plotted against Reynolds number in Figure 6.8. On observing Figure 6.8, it is found that similar trend to that with slow slag in Figures 6.5 and 6.6 happens in fast slag moving counter current experiments also, namely, the values of D/uL increase at first with increasing Reynolds number, reach a maximum at Reynolds number of about 900, then decrease.
Table 6.6. D/uL results for counter current experiments with fast slag moving using peak to peak method of mean time calculation.

<table>
<thead>
<tr>
<th>Reynolds Number</th>
<th>0/1500</th>
<th>2/1500</th>
<th>2/800</th>
</tr>
</thead>
<tbody>
<tr>
<td>342</td>
<td>0.18</td>
<td>0.2</td>
<td>0.2</td>
</tr>
<tr>
<td>513</td>
<td>0.12</td>
<td>0.26</td>
<td>0.19</td>
</tr>
<tr>
<td>684</td>
<td>0.27</td>
<td>0.28</td>
<td>0.23</td>
</tr>
<tr>
<td>855</td>
<td>0.26</td>
<td>0.27</td>
<td>0.3</td>
</tr>
<tr>
<td>1026</td>
<td>0.26</td>
<td>0.27</td>
<td>0.16</td>
</tr>
<tr>
<td>1197</td>
<td>0.16</td>
<td>0.18</td>
<td>0.26</td>
</tr>
<tr>
<td>1368</td>
<td>0.18</td>
<td>0.22</td>
<td>0.25</td>
</tr>
<tr>
<td>1538</td>
<td>0.21</td>
<td>0.22</td>
<td>0.23</td>
</tr>
<tr>
<td>1709</td>
<td>0.13</td>
<td>0.14</td>
<td>0.17</td>
</tr>
</tbody>
</table>

Figure 6.8. Plot of D/uL against Reynolds number for all conditions of counter current experiments with fast slag moving using peak to peak method of mean time calculation.
It is known that the Reynolds number for the transition from laminar to turbulent flow is different for different vessel geometries. For example, in the case of a packed bed, its value is 10. In the case of a channel, the transition occurs at the Reynolds number of about 1400 [Murphy, 1950] and in the case of coils, it occurs at 6000 [Levenspiel, 1993].

All of these transition the Reynolds numbers are for a developed flow, that is without any other interfering flow or disturbances, as is the case in the present instance of a bubbling gas. It is possible that these interferences by the gas flow somehow make the maximum of $D/uL$ occur at these lower Reynolds number. However, it is not immediately obvious as to the mechanism that may be involved in this case, and further research is clearly indicated.

The fact that in Figure 4.3, Sohn et.al. [1992] did not observe such a pattern as is shown in Figures 6.5 to 6.8, is difficult to explain. It may be due to the fact that the bubbling was carried out from the bottom rather than from the top, as in the present case. It may also be that in calculating the Reynolds number in these systems, emulsion density should be used rather than the density of the liquid as such, and this will, of course, lower the value of the transition Reynolds number. However, as in the previous comment, it can not be stated with certainty that this is the cause, and once again, further research is indicated to resolve this apparent anomaly.

In this context, it may be of interest to present the results obtained in the same experimental channel as used in the present work, but without gas bubbling. These results [Bohlsen, 1995] are given in Table 6.7 and shown in Figure 6.9,
for both single and two phase experiments. The values given here are different with the values that Bohlsen [1994] reported, because they were recalculated from the raw experimental data with different calculation method due to the reasons that has been discussed earlier.

Table 6.7. D/\(uL\) results of no bubbling experiments [Bohlsen, 1994].

<table>
<thead>
<tr>
<th>Reynolds number</th>
<th>D/(uL) counter current</th>
<th>D/(uL) single phase</th>
</tr>
</thead>
<tbody>
<tr>
<td>171</td>
<td>0.082</td>
<td>0.051</td>
</tr>
<tr>
<td>342</td>
<td>0.203</td>
<td>0.069</td>
</tr>
<tr>
<td>513</td>
<td>0.099</td>
<td>0.111</td>
</tr>
<tr>
<td>684</td>
<td>0.23</td>
<td>0.15</td>
</tr>
<tr>
<td>855</td>
<td>0.259</td>
<td>0.156</td>
</tr>
<tr>
<td>1025</td>
<td>0.284</td>
<td>0.159</td>
</tr>
<tr>
<td>1197</td>
<td>0.117</td>
<td>0.127</td>
</tr>
</tbody>
</table>

Figure 6.9. Plot of D/\(uL\) against Reynolds number of no bubbling experiments [Bohlsen, 1994].
Figure 6.9 clearly shows that for both cases, the $D/u_L$ increases with Reynolds number, reaches a maximum at about $Re = 1000$ and then drops off. Unfortunately, due to pumping restrictions, higher Reynolds number could not be obtained. Nevertheless, Figure 6.9 is considered to provide support for a suggestion that $D/u_W$, and by equation (2.10) also $D/u_L$, should parallel Figure 2.12.

### 6.3.3. EFFECT OF GAS FLOWRATE ON $D/UL$

Statistical regression method was used to assess the effect of gas flowrate ($Q$) on $D/u_L$. The results of the statistical regression calculation are given in Tables 6.8 to 6.11. The calculation has been done separately for high and low liquid flowrates due to different trends involved. Tables 6.8 and 6.9 are the results for counter current experiments with slow slag moving, for low and high flowrates, respectively. Tables 6.10 and 6.11 are for counter current experiments with fast slag moving, for low and high flowrates, respectively.

Table 6.8. Statistical regression calculation results of slow slag moving experiments, low liquid flowrates, for different gas flowrates.

<table>
<thead>
<tr>
<th>$Q$</th>
<th>Coefficient</th>
<th>$P$</th>
<th>Slope</th>
<th>$P$</th>
<th>$R$</th>
<th>standard error</th>
</tr>
</thead>
<tbody>
<tr>
<td>2/ 800</td>
<td>0.058</td>
<td>0.15</td>
<td>129 E-6</td>
<td>0.086</td>
<td>0.91</td>
<td>0.015</td>
</tr>
<tr>
<td>2/ 1500</td>
<td>0.048</td>
<td>0.336</td>
<td>129 E-6</td>
<td>0.168</td>
<td>0.83</td>
<td>0.023</td>
</tr>
<tr>
<td>2/ 2000</td>
<td>0.047</td>
<td>0.39</td>
<td>105 E-6</td>
<td>0.265</td>
<td>0.73</td>
<td>0.026</td>
</tr>
</tbody>
</table>
Table 6.9. Statistical regression calculation results of slow slag moving experiments, high liquid flowrates, for different gas flowrates.

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>P</th>
<th>Slope</th>
<th>P</th>
<th>R</th>
<th>standard error</th>
</tr>
</thead>
<tbody>
<tr>
<td>2/ 800</td>
<td>0.286</td>
<td>0.004</td>
<td>-125 E-6</td>
<td>0.026</td>
<td>0.86</td>
</tr>
<tr>
<td>2/ 1500</td>
<td>0.233</td>
<td>0.003</td>
<td>-82 E-6</td>
<td>0.045</td>
<td>0.82</td>
</tr>
<tr>
<td>2/ 2000</td>
<td>0.215</td>
<td>0.001</td>
<td>-77 E-6</td>
<td>0.015</td>
<td>0.9</td>
</tr>
</tbody>
</table>

Table 6.10. Statistical regression calculation results of fast moving experiments, low liquid flowrates, for different gas flowrates.

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>P</th>
<th>Slope</th>
<th>P</th>
<th>R</th>
<th>standard error</th>
</tr>
</thead>
<tbody>
<tr>
<td>2/ 800</td>
<td>0.162</td>
<td>0.172</td>
<td>87 E-6</td>
<td>0.488</td>
<td>0.72</td>
</tr>
<tr>
<td>2/ 1500</td>
<td>0.127</td>
<td>0.176</td>
<td>234 E-6</td>
<td>0.179</td>
<td>0.96</td>
</tr>
</tbody>
</table>

Table 6.11. Statistical regression calculation results of fast moving experiments, high liquid flowrates, for different gas flowrates.

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>P</th>
<th>Slope</th>
<th>P</th>
<th>R</th>
<th>standard error</th>
</tr>
</thead>
<tbody>
<tr>
<td>2/ 800</td>
<td>0.286</td>
<td>0.009</td>
<td>-48 E-6</td>
<td>0.43</td>
<td>0.36</td>
</tr>
<tr>
<td>2/ 1500</td>
<td>0.368</td>
<td>0.0003</td>
<td>-119 E-6</td>
<td>0.017</td>
<td>0.84</td>
</tr>
</tbody>
</table>

Tables 6.8 to 6.11 give the usual parameters of a linear regression which was used because the small number of points and their scatter did not justify the use of anything more complex, as for example, non-linear regression. The parameters in Tables 6.8 to 6.11 are those of an equation of a straight line with statistical estimates of confidence. Specifically, P is the probability of the
tabulated value of coefficient (intercept) and slope being equal to the stated value and $R$ is the regression coefficient with its standard error.

In the above, the usual convention applies, i.e. the smaller the $P$ value is, the more certain is the stated value, and, the lower does $R$ approach unity the better is the fit of the data to the regression line or equation. Taking Tables 6.8 and 6.9 as an example, it is seen that while the $R$ values are about the same, the $P$ values in Table 6.9 are one to two orders of magnitude smaller than those in Table 6.8. Similar result is also evident in Tables 6.10 and 6.11.

It should also be noted that in accord with the results in Figures 6.6 and 6.8, low liquid flowrates (Tables 6.8 and 6.10) are associated with a positive slope and high liquid flowrates (Tables 6.9 and 6.11) are associated with a negative slope.

Using the results in Tables 6.8 to 6.11, the effect of gas flowrate ($Q$) on $D/uL$ was calculated from the usual functional relationship assumption, i.e. $D/uL \propto Q^a$. The results are presented in Table 6.11 that suggest that the effect of gas flowrate is to decrease the vessel dispersion number ($D/uL$), although, realistically, an effect of an order of 0.04, whether positive or negative, would be questionable as an effect at all. This result can therefore be said to be approximately in agreement with that of Sohn et.al. [1992], namely, that the effect of gas flowrate in the counter current case is small. Noting that these authors used bottom gas bubbling, as against top gas bubbling in the present work, could be sufficient to account for a lack of a complete agreement.
Table 6.12. Values of exponent a.

<table>
<thead>
<tr>
<th>Condition</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>slow slag, low Re</td>
<td>-0.22</td>
</tr>
<tr>
<td>slow slag, high Re</td>
<td>-0.04</td>
</tr>
<tr>
<td>fast slag, low Re</td>
<td>0.27</td>
</tr>
<tr>
<td>fast slag, high Re</td>
<td>-0.04</td>
</tr>
</tbody>
</table>

In the case of low gas flowrates the effect of an order of 0.25 in Table 6.11, would normally be regarded as significant. The question in the present case is, if the effect is significant, why is it significant in the opposite sense for slow slag and for fast slag case. Theoretically, D/uL is the dispersion number of a vessel for a given set of flow parameters, including gas bubbling, and should not be affected by anything else. Specifically, it should not be affected by the presence of a slag phase whether slow moving or fast, at least within the range used in the present case. The reason for the latter restriction is that:

1) the 'slag/metal interface', i.e. the vessel geometry will be essentially the same in both cases, and

2) the velocity term (u) will not be significantly changed as a result of its relative motion with respect to the 'slag' flow.

From the above, therefore, it should be expected that the effect of gas flowrate on D/uL for low liquid rates should not be too different from its effect at high liquid rates at which also, as theoretically expected, the slag rate has no influence (Table 6.12).
On combining all of the above considerations, the only reasonable explanation seems to be that of error. This conclusion is supported by the very high P-values for the low liquid flowrates (Tables 6.8 and 6.10) compared with the very low P-values for the high liquid flowrates (Tables 6.9 and 6.11).

6.3.4. EFFECT OF BUBBLER POSITION ON D/UL

Similarly to the evaluation of the effect of gas flowrate on D/uL, statistical regression method was used to evaluate the effect of bubbler depth on D/uL. The regression parameters are given in Tables 6.13 to 6.16. It is noted that discussions regarding statistical evaluation and confidence level presented in Section 6.3.3 apply without change in this case also. Hence, they will not be restated. The values of the exponent b in the functional relationship $D/uL \propto P^b$, where P is the bubbler depth below the surface, are given in Table 6.17.

Table 6.13. Statistical regression calculation results of slow slag moving experiments, low liquid flowrates, for different bubbler depth.

<table>
<thead>
<tr>
<th>Depth</th>
<th>Coefficient</th>
<th>P</th>
<th>Slope</th>
<th>P</th>
<th>R</th>
<th>Standard error</th>
</tr>
</thead>
<tbody>
<tr>
<td>0/1500</td>
<td>0.057</td>
<td>0.258</td>
<td>105E-6</td>
<td>0.211</td>
<td>0.79</td>
<td>0.022</td>
</tr>
<tr>
<td>2/1500</td>
<td>0.048</td>
<td>0.336</td>
<td>129E-6</td>
<td>0.168</td>
<td>0.83</td>
<td>0.023</td>
</tr>
</tbody>
</table>
Table 6.14. Statistical regression calculation results of slow slag moving experiments, high liquid flowrates, for different bubbler depth.

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>P</th>
<th>Slope</th>
<th>P</th>
<th>R</th>
<th>standard error</th>
</tr>
</thead>
<tbody>
<tr>
<td>0/ 1500</td>
<td>0.24</td>
<td>0.001</td>
<td>-94 E-6</td>
<td>0.016</td>
<td>0.89</td>
</tr>
<tr>
<td>2/ 1500</td>
<td>0.233</td>
<td>0.003</td>
<td>-82 E-6</td>
<td>0.045</td>
<td>0.82</td>
</tr>
</tbody>
</table>

Table 6.15. Statistical regression calculation results of fast slag moving experiments, low liquid flowrates, for different bubbler depth.

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>P</th>
<th>Slope</th>
<th>P</th>
<th>R</th>
<th>standard error</th>
</tr>
</thead>
<tbody>
<tr>
<td>0/ 1500</td>
<td>0.055</td>
<td>0.819</td>
<td>263 E-6</td>
<td>0.593</td>
<td>0.6</td>
</tr>
<tr>
<td>2/ 1500</td>
<td>0.127</td>
<td>0.176</td>
<td>234 E-6</td>
<td>0.179</td>
<td>0.96</td>
</tr>
</tbody>
</table>

Table 6.16. Statistical regression calculation results of fast slag moving experiments, high liquid flowrates, for different bubbler depth.

<table>
<thead>
<tr>
<th>Coefficient</th>
<th>P</th>
<th>Slope</th>
<th>P</th>
<th>R</th>
<th>standard error</th>
</tr>
</thead>
<tbody>
<tr>
<td>0/ 1500</td>
<td>0.36</td>
<td>0.0005</td>
<td>-125 E-6</td>
<td>0.019</td>
<td>0.84</td>
</tr>
<tr>
<td>2/ 1500</td>
<td>0.368</td>
<td>0.0003</td>
<td>-119 E-6</td>
<td>0.017</td>
<td>0.84</td>
</tr>
</tbody>
</table>

Table 6.17. Values of exponent b.

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>slow slag, low Re</td>
<td>0.06</td>
</tr>
<tr>
<td>slow slag, high Re</td>
<td>0.11</td>
</tr>
<tr>
<td>fast slag, low Re</td>
<td>0.39</td>
</tr>
<tr>
<td>fast slag, high Re</td>
<td>0.11</td>
</tr>
</tbody>
</table>

The results in Table 6.17 suggest that the effect of bubbler depth is opposite in the slow and fast slag case. In the former, D/uL increases with bubbler depth.
and in the latter, it decreases. As there is no obvious reason, either theoretical or intuitive why this should be so, the difference is considered to be very likely an error. This is supported by the P-values in Tables 6.13 to 6.16, in the same way as for the effect of gas flowrate in Tables 6.8 to 6.11.

The above therefore means that the effect of bubbler depth is to increase \( D/u_L \). It is also supported by Treverrow [1995]. This conclusion is reasonable as in the limit the situation would be same as bottom bubbling and this is known to lead to a complete plume which sets up circulating motion of the liquid over the full height of the bath. In turn these circulations would be expected to increase mixing and by extension, dispersion also.

### 6.4. CONCLUDING REMARKS

As stated in Section 5.1, this work was specifically modelled on the dispersion part of that of Sohn et.al. [1992]. In particular, the channel dimension, the liquids used, the flow arrangements, and the thermal stimulus-response technique employed were identical. The only difference was the use of top gas bubbling instead of bottom gas bubbling as used by Sohn et.al. [1992]. The reason for this difference was that in the WORCRA furnace, which it was wished to highlight, top gas bubbling was used. Because Sohn et.al. [1992] reported good results and did not report any experimental problems, it was expected that the present work would yield similar outcome. However, as is clear from the results reported here, this was not the case.
When the work was carried out, everything seemed normal; there were no undue problems and the thermal tracer technique described by Sohn et.al. [1992] as ‘a novel tracer technique’ seemed to give reasonable response, even though the input signal was often far from an ideal pulse. This, as noted in Section 2.3.2, was corrected by using the variance difference in calculating the D/ul values.

According to Sohn et.al. [1992], this technique ‘was developed to measure eddy diffusivities in the two phases and to simulate mass transfer using the heat transfer’. Although, as summarised in Chapter 4, their dispersion results in the low gas flow studies using the thermal technique were claimed to be very good, they used a conductivity technique employing a concentrated KCl as the tracer in their high gas flow studies and RTD modelling, claiming ‘an excellent agreement’ between prediction and experiment. Of course, the use of thermal experiments is mandatory in studying heat transfer, using heat as a tracer in studying dispersion may be questionable, at least without appropriate corrections.

Unlike for other tracers listed in Table 2.1, tracer balance for a thermal tracer is an impossible in practice, because of heat loss between the point of ‘injection’ and its response at the outlet. If there is no tracer balance then the mathematics normally used for calculating RTD will give results which are in error that can be quite large [Levenspiel, 1993]. Additionally, proper boundary conditions must be used, otherwise more errors are introduced and tmeasured is greater than tcalculated. In fact, a result of tmeasured > tcalculated is a sure sign that there is something wrong [Levenspiel, 1993]. It is of interest to
remark that calculations of Sohn et.al. [1992] response curve in Figure 4.3 gives such a result.

Because of an appreciation of these kinds of errors, the calculations in the present thesis have included the so-called 'sloppy pulse' method of calculating \( D/uL \) with correction for the mean time obtained by the 'peak-to-peak' method, as well as calculating \( D/uL \) values independently by the 61 \% method, as a check on consistency. It is not claimed that the so calculated values are free of 'lack of tracer balance' error. They are nevertheless considered to be more correct than if the normally used calculating methods were employed.

It is believed that if the thermal tracer technique is to be used again for dispersion studies, it should be preceded by an investigation designed to establish correction factors for the many variables involved, including ambient temperature which may be significantly different between winter and summer. Of course, if the thermal tracer technique can be avoided, then, based on the knowledge obtained in the present work, it should be avoided and one or the other techniques listed in Table 2.1 be used instead.
The results of the investigation reported in the present work, involving a laboratory model of a channel reactor, have shown that:

1) For countercurrent flow of TCE and water, simulating metal and slag flow, the vessel dispersion number, $D/u_L$, increases with Reynolds number, reaches a maximum at a Reynolds number of about 900 and then it decreases. This behaviour was observed for all gas bubbling rates employed (800 - 2000 cm$^3$/min.)

2) For single phase flow, the vessel dispersion number, $D/u_L$, for all conditions employed, i.e. liquid flowrates of 0.10 - 0.52 cm/s and gas bubbling rates of 800 - 4000 cm$^3$/min., appeared independent of Reynolds number and within the experimental error of the measurement. This unexpected result was attributed to a possible effect of introducing the thermal tracer in a different way to that in the counter current experiments.

3) The average effect of the bubbling rate and the bubbler position in the two phase work, assessed by regression analysis, was found to be described, with different degrees of confidence, as:

   i) slow slag, low Re : $D/u_L \propto Q^{-0.22} \pm 0.06$
ii) slow slag, high Re : \( \frac{D}{uL} \propto Q^{-0.04} p^{0.11} \)

iii) fast slag, low Re : \( \frac{D}{uL} \propto Q^{0.27} p^{0.39} \)

iv) fast slag, high Re : \( \frac{D}{uL} \propto Q^{-0.04} p^{0.11} \)

4) The mean residence times for the single phase experiments were in good agreement with the calculated values but in the two phase experiments there was a large difference between the two values, attributed to increased heat losses from the systems.

5) The use of a thermal tracer appears to inherently problematic. A change to another tracer for future studies is recommended.
REFERENCES


References


