A study of flow interruptions in the HYL III reactor

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A STUDY OF FLOW INTERRUPTIONS IN THE HYL III REACTOR

A thesis submitted in fulfillment of the requirements for the award of the degree

Honors Master of Engineering

from

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by

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SYNOPSIS

Since January 1994, PT Krakatau Steel has been operating a new Direct Reduction plant with the Hyl III process, a continuous moving bed reactor. During the operation, some problems emerged regarding flow interruptions at the outlet of reactor.

It is quite difficult to observe the reason why the flow interruption occurs in actual operation because it requires considerable cost and will certainly disturb the production. Therefore, developing and implementing a model (one tenth of the actual size) representing the actual reactor is a method to simplify the observation.

Three possibilities have been identified as to why flow interruption occurs at the Hyl III reactor. The first possibility is the arching phenomenon. Arching is characterized using the arching equation to find the minimum diameter of arching at the Hyl III reactor outlet. The minimum diameter of arching is then compared to the actual diameter of the outlet. In the same manner, the arching minimum diameter in the physical model is calculated too and then compared to the actual diameter of the model outlet.

The second possibility of flow interruption is due to fluidization. The existence of fluidization was identified by measuring the actual minimum pressure drop in the Hyl III reactor compared to the calculated pressure in the actual Hyl III reactor. Additionally, several experiments have been conducted using the Hyl III reactor model. The pressure drop at the reactor model is measured for various
parameters such as size distributions, gas velocity and solid velocity. The measured pressure drop reflects the type of solid flow: stick slip flow or aeration flow (fluidised). The results show that the flow interruption (clogging) did not occur under this condition (stick slip and aerated flow). Aerated flow condition is the same condition of fluidization.

The last possibility is the effect of agglomerate formation. To identify agglomerate formation in the outlet Hyl III reactor, observations were made in the Hyl III reactor model. The objective was also to examine the various proportions of the agglomerate and sand mixture either with pressured air or without pressured air.

The present study has shown that in the Hyl III reactor, both arching and fluidization did not occur. On the other hand, a great number of agglomerates are formed. These agglomerates choked the outlet of the Hyl III reactor resulting in clogging. This is in accordance with the experiment result obtained in the Hyl III model reactor. The clogging is caused by an abundance of agglomerates blocking the outlet model reactor.

Basically, a few agglomerates did not cause clogging, however a large number of agglomerates do because they contact one another, leading to interlocking and in turn, the outlet is choked. The clogging occurrence can be prevented using some air in that air pressure can reduce the velocity of agglomerates falling to the bottom of the reactor toward the outlet. Therefore, if only one or two agglomerates contact each other, that does not cause clogging. However, should the size of agglomerates increase, the air pressure is not able to
reduce the agglomerate velocity. The number of agglomerates keeps increasing and the agglomerates get interlocked resulting in clogging.

Therefore, the best solution to prevent the flow interruption is by preventing the occurrence of agglomerates formation.
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Reproducibility
NOTATION

$\alpha l$ : hopper half angle

$B$ : diameter of circular outlet area

$C$ : the length scale (the ratio of a pertinent distance on the prototype to the corresponding distance on the model)

$\bar{d}$ : sample mean

$\bar{\delta}$ : mean particle diameter

$di$ : data of $i^{th}$ sample

$dt$ : effective diameter

$\delta$ : The effective angle of internal friction.

$\Delta P$ : pressure drop

$\Delta u$ : linear velocity of gas relative to solids

$\varepsilon$ : voidage

$ff$ : flow factor

$g$ : gravitational constant

$H(\theta)$ : Function that depends on the hopper type and angle as shown in Figure 2.8

$\gamma$ : bulk density

$\phi$ : kinematics angle of friction between the solid and the bin wall
\( \phi_s \) : sphericity

\( M_p \) : particle mass

\( n_d \) : number of data

\( \rho_s \) : density of solid

\( \rho_g \) : density of gas

\( \sigma_1 \) : the major consolidating pressure at a point in the hopper

\( \sigma_1 \) : the minimum stress in the abutment of an arch at this point

\( T \) : uniform thickness

\( \tau \) : Stress component in the shear direction.

\( U_0 \) : superficial velocity of gas

\( U_s \) : velocity of solids

\( U_{mf} \) : superficial velocity at minimum fluidization

\( V_p \) : particle volume

\( \mu \) : Viscosity of fluid.
CHAPTER ONE

INTRODUCTION

1.1 Introduction

PT Krakatau Steel, since January 1994, has been operating a new Direct Reduction plant with the Hyl III process, a continuous moving bed reactor. This process is different from the previous Direct Reduction plant which uses a fixed bed reactor.

After two years operating, several problems emerged, such as: charging material to reactor, metal dusting in the gas heater, cooling plenum, and reduction quench tower, up to flow interruption problem at the outlet Hyl III reactor. Metal dusting and charging material problems have been under researched by an expert team in PT Krakatau Steel. Since the two frequent problems have been taken care of by the team, this research is directed to solve the flow interruption problem at the outlet Hyl III reactor.

To observe the cause of flow interruption in the actual Hyl III reactor is quite difficult because of the cost, and it disturbs production. Therefore, the problem can be investigated using a physical model of the Hyl III reactor. Using this model experiments can be conducted to observe the cause of flow interruption.
1.2 Overview

Flow interruption at the Hyl III reactor is possibly caused by arching, fluidization, or agglomerates formation.

The first possibility, flow interruption, is caused by arching. Arching in mass flow bins can be due to the interlocking of a few particles that are large with respect to the outlet or due to the formation of a cohesive arch (Carson and Johanson, 1985). To overcome the first type of arching, the outlet size of a circular bin should be at least five to six times the size of the largest particles if a circular outlet is used, or three to four times (i.e. minimum width) if rectangular outlet is used. Flow interruption caused by arching has certainly already been considered as a possibility in the basic engineering calculations by Engineers HYLSA. So there is just little possibility that flow interruption is caused by arching.

The second possibility, flow interruption is caused by fluidization. Fluidization is the operation by which fine solids are transformed into a fluid-like state through contact with gas or liquid [Kunii and Levenspiel, 1977]. The Hyl III is a continuous moving bed reactor. This is not a fluidized bed system, therefore, operating conditions do not match the requirements of a fluidized bed. So that there is just a little possibility that flow interruptions is caused by fluidization.

The third possibility is flow interruption is caused by agglomerate formation. While clogging occurs at the charge or discharge reactor Hyl III, solid on the reactor received hotter reducing gas resulting in some hot spots throughout the reactor and finally, some solid undergo agglomerate formation. Agglomerate
formation occurs for pellets that have a sticking index above 30% (recommended by HYLSA). Sticking index is calculated as the area under the resulting curve of the percentage of clusters as a function of the number of drops. Agglomerate formation is caused by fault operation reduction. Long or short operation reductions depend on sticking index, short operation reduction if sticking index is high and vice versa. The frequency of problems influences agglomerate formation and the possibility that flow interruptions cause agglomerate formation is high.

1.3 Research Question

The three possible causes of the flow interruption are to be researched. In any research, there is always at least one question that one is trying to answer. Therefore, this research is trying to answer the major questions:

- What are the actual causes of flow interruption?
- How can the flow interruption be prevented?

1.4. Research Methodology

The research began with a literature review at the preliminary state. The literature review was the primary method to gain basic knowledge and an overview. To prove flow interruption caused by arching, it is required to calculate the minimum outlet diameter of Hyl III reactor causing arching using the arching equation and then it is compared to the existing outlet Hyl III reactor. Arching occurs if the calculated minimum diameter of the outlet Hyl III reactor is bigger.
than the existing outlet diameter of Hyl III reactor. The above step is carried out for the outlet Hyl III reactor model in the same manner. To prove that flow interruption at the outlet Hyl III reactor is caused by fluidization, it is required to recalculate the minimum pressure drop for fluidization in the Hyl III reactor using the fluidization equation and then the result is compared to the actual pressure drop in the Hyl III reactor. Fluidization occurs if the calculated minimum pressure drop is lower than the actual pressure drop in the Hyl III reactor. Experimental work was carried out by using the Hyl III reactor model. To prove that flow interruptions at the outlet Hyl III reactor was caused by agglomerate formation, cold experiments were conducted using the Hyl III reactor model.

Flow interruption caused by agglomerate formation occurs when the mixture of sand and agglomerates cannot flow at the Hyl III model reactor.

1.5 Limitation of Study

This research is simply to identify the causes of flow interruption as well as to identify the ways to prevent flow interruption. In this research, the inference drawn regarding the causes of flow interruption due to arching was based on the calculation only without any experiment conducted. In the same manner, the conclusion drawn regarding the causes of flow interruption due to fluidization was based on calculation and an experiment on a model. Flow interruption because of agglomerate formation was observed with experiments. The experiment was conducted with various mixture of agglomerates and sand. These experiment did
not directly study agglomerate formation. However, it observed flow interruption after agglomerate formation occurs. Therefore, it was hoped this experiment would explain the effect of the composition of agglomerate and sand mixture. In addition, it was expected to explain the effect of gas pressure on the flow interruption caused by agglomerate formation.

This study basically concerned possible mechanisms of flow interruptions. The results of this study were be compared and confirmed with the actual process.

1.6 Expected Outcome

This research was expected to identify the causes of flow interruption and therefore, allow the prevention of this phenomenon. In addition, it was expected that the research would pave the way for further research regarding flow interruption.

1.7 Structure of The Thesis

Chapter 2 consists of a literature review on reactor Hyl III, arching, fluidization and agglomerate formation.

Chapter 3 describes design criteria for the Hyl III reactor model, construction of the Hyl III reactor model, the Rotary valve model and material used in this experiment.

Chapter 4 describes the result and discussion about flow interruption both in the Hyl III reactor and in the model caused by arching, fluidization and agglomerate formation. In addition, this chapter discusses how to prevent flow interruption.
CHAPTER TWO

THEORY

2.1 Reactor

The HYL III Reactor is a continuous moving bed reactor. The iron ore is charged at the top and flows countercurrently to the reducing gases in the upper part. The iron ore is reduced to DRI and in the lower part it is carburized and cooled.

The HYL III Reactor is a vessel constructed of carbon steel plate. The upper part is a semispherical dome, followed by a cylindrical section that connects with another cylindrical section of larger diameter than the previous one and finally conical section in the lower part (Figure 2.1).

Except for the lower conical section, the reactor is lined with insulating refractory castable and refractory brick.

The exterior of the reactor has a manhole at the upper dome. It also has an inlet and outlet reducing gas and cooling gas connections, as well as thermocouples and pressure tap connections.
Figure 2.1. The Hyl III reactor.
The iron ore, either as pellet or lump or mixture of both, is charged into the reactor through the upper part and flows downwards by gravity.

The cooled (less than 60°C) Direct Reduced Iron (DRI) is discharged through the rotary valve that is located at the bottom of the lower conical section.

The reactor is divided into three zones:

a) Reducing zone.

b) Isobaric zone.

c) Cooling zone.

2.1.1 Reduction zone.

The reduction zone is made up of three sections (Figure 2.2): the first is the upper dome that has semispherical form.

Its interior is lined with insulating refractory castable. There are four iron ores feeding pipes located at the top of this section that enter the reactor. These four inlet tubes extended down into the reactor and serve the purpose of evenly distributing the iron ore across the internal cross section of the reactor.

Externally, there is a manhole for maintenance and inspection, as well as the reduction gas outlet that has a pressure tap and a thermocouple.

The second section is a 5.0 m diameter cylindrical shape and 5.124 m in height. It is lined with insulating refractory castable, insulating fire brick and refractory brick. There are two thermocouples located at 180° in the lower part of this section.
Figure 2.2. The Reduction Zone.
The last section of this zone is 3.841 m high in which its internal diameter increases at an angle of 2° with the vertical until it meets the inlet gas level. In the lower part is the inlet gas plenum as a ring built with refractory brick (Figure 2.3). The plenum is hollow and has 90 rectangular nozzles of 0.055x0.0190 m evenly distributed around the interior of the plenum to distribute the reducing gas uniformly around the circumference of the reactor. The plenum has only one reducing gas inlet.

In this section there are two thermocouples at 180° placed above the reducing gas plenum, also the plenum has a pressure gauge and a thermocouple in the reducing gas inlet.

The hot reducing gas coming from the gas heater enters the reactor through the reducing plenum. The gas is distributed uniformly around the plenum. The gas enters the reactor through the rectangular nozzles that direct the gas to the inside and then it flows upwards countercurrently to the iron ore.

As the reducing gas flows upward, it reduces the iron ore that flows downwards. In this manner the reducing potential of the reducing gas is utilized to maximum. In the upper part, the gas is collected in the void space (dome) before it leaves the reactor.

The reduction zone is the part of the reactor where almost the whole reduction process takes place.
Figure 2.3. The Reduction Plenum.
2.1.2 Isobaric Zone

It is a practically cylindrical section lined in the interior with insulating refractory castable, insulating fire brick and refractory brick. It increases its internal diameter towards the bottom forming an angle of $1^\circ$ with the vertical (Figure 2.4). This section is 5.486 m high and 5.664 m in diameter at the upper part; it has two thermocouples at $180^\circ$ in the lower part and two thermal expansion joints in the refractory.

This section is the interphase that separates the reducing gas stream and the cooling gas stream.

The outlet plenum for the cooling gas that is of similar construction to the reducing gas plenum is located in the lower part of this section.

The plenum has two thermocouples at $180^\circ$ and one pressure gauge. The cooling gas outlet also has one thermocouple.
Figure 2.4. The Isobaric Zone.
2.1.3 Cooling Zone

This conical section is located at the lower part of the reactor immediately below the Isobaric Zone, and has one more conical section incorporated to it (Figure 2.5). This section is where the carburizing and cooling of DRI takes place. It has no refractory on the inside, and presents a smooth metallic face to the solids to allow them to flow freely.

As a safety precaution in the upper part of this zone where the temperatures are higher, there is a water cooled exterior jacket to prevent any localized heating of the metal in this part.

The cooling gas before it leaves the reactor is collected in a plenum in the shape of a ring around the conical section to ensure a uniform distribution of the gas, and from here, the gas goes to outlet duct (Figure 2.6).

The plenum is formed by 36 metallic plates each one supported by two pipes that cross the reactor wall. The metallic plates serve to support the internal refractory of the isobaric zone.

Similarly to the inlet plenum of the reducing gas (Figure 2.3), the cooling gas enters the reactor through a metallic plenum (Figure 2.6), located slightly below the middle part of the conical section. The plenum forms a hollow ring around the cone and has four inlet gas connections located at 90° one from the other.

There is an expansion joint and a sliding cut-off valve (guillotine) in lower part of the cooling zone. There are two clusterbreakers located above this valve.
that contain hydraulic cylinders attached to steel bars and are use for breaking cluster formation that could be present. When operated, these devices are pushed and pulled repeatedly to break the cluster. However, they are used only in emergency cases.

At the bottom of the cooling zone is the rotary valve that controls the rate of discharge of DRI.

Figure 2.5. The Cooling Zone.
Figure 2.6. The Cooling Gas Outlet Plenum.
One of the common problems appears in discharging the direct reduced iron (DRI) is flow interruption. To observe the cause of flow interruptions in the actual Hyl III reactor is hardly possible because it is very costly and also disturbs the production. To simplify the observation of the problem, a model was made which is one tenth of the actual size and similar to the actual reactor. The flow interruption at the outlet of an Hyl III reactor is possibly caused by an arching, fluidization or agglomerates formations are all considered.

2.2 Arching in Mass Flow Bins

This can be due to the interlocking of few particles that are large with respect to the outlet or due to the formation of a cohesive arch [Carson and Johanson, 1985].

To overcome the first type of arching, the outlet size of a circular bin should be at least five to six times the largest particle size if a circular outlet is used, or three to four times (i.e. minimum width) if rectangular outlet is used.

The procedure used to assure that a cohesive arch will not form is somewhat more complicated. First, it is necessary to compute the flow factor of the hopper, defined as:

\[
\text{Flow factor (ff) = (major consolidating pressure at a point in the hopper) / (minimum stress in the abutment of an arch at this point).}
\]
Then, as shown in Figure 2.7, a line is drawn starting at the origin having an inverse slope of flow factor.

![Diagram showing flow factor (ff) and Flow Function (FF) relationship]

**Figure 2.7. Flow or No Flow Condition for Mass - Flow Design [source: Carson and Johanson, 1985].**

The ability of a bulk material to flow depends on the strength developed by the material due to consolidation and whether. As a result of this strength, the material is able to form a stable arch or pipe. The unconfined yield pressure is a measure of the material's strength at a free surface and is a function of the major consolidating pressure. This is referred to as the "Flow Function (FF)" of the material.
Flow Function (FF) is representing the strength of the material.

There are now several possibilities:

1. The flow function lies entirely below the flow factor.

   This indicates a material that is easy flowing. Consideration of particle interlocking and flow rate dictate the minimum outlet size, not cohesive arches [Carson and Johanson, 1985].

2. The flow function intersects the flow factor.

   Naming the value at the point of intersection as $\sigma_1$, the minimum outlet size $B$ required to prevent cohesive arches is then:

   \[
   B = \frac{\sigma_1 H(\alpha)}{\gamma} \quad (2.1)
   \]

   where $H(\alpha)$ : function as shown in Figure 2.8.

   $\gamma$ : bulk density of solid

   Since the flow factors for conical and wedge hoppers are essentially the same for the same value of effective angle of friction ($\delta$), the fact that $H(\alpha)$ for circular outlet is approximately double that of a rectangular one in Figure 2.8, suggests that the minimum outlet size of conical hopper is approximately twice the minimum width of a wedge hopper.

3. The flow function lies entirely above the flow factor.

   This indicates that gravity alone is not sufficient to overcome cohesive arching. Flow aids such as vibrators or air blasters must be used, or the materials flow function must be changed by, for example lowering its
moisture content or temperature or adding a free flow additive [Carson and Johanson, 1985].

\[ H(\alpha) \]

---

![Diagram](image)

**Figure 2.8. Function H (\(\alpha\)) [Arnold, 1992].**

In the design of flow bins it is assumed that principle potential obstruction to flow is due to the formation of stable cohesive arches. There are many shapes of arches that could form, but for illustration purposes, a simple analysis upon which the Jenike Theory [Arnold, 1992] was originally given as follows:

Consider an arch of uniform thickness (T) as shown in Figure 2.9.
Neglecting any dynamic affects the equilibrium of the arch may be analyzed.

For a circular opening diameter (B):

\[
\sigma_1 \pi B T \sin \beta \cos \beta = W \tag{2.2}
\]

Where:

\[
W = \gamma . T. \pi. (B^2/4) \tag{2.3}
\]

\[
\gamma = \rho \cdot g
\]

\[
\rho : \text{solid density}
\]

\[
g : \text{acceleration due gravity}
\]
Choosing, arbitrarily, angle $\beta$ of the arch to be $45^\circ$ so as to maximize the vertical component of the supporting forces, then the major principal stress $\sigma_1$ in the obstructing arch is minimized.

Thus the equation for equilibrium can be written in the generalized form.

$$\sigma_1 = \gamma \cdot (B/2) \quad (2.4)$$

$$\sigma_1 = (1/2) \rho \cdot g \cdot B \quad (2.5)$$

Jenike and Leser [Arnold, 1992] refined the arch analysis by considering the variation in the thickness of the arch. They developed the expression:

$$\overline{\sigma_1} = (\gamma \cdot B) / (H(\alpha)) \quad (2.6)$$

Function $H(\alpha)$ is given in graphical form in Figure 2.8.

The ratio $\sigma_1/\overline{\sigma_1}$ is termed the flow factor ($ff$) for the hopper. That is

$$ff = \sigma_1 / \overline{\sigma_1} \quad (2.7)$$

The flow factor is used to indicate the “flowability” of a channel.

To prove whether the flow interruption is caused by an arching or not is done by using an arching equation to calculate a minimum outlet diameter of Hyl III reactor cause an arching. The result of the calculation is then compared to the actual condition. The same method is also applied to evaluate the flow interruption in the model.
2.3 Fluidization Phenomena

Fluidization is the operation by which fine solids are transformed into a fluidlike state through contact with gas or liquid [Kunii and Levenspiel, 1977].

When some fluid is passed upward through a bed of fine particles at low flow rate, the fluid merely percolates through the void spaces between stationary particles. This condition defines a fixed bed. With an increase in flow rate, particles move apart and a few may be seen to vibrate and move about in restricted region. This defines an expanded bed. At a still higher velocity a point is reached when the particles are all just suspended in the upward flowing gas or liquid. At this point the frictional force between a particle and fluid counterbalances the weight of the particle, the vertical component of the compressive force between adjacent particles disappears, and the pressure drop through any section of the bed about equals the weight of fluid and particles in that section. The bed in this condition is considered to be just fluidized or a bed at minimum fluidization [Kunii and Levenspiel, 1977]. Under condition at fluidization, the pressure drop ($\Delta P$) relatively remains constant with the increase of $\Delta U$, which is the relative velocity between gas and solid. For condition of stick slip flow the pressure drop ($\Delta P$) increases in line with the relative velocity between gas and solid ($\Delta U$).
2.4 Minimum Fluidization Velocity

The superficial velocity at minimum fluidization condition \( (U_{mf}) \), is found by Wen and Yu [Kunii and Levenspiel, 1977], to be:

\[
U_{mf} = \frac{dp^2 (\rho_s - \rho_g) g}{(1650 \mu)} \quad (2.8)
\]

for small particles and

\[
U_{mf}^2 = \frac{dp (\rho_s - \rho_g) g}{(24.5 \rho g)} \quad (2.9)
\]

for large particles

2.5 Flow of High Bulk Density Mixtures

Theoretically, there are two forms of high bulk density flow, stick slip flow and aerated flow [Kunii and Levenspiel, 1977]. Stick slip flow is a jerky movement of compacted grains with velocity of particles at the walls slightly lower than that in the rodlike core. It occurs primarily with large particles and the solid flow is always downward.

In aerated flow solids are fluidized or suspended by the gas and have a high mobility compared to stick-slip flow. For aerated solids, the pressure drop is named by two terms: the static pressure and the frictional loss term. Aerated flow is usually limited to fine particles and can be used to transport solids in any direction as well as in U-tubes.

To illustrate the relation between these two types of flow and the corresponding pressure variations, consider flow in a vertical tube. Four cases may be distinguished as shown in Figure 2.10. In Figure 2.10 a, since solids are moving upward they must be fluidized. Therefore, the relative velocity between gas and
solid $\Delta U$ must be sufficient to keep the solids suspended. The pressure drop in the pipe will slightly exceed the static head of suspended solids. The gas moves up faster than the solids. Figures 2.10 b and 2.10 c may also represent aerated flow as long as $\Delta U$ satisfies condition of case (a). If the relative velocity is not high enough to fluidize the solids, stick slip flow will be the result. Figure 2.10 d can only represent stick - slip flow. Solids move from a high to low pressure region. The gas descends faster than the solid.

\[ \text{Aerated flow if } \Delta u > u_{mf} \]

\[ \text{Stick-slip flow} \]

Figure 2.10 Types of flow, relative velocities, and pressure gradients in flowing gas-solid mixtures with high bulk density of solid.
These sketches show that the flow of Figure 2.10 d is used when solids descend from vessel at high pressure to one at lower pressure, whereas the flows of Figure 2.10 b and 2.10.c are used when solids descend to a vessel at higher pressure. Figure 2.10 c also shows that gas can be made to flow from low to high pressure, or with proper design gas can remain essentially stationary and act as a seal.

To differentiate between stick slip flow and aerated flow at system shown in Figure 2.10 b to 2.10 c use the linear velocity of gas relative to solids ($\Delta U$) and the minimum fluidization velocity ($U_{mf}$). Theoretically, stick slip flow happens if $\Delta U$ is lower than $U_{mf}$. And if $\Delta U$ is higher than $U_{mf}$, it means aerated flow.

### 2.5.1 Pressure drop in Stick - Slip Flow

In stick slip flow, the movement of gas relative to the solids (not relative to walls) determines the pressure gradient. This is because the frictional resistance between gas and solid overshadows that between gas and walls. Thus, the pressure difference between two points in pipe is given by the packed bed expression of the Ergun equation [Kunii and Levenspiel, 1977].

\[
\Delta P = \frac{150 (1-\varepsilon)^2}{\varepsilon^3} \frac{|\Delta u|}{(\phi s \, dp)} + \frac{1.75 (1-\varepsilon)}{\varepsilon^3} \frac{\rho g (\Delta u)^2}{(\phi s \, dp)} \quad (2.10)
\]

where $\Delta u$ is linear velocity of gas relative to solids

\[
\Delta u = (U_0/\varepsilon) - U_s \quad (2.11)
\]
Equation (2.11) is used for Figure 2.10 d. As it is shown figure 2.10. b the solids move downward while the gas moves up, so the relative linear velocity between the gas and the solids ($\Delta U$) is the sum of gas velocity $(U_0/\varepsilon)$ and the solids velocity $(U_s)$ or expressed by the equation

$$\Delta U = \frac{U_0}{\varepsilon} + U_s \quad (2.11a)$$

If the pressure drop is too high, then $\Delta u$ increases sufficiently so that solids will fluidize to give aerated flow. The pressure drop for fluidization is then given by:

$$\Delta P = \rho_s * (1-\varepsilon) * g * \Delta h \quad (2.11b)$$

### 2.5.2. Pressure Drop in Aerated Flow

For aerated solids the pressure drop is named by two terms, the static pressure and the frictional loss term. Thus, the Bernoulli equation between the lower point 1 and upper point 2 in a pipe inclined at any angle $\theta$ to the horizontal becomes:

$$P_1 - P_2 = \rho_s * g * (h_2-h_1) \pm |\Delta P \text{ friction}| \quad (2.12)$$

$$P_1 - P_2 = \rho_s * g * (\Delta L \sin \theta) \pm |\Delta P \text{ friction}| \quad (2.13)$$

and

$$|\Delta P \text{ friction}| = 32\mu U_s L / dt^2 \quad (2.14)$$

Where "+" refers to upward gas flow, "-" refer to downward gas flow, and the friction term is always positive.

For a dense bubbling fluidized bed the static head term in equation (2.12) is much greater than the friction term, or

$$\rho_s * g * (h_2-h_1) \gg |\Delta P \text{ friction}|$$
hence equation (2.12) reduces to:

\[ P_1 - P_2 = \rho_s \times g \times (h_2 - h_1) \] (2.15)

2.6 Properties of Granular Materials

2.6.1 The Bed Porosity

The definition of the porosity as cited by Dullien [Fayed and Otten, 1994] is

\[ \varepsilon = \frac{\text{volume of void in packing}}{\text{bulk volume of packing}} \] (2.16)

2.6.2 Particle Size

Particle size can be measured in a number of ways. For large particles, direct measurement is conducted with micrometer. For very small particles in the microscopic range indirect method relying on settling rate, Brownian Movement is used [Kunii and Levenspiel, 1977].

For non-spherical particles the diameter may be defined as:

\[ d_p: \text{diameter of sphere having the volume of the particle} \]

Solids with a distribution of sizes have a mean particle size that can be calculated with equation (2.18) [Kunii and Levenspiel, 1977].

\[ d_p = \frac{1}{\sum (\Delta d_p) i/d_{pi}} \] (2.17)

\[ d_p = \frac{1}{\sum (x/dp)i} \] (2.18)
2.6.3 Particle Shape

It is known that particle shape influences such properties as packing and interaction with fluids, although not much work has been carried out on these relationships.

For non-spherical particles a variety of measures of sphericity exist, the simplest of which is sphericity $\phi_s$, defined as

$$\phi_s = \frac{\text{surface area of a sphere}}{\text{surface area of particle}}$$

both of same volumes. With this definition, $\phi_s = 1$ for sphere, and $0 < \phi_s < 1$ for all other particles shapes.

2.6.4 Particle Density

Particle density can be calculated with equation (2.20).

$$\rho_s = \frac{M_p}{V_p}$$

2.6.5 Bulk Density

The mathematical expression of bulk density is:

$$\gamma = \rho_s(1-\varepsilon)$$

2.6.6 Flow Function

This material property is defined as follows:

Flow function: the relationship between cohesive strength of bulk solid and consolidating pressure.
2.7 Similarity Criteria

For economical and safety reason, usually a physical modeling of the system is required. In similarity, experimental work becomes a key factor in studying the system. A knowledge of similarity criteria is essential as noted by Zlokarnik [Subagyo, 1997] to achieve greater confidence in studying physical phenomena.

In order to study the operation of an existing process in the industrial scale on a suitable model, researchers are frequently faced with problem of defining the minimum possible factors which will adequately represent the geometry and operating condition of system. The model need not reproduce the entire system but only the characteristic aspect under the study. The concept of general similarity originated in the geometric similarity of form such as the similar triangles. The necessary condition for one system to be similar to another are expressed by similarity criteria. Knowledge of the applicable similarity criteria provides an insight into the system and can be used to minimize the experimentation necessary to relate the important process variables.

In general, the configuration (geometric and dynamic) similarity of system can be defined by ratio of magnitudes within the system, which do not depend on the unit of measurement, to quote Szekely [Purnomo, 1995].

For instance, a fixed ratio of outlet diameter of a cylindrical reactor and inlet diameter of cylinder can be considered as the necessary condition for geometric similarity. Geometric similarity can be see in Figure 2.11.
reactor outlet diameter
\[
\frac{\text{model outlet diameter}}{= C} \quad (2.22)
\]

Reactor inlet diameter
\[
\frac{\text{model inlet diameter}}{= C} \quad (2.23)
\]

Figure 2.11. Geometric Similarity for Model Reactor.

2.7. EXPERIMENTAL DESIGN

2.7.1. Fractional Factorial Design.

Two-level, full factorial designs are very powerful because they provide information about all main effects and two-factor interactions. However, the
The designs that are used most frequently for screening experiments are two level fractional factorial designs, a fraction (subset) of a full factorial. The number of combinations of factor settings included in the design is the sample size for an unreplicated design. For example, a full factorial design with three factors would have a sample size of $2^3 = 8$ combinations of high and low settings (Table 2.1). Similarly, full factorials for 4, 5, 6 factors have sample sizes of 16, 32, and 64. However, if a design with four factors has a sample size of 8, then it would be one-half fraction of the full factorial. Similarly a six factor experiment with a sample size of six i.e. of 16 would be one-fourth fractional factorials and so on.

Table 2.1. A $2^3$ Factorial Design

<table>
<thead>
<tr>
<th>Run</th>
<th>A</th>
<th>B</th>
<th>C</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>2</td>
<td>+</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>3</td>
<td>-</td>
<td>+</td>
<td>-</td>
</tr>
<tr>
<td>4</td>
<td>+</td>
<td>+</td>
<td>-</td>
</tr>
<tr>
<td>5</td>
<td>-</td>
<td>-</td>
<td>+</td>
</tr>
<tr>
<td>6</td>
<td>+</td>
<td>-</td>
<td>+</td>
</tr>
<tr>
<td>7</td>
<td>-</td>
<td>+</td>
<td>+</td>
</tr>
<tr>
<td>8</td>
<td>+</td>
<td>+</td>
<td>+</td>
</tr>
</tbody>
</table>
2.7.2. The Pareto Principle

A statistical analysis provides estimation of how strongly experimental factors affect process performance. This estimation reveals which factors are most important and how changing these settings affects process performance. The results analysis is necessary to establish the magnitude of the error involved so that the effect of parameters can be correctly assessed. The overall or pooled standard deviation needed for the evaluation of experimental error is found from equation (2.24).

\[
S_{\text{overall}} = \left[ \frac{S_1^2 + S_2^2 + \ldots + S_8^2}{8} \right]^{1/2} \tag{2.24}
\]

where: \( S_1^2, S_2^2, \ldots, S_8^2 \) are the variances of each experiment \((S_1^2)\) from run 1 to run 8, calculated from equation (2.25) by standard methods [Spigel,1972].

\[
\text{Variance} = S^2 = \frac{\sum_{i=1}^{N} (X_i - \bar{X})^2}{N} \tag{2.25}
\]

Using the 3 S or 99 % confidence interval to know most important effect and other important effects or the noise.

The Pareto chart in Figure 2.5, graphically displays the magnitudes of the effects sorted from the largest to the smallest. Using the 3 S it is clearly shown that the most important effects are located at the upper 3S line and other unimportant effect(s) or the noise is located at the lower 3S line.
To prove whether the flow interruption at the outlet Hyl III reactor is caused by fluidization or not, pressure drop was calculated from the equation of minimum pressure drop for fluidization and then compared with the actual pressure drop in the reactor. The same method was also applied to evaluate the flow interruptions caused by fluidization in the model.
2.8 Agglomerate Formation

It is well known that productivity of Direct Reduction reactor can be effectively increased by increasing the temperature of the incoming reducing gas. It has been shown that increasing the reduction gas temperature by 50 °C increases productivity by 40-50 % (Narita, et al, 1980). Additionally, raising the temperature of the reducing gas is also advantageous for prevention of reoxidation of the products.

Unfortunately, this method of increasing the temperature of incoming reducing gas is restricted due to the formation of the so called “cluster”, a phenomenon in the high temperature region of the reduction zone in the reactor.

Understanding the influence of agglomerates to the flow interruptions was done in the model by varying the mixture of agglomerates and sand, either with or without gas flow. The experimental results show which composition of the agglomerates and sand mixture stops the flow (with or without gas flow).
CHAPTER THREE

EXPERIMENTAL EQUIPMENT

AND PROCEDURE

3. 1. The Hyl III Reactor Model

3.1.1. Design Criteria for the Hyl III Reactor Model

3.1.1.1 Similarity Material

Iron ore for Hyl III reactor is categorized as nonspherical material. Its sphericity, measured according to equation (2.19), is 0.9. To fulfill similarity criteria of the packing material, the model material ought to have the same sphericity. For this purpose a packing material called "feroxide" with specification of solid density 4615.4 Kg/m$^3$, bulk density of 3000 kg/m$^3$ and sphericity of 0.9 was chosen initially. For the conditions of the experiment, the minimum fluidization of "feroxide" is 622.5 m$^3$/h (see Appendix D1). This flow rate is higher than supply capacity from Wire Rod Plant Utility of 500 m$^3$/h (see Section 3.2.2). Fluidization cannot occur if this material is used.

Therefore, "feroxide" material was substituted with sand with sphericity of 0.8 and bulk density of 1250 kg/m$^3$. Those figures were lower than those of the "feroxide", as required. Furthermore, its sphericity is also lower than the actual
sphericity of iron ore. So, As the difference of both sphericities is only 11.1% this was considered acceptable to the design similarity criteria.

The actual granular metric analyses after screening were:

<table>
<thead>
<tr>
<th>Size distribution</th>
<th>% weight</th>
</tr>
</thead>
<tbody>
<tr>
<td>16 - 13 mm</td>
<td>45 %</td>
</tr>
<tr>
<td>13 - 10 mm</td>
<td>45 %</td>
</tr>
<tr>
<td>10 - 6 mm</td>
<td>10 %</td>
</tr>
</tbody>
</table>

bulk density : 1700 kg/m³

It is difficult to fulfill the criteria similarity using the same size distribution of 1:10 but having the same weight percentage. Therefore, in this experiment two size distributions were chosen where one has the same weight percentage and the other has different weight percentage.

Size distribution 1:
- 0.25 - 0.5 mm = 33.33 %
- 0.5 - 1 mm = 33.33 %
- 1 - 2 mm = 33.33 %

Size distribution 2:
- 0.25 - 0.5 mm = 45 %
- 0.5 - 1 mm = 45 %
- 1 - 2 mm = 10 %

3.1.1.2 Similarity of The Hyl III Reactor Model

The Hyl III Reactor model design was based on dynamic and geometric similarity.
For practical reasons and restriction of loading and lifting services available, the reactor model size was determined 10% of the actual Hyl III reactor size (see Section 2.6).

\[
\frac{\text{reactor outlet diameter}}{\text{model outlet diameter}} = 10
\]
\[
\frac{\text{reactor inlet diameter}}{\text{model inlet diameter}} = 10
\]

The bottom of the Hyl III reactor model is conical with an angle of 16° from the vertical line. This structure is precisely the same as the cooling zone in the Hyl III reactor.

### 3.1.1.3 Similarity of the Processes in The Hyl III Reactor Model.

Minimum fluidization in the Hyl III Reactor, from equation (2.9), is

\[
\text{Umf}^2 = (dp)(\rho_s - \rho_g)(g)/(24.5*\rho_g)
\]
\[
\text{Umf}^2 = (0.01)(3230-1.186)(9.8)/(24.5*1.186)
\]
\[
= 3.299 \text{ m/s}
\]

where:

- \(dp\) = mean particle diameter = 0.01 m
- \(\rho_s\) = solid density = 3230 kg/m³
- \(\rho_g\) = gas density = 1.186 kg/m³
- \(g\) = gravity = 9.8 m/s²
The flow rate of the cooling gas at inlet is 50615 m³/h and cross section area is 9.24 m². The velocity of actual cooling gas is therefore 1.521 m/s. Compared to the above minimum fluidization velocity calculated from equation 2.9 (3.299 m/s), this figures is only 46.12%.

As noted earlier (Section 3.1.1), when "feroxide" was initially used, fluidization did not occur because available air capacity was not enough for fluidization, so that the packing material ("feroxide") was substituted with sand. The sand used had a bulk density of 1250 kg/m³ and its voidage from measurement (Appendix H) was 0.401 and mean particle diameter was 6.43 x 10⁻⁴ m. Using the sand as the packing, minimum velocity for fluidization (Umf) is 0.269 m/s (see Appendix D2).

The actual experimental velocity is:

\[ U_{exp} = 0.4612 \times 0.269 \]

\[ = 0.124 \text{ m/s}. \]

Therefore, the actual velocity is 0.124 m/s. Compared to the minimum fluidization velocity (Umf) for sand of 0.264 m/s (see Appendix D2), this figure is 46%, and this is the same as the proportion of the velocity of actual cooling gas with the minimum fluidization velocity in the Hyl III reactor.

The velocities of gas used in this study were 0.13 m/s and 0.26 m/s which are greater than 0.124 m/s as required.
3.1.1.4 Similarity of Solid Flow rate.

The actual flow rate Hyl III reactor is 4.335 kg/s and with diameter outlet of the Hyl III reactor is 0.57 m, the cross section area is 0.255 m$^2$. So, the bulk density of solid is 1700 kg/m$^3$.

The actual solid velocity in the Hyl III reactor uses the following equation:

$$ v = \frac{Q}{(A \times \gamma)} $$

where $Q$ : flow rate = (kg/s)

$A$ : cross section area (m$^2$)

$\gamma$ : bulk density of solid (kg/m$^3$)

$$ v = \frac{4.335}{(0.255 \times 1700)} = 0.01 \text{ m/s} $$

With velocity of solid 0.01 m/s the flow rate of the Hyl III reactor model can be calculated using equation $Q = A \gamma v$.

If the diameter outlet of the Hyl III reactor model is 0.343 m, the cross section area is 0.0924 m$^2$ and the bulk density of solid is 1250 kg/m$^3$, so, the final result calculation flow rate is 1.155 kg/s.

The velocities of solid used in this study were 0.01 m/s and 0.02 m/s.

3.1.2 Construction of The Hyl III Reactor Model

The model reactor used for the study consists of three sections, one cylindrical and two conical. Each section is made up of 3.5 mm thick mild steel. Flanges are provided at two upper locations so that sections can be removed as required. Each section and others are joined with bolts at flanges.
Section 1 (upper) is cylindrical and closed by flanges. Section 2 (center) is conical with an angle of $16^\circ$ from vertical line. This section contains one outlet hole for gas pressure outlet pipe and a pressure indicator outlet. Section 3 (bottom) is conical too, and also with an angle of $16^\circ$ from the vertical line. This section contains four inlet pipe holes for gas pressure inlet and a pressure indicator inlet. After this section, a gate valve is provided for starting or stopping flow of the material.

The Hyl III reactor model was designed by the candidate, drawn by Moh.Jamil and manufactured in the PT. Krakatau Steel workshop. The installation of the experimental apparatus was done by Field Maintenance of PT Krakatau Steel.

A line diagram of the experimental apparatus is shown in Figure 3.1, and a photograph of the Hyl III reactor model set-up is exhibited in Figure 3.2. The dimensions are shown in Figure 3.3.
Figure 3.1. Line diagram of the experimental apparatus.
Figure 3.2. A photograph of the Hyl III reactor model set-up.
Figure 3.3. The Hyl III Reactor Model.
3.1.3 The Rotary Valve Model

The rotary Valve is the reactor discharge system. The system's function is to assure a continuous material flow through the reactor. The rotary valve model was provided for discharging the solids from the model reactor. The rotary valve model was designed by the candidate, and drawn by Moh. Jamil. Finally, it was manufactured in the PT Krakatau Steel workshop. It was made from high tensile steel and all parts were machined to a 400 grade finish, and driven by an AC motor with a power of 1.1 KW and output revolutions of 64 rpm. The dimensions of the rotary valve model can be seen in Figure 3.4. The Main parts of the rotary valve are the shaft with four impellers and bushing with holes for the inlet of the solid material and for the outlet of the solid material.

Figure 3.4. The rotary Valve Model.
3.1.4. Pressure Difference Measurement Device

Pressure difference between inlet and outlet air in the Hyl III model reactor was measured with a plastic tube manometer filled with red colored water.

3.2. Materials

3.2.1 Granular Material

The packing material used in this experiment was sand. Prior to the experiment, the sand was dried under the sun and then screened using 0.25 mm, 0.5 mm, 1 mm and 2 mm aperture screens.

The specification of the sand used as the packing is as follows:

-Size distribution 1: 0.25 - 0.5 mm = 33.33 %
  
  0.5 - 1 mm = 33.33 %
  
  1 - 2 mm = 33.33 %

-Size distribution 2: 0.25 - 0.5 mm = 45 %

  0.5 - 1 mm = 45 %
  
  1 - 2 mm = 10 %

-Bulk density = 1250 kg/m³

3.2.2. Air Supply

The air supply used was from the utility workshop of the Wire Rod Plant, rated at 500 m³/h and pressure at 4 bar. This air was used to simulate the cooling gas of the Hyl III reactor in the Hyl III reactor model.
3.2.3 Agglomerates

The actual size agglomerates when sticking occurs at the Hyl III reactor varies between 100 mm to 200 mm with non uniform shape. The size of agglomerates depends on the length of sticking time. For this experiment, 200 mm size agglomerates were chosen and then after scaling down by one tenth, the size of agglomerates becomes 20 mm.

Sand agglomerates was made up of sand, cement and water with ratio of 2:1:1. The size of agglomerates was approximately 20 mm. Finally, they were dried under the sun.

3.3. Procedure

3.3.1 Preparation

Measuring Sand, Sponge iron and “Feroxide” density and bed porosity

1. Prepare Sand with size distribution 1.
2. Weigh sample of sand.
3. Fill water into the glass, measure initial volume, put a sample sand into the glass and measure volume of glass. The difference of the volume of glass is the volume of the sample.
4. Calculate particle density of sand using equation (2.20).
5. Repeat procedure 2 until 4 for sand with size distribution 2, sponge iron and “feroxide”.
6. Prepare 5 liters of sample for sand with size distribution 1.
Experimental Equipment and Procedure


8. Repeat procedure 6 and 7 for other samples.

9. Calculate the porosity of packed bed for sand with size distribution by equation (2.21).

3.3.2 Experiment

3.3.2.1. Experiments of arching phenomena

Basically the experimental procedure consisted of the following steps:

a) Fill the Reactor with sand of given size distribution, after mixing.

b) Open the solid gate valve, start the rotary valve with a predetermined flow rate.

c) Open the air supply gate valve with a pre-determined flow rate.

d) Record the pressure difference between the inlet and outlet flow of air.

The above procedure was followed for each of the eight different conditions used according to a $2^3$ factorial design (Section 2.7). The details of these runs are given in Table 3.1.
Table 3.1 Details of The Runs

<table>
<thead>
<tr>
<th>Run</th>
<th>Sand Size Distribution, %</th>
<th>Velocity of Solids, m/s</th>
<th>Velocity of gas, m/s</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.25-0.5 mm</td>
<td>0.5-1 mm</td>
<td>1-2 mm</td>
</tr>
<tr>
<td>1</td>
<td>33.33</td>
<td>33.33</td>
<td>33.33</td>
</tr>
<tr>
<td>2</td>
<td>33.33</td>
<td>33.33</td>
<td>33.33</td>
</tr>
<tr>
<td>3</td>
<td>33.33</td>
<td>33.33</td>
<td>33.33</td>
</tr>
<tr>
<td>4</td>
<td>33.33</td>
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<td>45</td>
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<td>6</td>
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</tr>
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<td>7</td>
<td>45</td>
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<td>10</td>
</tr>
<tr>
<td>8</td>
<td>45</td>
<td>45</td>
<td>10</td>
</tr>
</tbody>
</table>

3.3.2.2. Experiments on flow interruption using agglomerates

Basically the experimental procedure consisted of the following steps:

a) Mix a certain amount of sand and sand agglomerate according to predetermined composition (see Table 3.2) and fill them in the Reactor.

b) Open the solids gate valve and start the rotary valve with predetermined flow rate.

c) Open the air supply gate valve with a predetermined flow rate, if needed.

d) Observe the appearance of flow interruptions.

The above procedure was followed for each of the ten different conditions. The details of these runs are given in Table 3.2.
Table 3.2. Details of the Runs with agglomerates.

<table>
<thead>
<tr>
<th>NO</th>
<th>MIX A/S</th>
<th>AIR</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>1/5</td>
<td>NO</td>
</tr>
<tr>
<td>2</td>
<td>1/4</td>
<td>NO</td>
</tr>
<tr>
<td>3</td>
<td>7/24</td>
<td>NO</td>
</tr>
<tr>
<td>4</td>
<td>1/3</td>
<td>NO</td>
</tr>
<tr>
<td>5</td>
<td>1/2</td>
<td>NO</td>
</tr>
<tr>
<td>6</td>
<td>1/5</td>
<td>YES</td>
</tr>
<tr>
<td>7</td>
<td>1/4</td>
<td>YES</td>
</tr>
<tr>
<td>8</td>
<td>7/24</td>
<td>YES</td>
</tr>
<tr>
<td>9</td>
<td>1/3</td>
<td>YES</td>
</tr>
<tr>
<td>10</td>
<td>1/2</td>
<td>YES</td>
</tr>
</tbody>
</table>

3.4 Reproducibility

Reproducibility is a key factor to verify the validity of an experiment. The reproducibility of the measurement was calculated according to equation (3.1) proposed by Davies and Goldsmith [1977].

\[
R = \frac{\sum_{i=1}^{nd} \left| \frac{d_i - d}{d} \right|}{nd} \times 100\% \quad (3.1)
\]
Chapter Four

RESULTS AND DISCUSSIONS

This chapter will first discuss arching in the Hyl III reactor and in the Hyl III model reactor and then this will be followed by a discussion of fluidization in the same two reactors. Agglomerates formation in the Hyl III reactor will be discussed after arching and fluidization discussion.

4.1 Flow Interruptions

4.1.1 Arching in the Hyl III Reactor

As noted in Section 2.1, arching was caused by the interlocking of a few particles which are large with respect to the outlet. It can be due to the formation of a cohesive arch [Carson and Johanson, 1985].

The critical outlet diameter from calculation with effective angle of friction ($\delta$) = $55^\circ$, kinematic angle of friction between the solid and the bin wall ($\phi$) = $30^\circ$ and hopper half angle ($\alpha$) = $16^\circ$ in Appendix A, is 0.3 m. This is lower than the actual outlet diameter of the Hyl III reactor of 0.57 m. Therefore, it is safe to conclude that arching cannot occur, because of the dimensions of the particles used being smaller that they cannot interlock. On
this basis, the several times observed clogging in the outlet of the Hyl III reactor must have been due to some other cause and not arching.

In normal temperature the temperature in the reduction zone of the reactor reaches 900°C. At this or higher temperature, the iron oxide materials that are being reduced some times stick together causing flow problem. Sticking occurs due to the increasing volume of reducing gas so that large agglomerates are formed with size between 100 mm to 200 mm. The various size of agglomerates are formed depends on the length of sticking time. With outlet diameter reactor Hyl III of 0.57 m and the size of formed agglomerates about 100 mm and with the number of agglomerates of minimum six, arching will occur [Carson and Johanson, 1985].

4.1.2 Arching in the Hyl III Model Reactor

In the Hyl III model Reactor, the critical outlet diameter from calculation with effective angle of friction (\( \delta \)) = 60°, kinematic angle of friction between the solid and the bin wall (\( \phi \)) = 28° and hopper half angle (\( \alpha \)) = 16°, given in Appendix B, is 0.03 m. Again, this is lower than the actual used Hyl III model reactor outlet diameter (0.057 m) and again arching cannot occur.

Experimental runs with sand of various sizes did not result in any flow interruptions, either with or without the countercurrent flow of air, provided the air flow was below the minimum fluidization velocity. The reason that arching can not occur is because the diameter of particles involved (0.25 - 2 mm) is very small compared with the
E:q)enmental Results and Discussions

outlet diameter of the model reactor (57 mm). The outlet diameter of the model reactor is
228 - 28.5 times the particle diameter. Arching occurs when the maximum outlet diameter
size is at least five to six times the largest particle size if circular outlet is used [Carson and
Johanson, 1985].

For confirmation purposes of the above conclusion, the packing material was
substituted from sand to sponge iron pellets with diameter 10 - 18 mm. Arching could be
readily induced because the outlet diameter is only 5.7 - 3.16 times the particle diameter.
This is according to Carson and Johanson [1985], can readily lead to interlocking of a few
particles that are large with respect to the outlet.

4.1.3. Fluidization in the Hyl III Reactor.

Fluidization is the operation by which fine solids are transformed into a fluid-like
state through contact with fluids.

In the Hyl III Reactor fluidization of the sponge iron, if it occurs, would occur
through contact with the Cooling gas. Cooling gas inlet capacity is 50615 m$^3$/h at a
pressure of 5.27 kg/cm$^2$ and cooling gas outlet capacity is 47520 m$^3$/h at a pressure of
4.17 kg/cm$^2$.

The calculated minimum pressure drop for fluidization (Appendix C) is 133280 Pa.
Actual pressure drop in the Hyl III Reactor is 0.1 kg/cm² or 9800 Pa. This is an order of magnitude different and, accordingly, fluidization under these conditions in the Hyl III Reactor cannot occur.

If gas distribution is non-uniform it will lead to poorly fluidized beds. Thus the large pressure fluctuation result in a slugging bed, whereas an absence of the characteristic sharp change in slope minimum fluidization and the abnormally low pressure drop lead to incomplete contacting with particles only partly fluidized (channelling).

Because of slugging or channeling, the possibility of clogging as result of formed agglomerates becomes higher. This is because the pressure fluctuation or channelling causes the agglomerates became easy to gather resulting in the more possibility of clogging.

4.1.4. Fluidization in the Hyl III Reactor Model

Experimental fluidization runs are carried out in the model reactor with two packing materials, first with “feroxide” and then with sand.

As noted earlier (Section 3.2.2) the maximum air supply was 500 m³/h and because of feroxide density, fluidization did not occur when it was used because the air capacity was insufficient for fluidization.

The inlet gas flow needed to reach minimum fluidization from calculation in Appendix D1, is 622.5 m³/h.
Experimental Results and Discussions

As noted in Section (3.1.1), this is why sand was substituted for "feroxide" as the packing material.

4.1.4.1 Reproducibility of the Results.

In presenting the results of a technical investigation or study it is essential to be able to tell quantitatively how reproducible they might be. For this reason, replication of runs were made and the results of the replications were used to establish the reproducibility of the results by quoting mean value, variance and the percentage error involved. Judgments are then made to the validity of the conclusions.

In the present study the only experimental part that could be treated in this way to obtain statistical data was the fluidization, whereas the other parts were essentially yes or no investigations. However, it is here noted that the reproducibility of these tests was found to be complete. In other words, on replicating the conditions, the result was the same, namely flow or no flow, as the case might be.

Five replicate tests were made using a $2^3$ factorial design as discussed in Chapter 2. The results of pressure drop measurement are given in Appendix J and in Figure 4.1.
showing that a good reproducibility was obtained, the absolute deviation from the mean of the five replications was 8 %, although a maximum as high as 12 % was observed.

The average reproducibility of 8 %, was considered good for this type of experiment due to the nature of the solid phases, which are random systems and cannot be exactly duplicated [Fayed and Otten, 1984].
Figure 4.1. Reproducibility of the results.
Before the results are analyzed it is necessary to establish the magnitude of the error involved so that the effect of the parameters can be correctly assessed. The overall or pooled standard deviation needed for the evaluation of experimental error is found from equation (2.24).

On substitution of the variances from Table 4.1 into equation (2.24) gives $S_{\text{overall}} = 3.5 \text{ cm H}_2\text{O}$. The 3 $S$, or, the 99 % confidence interval, is $=3 \times 3.5 \text{ cm H}_2\text{O} = 10.5 \text{ cm H}_2\text{O}$. 
<table>
<thead>
<tr>
<th>Run</th>
<th>Sand Size Distribution,%</th>
<th>Velocity of Solids (Us), m/s</th>
<th>Velocity of gas (U/o), m/s</th>
<th>Pressure Drop</th>
<th>Average cm H₂O</th>
<th>Variance $\Sigma^2$</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.25-0.5 mm</td>
<td>0.5-1 mm</td>
<td>1-2 mm</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1</td>
<td>33.33</td>
<td>33.33</td>
<td>33.33</td>
<td>0.01</td>
<td>0.13</td>
<td>54.6</td>
</tr>
<tr>
<td>2</td>
<td>33.33</td>
<td>33.33</td>
<td>33.33</td>
<td>0.01</td>
<td>0.26</td>
<td>71.4</td>
</tr>
<tr>
<td>3</td>
<td>33.33</td>
<td>33.33</td>
<td>33.33</td>
<td>0.02</td>
<td>0.13</td>
<td>61.8</td>
</tr>
<tr>
<td>4</td>
<td>33.33</td>
<td>33.33</td>
<td>33.33</td>
<td>0.02</td>
<td>0.26</td>
<td>71.8</td>
</tr>
<tr>
<td>5</td>
<td>45</td>
<td>45</td>
<td>10</td>
<td>0.01</td>
<td>0.13</td>
<td>62.6</td>
</tr>
<tr>
<td>6</td>
<td>45</td>
<td>45</td>
<td>10</td>
<td>0.01</td>
<td>0.26</td>
<td>81.2</td>
</tr>
<tr>
<td>7</td>
<td>45</td>
<td>45</td>
<td>10</td>
<td>0.02</td>
<td>0.13</td>
<td>68</td>
</tr>
<tr>
<td>8</td>
<td>45</td>
<td>45</td>
<td>10</td>
<td>0.02</td>
<td>0.26</td>
<td>83.6</td>
</tr>
</tbody>
</table>
Experimental Results and Discussions

From Table 4.1 the variance at the run number 1 is 7.3. This is the lowest value compared to other runs, whilst the variance at the run number 4 is 18.7 and is the highest value. The average variance is 12.5, and was considered good for this type of experiment due again to the nature of solids phases, which are random systems and cannot be exactly duplicated.

Using factorial design, referred to Section (2.7), the experimental results can be seen in Table 4.1. The effects of main parameters and their interactions, as found by calculation in Appendix G, are show in Table 4.2. The Pareto chart is show in Figure 4.2 with the confidence interval 10.5 cm H₂O given earlier.

Table 4.2 Calculated main effects and interactions

<table>
<thead>
<tr>
<th>Effect</th>
<th>Pressure Drop (cm H₂O)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Average</td>
<td>69</td>
</tr>
<tr>
<td>Gas Flow (G)</td>
<td>15.85</td>
</tr>
<tr>
<td>Solid Flow (S)</td>
<td>3.25</td>
</tr>
<tr>
<td>Size Distribution (D)</td>
<td>-9.55</td>
</tr>
<tr>
<td>G*S</td>
<td>-1.85</td>
</tr>
<tr>
<td>S*D</td>
<td>0.65</td>
</tr>
<tr>
<td>G*D</td>
<td>1.25</td>
</tr>
<tr>
<td>G<em>S</em>D</td>
<td>0.35</td>
</tr>
</tbody>
</table>
From Table 4.2 the parameters having the biggest effect on fluidization are gas flow (15.85) and size distribution (-9.55). Solids flow has small effect (3.25) while neither the two factors nor the three factor interactions have any effect.

Figure 4.2 The Pareto Chart.
Figure 4.2 with the confidence interval of 3s (10.5 cm H2O) included confirms this and shows that the parameters having the biggest effect on fluidization are the gas flow (15.85). This is the most important effect to improve process performance and it is obvious that we should begin by adjusting this factor. The second largest effect is the size distribution (-9.55) whilst solids flow rate two factor interactions and three factor interactions have no effect, just noise levels.

Next, experimental results are compared to the calculations (see sample calculation, Appendix E) in Table 4.3.

Table 4.3. Comparison of Average Experimental and Calculated results.

<table>
<thead>
<tr>
<th>Run</th>
<th>Experimental (cm H2O)</th>
<th>Calculation (cm H2O)</th>
<th>Us (m/s)</th>
<th>Uo/s (m/s)</th>
<th>ΔU (m/s)</th>
<th>Umf (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>54.6</td>
<td>59.3</td>
<td>0.01</td>
<td>0.13</td>
<td>0.14</td>
<td>0.264</td>
</tr>
<tr>
<td>2</td>
<td>71.4</td>
<td>77.04</td>
<td>0.01</td>
<td>0.26</td>
<td>0.27</td>
<td>0.264</td>
</tr>
<tr>
<td>3</td>
<td>61.8</td>
<td>63.4</td>
<td>0.02</td>
<td>0.13</td>
<td>0.15</td>
<td>0.264</td>
</tr>
<tr>
<td>4</td>
<td>71.8</td>
<td>77.04</td>
<td>0.02</td>
<td>0.26</td>
<td>0.28</td>
<td>0.264</td>
</tr>
<tr>
<td>5</td>
<td>62.6</td>
<td>67</td>
<td>0.01</td>
<td>0.13</td>
<td>0.14</td>
<td>0.187</td>
</tr>
<tr>
<td>6</td>
<td>81.2</td>
<td>77.04</td>
<td>0.01</td>
<td>0.26</td>
<td>0.27</td>
<td>0.187</td>
</tr>
<tr>
<td>7</td>
<td>68</td>
<td>72.12</td>
<td>0.02</td>
<td>0.13</td>
<td>0.15</td>
<td>0.187</td>
</tr>
<tr>
<td>8</td>
<td>83.6</td>
<td>77.04</td>
<td>0.02</td>
<td>0.26</td>
<td>0.28</td>
<td>0.187</td>
</tr>
</tbody>
</table>
In this experiment, the conditions are like those in Figure 2.4 b (Section 2.4), that is solid's move from low to high air pressure. There are two possible conditions for this case, namely stick slip flow and aerated flow, depending on the $\Delta U$ value compared to that of the minimum velocity for fluidization (Umf). Umf for run numbers 1,2,3,4 is for size distribution 1 and is 0.264 m/s (Appendix D), and Umf for run numbers 5,6,7,8 is for size distribution 2 and is 0.187 (Appendix D).

Using equation (2.11a) and on replacing given values we find $\Delta U$ as tabulated in Table 4.3. In run numbers 1,3,5,7 the condition, according to theory (Section 2.4), is stick slip flow because the linear velocity of gas relative to solids ($\Delta U$) is lower than Umf and pressure drop ($\Delta P$) can be calculated with equation (2.10).

Run numbers 2,4,6,8 are aerated flow of solids condition because the linear velocity of gas relative to solids ($\Delta U$) is higher than Umf and pressure drop ($\Delta P$) can be calculated with equation (2.12).

From Table 4.3, the average difference between experimental and calculated results are 8%.

This difference is considered reasonable because the experiments involving packing are those of a random system and can not exactly be reproduced.

The Table 4.3 shows average pressure drop values from experiment are compared to $\Delta U$ for Stick slip flow condition and for Aerated flow condition as shown Table 4.4.
Table 4.4 Pressure Drop for Stick Slip Flow Condition and Aerated Flow Condition

<table>
<thead>
<tr>
<th>No</th>
<th>Condition</th>
<th>$\Delta U$ (m/s)</th>
<th>Pressure Drop (Cm H20)</th>
<th>Size Distribution 1</th>
<th>Size Distribution 2</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Stick Slip Flow</td>
<td>0.14</td>
<td>54.6</td>
<td>62.6</td>
<td></td>
</tr>
<tr>
<td>2</td>
<td>Stick Slip Flow</td>
<td>0.15</td>
<td>61.8</td>
<td>68</td>
<td></td>
</tr>
<tr>
<td>3</td>
<td>Aerated Flow</td>
<td>0.27</td>
<td>71.4</td>
<td>81.2</td>
<td></td>
</tr>
<tr>
<td>4</td>
<td>Aerated Flow</td>
<td>0.28</td>
<td>71.8</td>
<td>83.6</td>
<td></td>
</tr>
</tbody>
</table>

From Table 4.4 with an increase in $\Delta U$ the pressure drop increases too. This is according to equation (2.10) pressure drop for stick-slip condition being proportional to $\Delta U$. Table 4.4 also shows that the pressure drop for size distribution 2 is higher than for size distribution 1. As size distribution 2 has a mean particle diameter lower than that of size distribution 1, then according to equation (2.10) pressure drop for stick-slip condition is inversely related to particle diameter, and decreasing mean particle size will increase pressure drop. From Table 4.4 for aerated flow with increase $\Delta U$ the pressure drop remains relatively constant.

From all these experimental fluidization runs in the Hyl III model reactor, clogging was not caused by fluidization, because of aerated flow condition (same condition for fluidization) clogging did not occur for the stick slip flow condition. In the Hyl III reactor,
Experimental Results and Discussions

Fluidization could not occur because the actual pressure drop was lower than the calculated minimum pressure drop for fluidization. Because in the Hyl III reactor fluidization can not occur then, accordingly, clogging due to fluidization can not occur also. On this basis the several times, observed clogging in the outlet of the Hyl III reactor must be due to some other cause and not by fluidization or arching.

4.1.5 Flow interruption with agglomerates.

Following the procedure given in Section 3.3.2.2, the experimental results obtained for agglomerates and sand are summarized in Table 4.5.

Table 4.5. Experimental Results of Flow Interruptions with Agglomerates.

<table>
<thead>
<tr>
<th>NO</th>
<th>MIX A/S</th>
<th>AIR</th>
<th>( \Delta U \text{(cm H}_2\text{O)} )</th>
<th>CLOSED OUTLET (YES/NO)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>1/5</td>
<td>NO</td>
<td>-</td>
<td>NO</td>
</tr>
<tr>
<td>2</td>
<td>1/4</td>
<td>NO</td>
<td>-</td>
<td>NO</td>
</tr>
<tr>
<td>3</td>
<td>7/24</td>
<td>NO</td>
<td>-</td>
<td>YES</td>
</tr>
<tr>
<td>4</td>
<td>1/3</td>
<td>NO</td>
<td>-</td>
<td>YES</td>
</tr>
<tr>
<td>5</td>
<td>1/2</td>
<td>NO</td>
<td>-</td>
<td>YES</td>
</tr>
<tr>
<td>6</td>
<td>1/5</td>
<td>YES</td>
<td>73</td>
<td>NO</td>
</tr>
<tr>
<td>7</td>
<td>1/4</td>
<td>YES</td>
<td>70</td>
<td>NO</td>
</tr>
<tr>
<td>8</td>
<td>7/24</td>
<td>YES</td>
<td>68</td>
<td>NO</td>
</tr>
<tr>
<td>9</td>
<td>1/3</td>
<td>YES</td>
<td>64</td>
<td>YES</td>
</tr>
<tr>
<td>10</td>
<td>1/2</td>
<td>YES</td>
<td>62</td>
<td>YES</td>
</tr>
</tbody>
</table>
In the experiments with mixtures of agglomerate and sand in the ratio of 1/5 and 1/4, it was observed that agglomerates and sand moved together out from the outlet of the model reactor and this did not result in clogging or arching (Table 4.5). The reason that clogging or arching did not occur is almost certainly because of the amount of agglomerates was very small compared to that of sand and there was little probability of agglomerates meeting with other agglomerates. When only sand flows out from the outlet of the model reactor, the diameter of particles involved (0.25-2 mm) was very small compared to the outlet diameter of the model reactor, i.e. 228-28.5 times the particle diameter, so flow interruption and arching is most unlikely.

When experiments with mixtures of agglomerate and sand in the ratio of 7/24 were carried out it was observed that although agglomerates and sand also again moved together out from the outlet Hyl III model reactor, clogging at the outlet of the model reactor did result this is because, the ratio of agglomerates is higher than before so that the probability of agglomerates meeting other agglomerates is also higher. When three or more agglomerates meet at the outlet of the model reactor arching can occur because the diameter of agglomerates involved (approximately 20 mm) is only 2.85 times or more compared to the outlet model reactor (57 mm), and this, as noted earlier, can readily lead to interlocking.
Experimental Results and Discussions

When the experiments with mixtures of agglomerate and sand in the ratio of 1/3 and 1/2 were carried out, it was observed that agglomerates and sand also moved downward together and also that clogging or arching at the outlet model reactor resulted. As the ratio of agglomerates is still higher in this condition than before, the probability of agglomerates meeting other agglomerates is also higher.

The experiments with mixtures of agglomerate and sand in the ratio of 1/5 and 1/4 and using some air with flow rate of 500 m$^3$/h were carried out. It was observed that agglomerates and sand moved together out from the outlet of the model reactor. This condition did not result in clogging or arching (Table 4.5). The reason that clogging or arching did not occur was almost certainly the same as the case of no air, namely the amount of agglomerates was very small compared to that of sand and in turn this prevented agglomerates meeting other agglomerates. When only some sand flew out from the outlet of the model reactor, for similar reasons, flow interruptions and arching was most unlikely. In the existence of air flow, this condition is the same as the stick slip flow condition, the solids moved downward from low air pressure to high air pressure.

On the experiment with mixtures of agglomerate and sand in the ratio of 7/24 and using some air with flow rate of 500 m$^3$/h, outlet closure was not observed. It was observed that again agglomerates and sand move together out from the outlet of the model reactor. This condition, unlike that for the no air case, did not result in clogging or arching (Table 4.5). The reason that clogging or arching did not occur was because the air decreased the
possibility of agglomerates meeting other agglomerate. The explanation about that can be seen in Figure 4.3 and Figure 4.4. With mixture of agglomerates and sand in the ratio 7/24, one layer in the model reactor was observed. This layer consisted of three agglomerates and some amount of sand. The agglomerates dispersed equally, one in the center, and the other two were at right and left side. With velocity of solid (Us) of 0.02 m/s and velocity of gas (Uo/e) of 0.26 m/s, the linear velocity of gas relative to solids (AU) was 0.28 m/s downward (equation 2.11a). Minimum fluidizing velocity (Umf) for agglomerate from Appendix F is 3.62 m/s, and for sand with size distribution 1 from Appendix D is 0.264 m/s.

In the case of sand, AU is higher than Umf which is aerated flow condition (Section 2.4), and for agglomerate, AU is lower than Umf that is stick slip flow condition (Section 2.4). The overall condition was agglomerates and sand moved downward with AU of 0.28 m/s.

Figure 4.3 shows the agglomerates and sand in the model reactor. Some part of the gas coming to the model reactor flew to the bottom of reactor and the rest went to the top. After the pressure of gas at the bottom had been the same as that at the incoming gas, it moved upward to cone 1 and cylinder because the top part has lower pressure. The gas pressure mixed in the centerline from round about cone 1 (Figure 4.4 step 1), and then in the centerline the pressure of gas was higher than round about cone 1. This means in the centerline drag force was higher than round about cone 1. The pressure of gas suspended one agglomerates in the Hyl III reactor model centerline. One agglomerate on the right and
one agglomerate on the left accept lower gas pressure than that in the centerline and then the velocity of the two agglomerates was faster than one in the centerline (Figure 4.4 step 2). After a period of time, all agglomerates went down from the bottom of cone 1 to the bottom of cone 2. However, the agglomerate in the center moved down a little bit slower than the other two on the left and right side because it was restricted by the gas pressure since the gas pressure in the center was higher than that on the left and the right side (Figure 4.4 step 3). In the cone 2 region the gas pressure is the same and then two agglomerates and sand move downward just by gravity force (Figure 4.4 step 4), finally the two agglomerates were out together in the outlet of model reactor, without any clogging occurred (Figure 4.4 step 5), because the outlet diameter (57 mm) is still bigger than the sum of the two agglomerates (40 mm).
Figure 4.3 Agglomerates and sand in the model reactor with 7/24 ratio using air.
Figure 4.4 A step by step flow of agglomerates and sand with 7/24 ratio using air.
Experimental Results and Discussions

When experiments with mixtures of agglomerate and sand in the ratio of 1/3 and 1/2 using some air with flow rate 500 m$^3$/h were carried out, it was observed that agglomerates and sand also moved together but then clogging or arching at the outlet model reactor resulted. This occurred under condition the ratio of agglomerates in the this experiment is higher than that in the previous experiment and the probability of agglomerates meeting other agglomerates was also higher.

The reason that clogging or arching resulted can be explained with Figure 4.5 and Figure 4.6. As the ratio of agglomerates used in this experiment was higher than before, then in one layer the number of agglomerates was large too. In one layer the number of agglomerates was seven (Figure 4.5 or Figure 4.6 step 1). One agglomerate was located in the center, three on the left and another three on the right side. And then with the same gas pressure as before, the drag force for seven agglomerates was lower than that with three agglomerates in the previous experiment so that the drag force suspending these agglomerate was lower compared to that suspending three agglomerates (Figure 4.6 step 2). Because the drag force was low, just one or two agglomerates can be suspended in the centerline and the other agglomerates have higher velocity than those in the centerline (Figure 4.6 step 3).

After a time period, one or two agglomerate in the centerline was left compared to other agglomerates (Figure 4.6 step 3), while the agglomerate went down from the bottom cone to the bottom of cone 2 (Figure 4.6 step 4). In the cone 2 section the gas pressure is
the same and then agglomerates and sand move downward just by gravity force. Finally, three or more agglomerates closed the model reactor outlet, resulting clogging, due to the diameter of outlet (57 mm) is smaller than the sum of three or more agglomerates’ diameter (60 mm or more).

Figure 4.5 Agglomerates and sand in the model reactor with 1/3 ratio using air.
Figure 4.6 A step by step flow of agglomerates and sand with 1/3 ratio using air
From all experiments on flow interruption in the Hyl III model reactor, the ratio of agglomerates in the Hyl III model reactor was influential on causing clogging. If the ratio of agglomerates is big, the probability of clogging will be high too and if the ratio of agglomerates is small the probability of clogging will also be small. The presence of the gas flow in the reactor resists the movement of agglomerates. When a few agglomerates present in the mixture, some agglomerates moving downward are not sufficient to stop the flow, but when a great number of agglomerates are present there will be more moving downward and stopping the flow.

From the results of all the calculations and experiments done in the model, it is shown that flow interruption in the Hyl III model reactor containing agglomerates occurs mostly caused by the presence of agglomerates. The presence of gas retards some of the agglomerates moving down, so that it help the flow of solids in a way that the agglomerate will not move down together at the same time resulting the absence of clogging. This occurs when the agglomerate consist of low composition of agglomerate and sand. However, in a higher composition of agglomerates the gas flow does not have enough energy to prevent clogging.

4.1.6 Application in Hyl III Reactor

As described in the preceding section, the possible solids flow interruptions in an Hyl III reactor is crucial problem. Based on the calculations about an arching
possibility (as it is describe in section 4.1.1 and section 4.1.2), about fluidization possibility (as it is describe in section 4.1.3 and 4.1.4), and from experimental results, the solids flow interruptions that has been observed in daily operation, might not because of arching or fluidization, but must have been due to agglomerates formation. As shown in Table 4.5, the presence of agglomerates equal to 7/24 or more of the solids without air flow will cause flow interruptions, and of agglomerates equal to 1/3 or more with air flow will cause flow interruption too. Although these conditions in the model are not really the same with the actual conditions, these facts indicate that the more agglomerates in the solids the more its tendency to have flow interruptions in the reactor.

From observation in the actual Hyl III reactor when sticking occurs, the agglomerates are formed with size between 100 mm and 200 mm and with the ratio between agglomerates and direct reduction iron is higher than 1/3. Therefore the clogging is eventually happening.

To prevent flow interruptions formation of agglomerates must be restricted.

Finally, it may be expected that the results in all the investigations of this research may be advantageously used as basic information for any activity relating to flow interruptions (flow system) in operating the Hyl III reactor area in general, and in PT Krakatau Steel in particular.
Chapter Five

CONCLUSIONS

The results of the present study have shown that in the Hyl III reactor arching does not occur and neither does fluidization. Clogging occurs caused by sticking or agglomerate formation. This closes the Hyl III reactor outlet. In line with the experiment in the Hyl III model reactor, clogging is caused by the formation of an abundance of agglomerates in the model reactor.

A few agglomerates cannot cause clogging but a great number of agglomerates can, because agglomerates meet each other and cause interlocking in the outlet and close it. The use of air can prevent the occurrence of clogging because the pressure gradient (drop) can reduce the velocity of agglomerates and then just one or two agglomerates meet in the reactor outlet, resulting in no clogging. If the proportion of agglomerates is large, the drag force cannot reduce the velocity of agglomerates so that they meet each other in the reactor outlet, leading to the occurrence of clogging.
REFERENCES


LKAB,1995,” A Study of Pellet sticking”, internal Report,Malmberget.


A. Calculation of Arching in the Hyl III Reactor

The calculation of arching in order to determine the arching phenomenon, requires the use of Convex Wall Yield Locus chart and Flow Factor chart after calculating the arching equation.

The arching equation is:

\[ \sigma_1 = \frac{(B \times \gamma)}{H(\alpha)} \]

where:

- \( \sigma_1 \) refers to the minimum stress in the abutment of an arch at this point.
- \( B \) refers to the estimate of minimum discharge opening.
- \( \gamma \) refers to the bulk density of iron ore. According to the measurement, the value of \( \gamma \) is 1700kg/m^3
- The function \( H(\alpha) \) depends on the outlet shape and hopper half angle (\( \alpha \)).
The value of the function $H(\alpha)$ can be read out from graphic of $H(\alpha)$ function exhibited on figure 2.8. Since the actual of $\alpha = 16^\circ$ and outlet shape circular, it is found that $H(\alpha) = 2.3$.

- $\sigma_1$ refers to the major consolidating pressure at a point in the hopper.
- $ff$ refers to flow factor. Flow factor which describes the stress condition in the hopper during flow. The flow factor is given in the equation (2.7).

$$ff = \frac{\sigma_1}{\sigma_1}$$

From equation 2.1 it is found that

$$\sigma_1 = \frac{(B \ast \gamma)}{H(\alpha)} = \frac{(0.5 \ast 1700)}{2.3} = 369.56 \text{ Pa.}$$

Using equation (2.7) and with estimate flow factor ($ff$) as 1.1,

$$\sigma_1 = ff \ast \sigma_1 = 1.1 \ast 369.56 = 406.51 \text{ Pa}$$

Having the value of $\sigma_1 = 406.51 \text{ Pa}$ and the estimate angle of friction ($\delta$) of $50^\circ$, the next step is founding the value of kinematic angle of friction between the solid and the bin wall ($\phi$) using Convex Wall Yield Locus chart.

The procedure to make Convex Wall Yield Locus chart are as follows:

1. Construct the Effective Yield Locus (EYL) with effective angle of friction ($\delta$) of $50^\circ$.

2. At the point $\sigma_1$ construct the Mohr Semi Circle through this tangent to the EYL (Figure A1).
Appendix

3. Plot the wall yield locus from measurement in the Convex Wall Yield Locus chart.

4. The line through the origin and the point of intersection of the Wall Yield Locus (WYL) and the Mohr Semi Circle.

The steps to make Convex Wall Yield Locus chart can be seen in Figure A1.

From this figure it is found the kinematic angle of friction between the solid and the bin wall ($\phi$) is 32°.

The last step is determining the actual value of flow factor ($f_f$) based on the flow factor chart for ($\delta$) = 50° (see Figure A2). With the value of ($\phi$) = 32°, $\alpha$ = 16° it is found that flow factor was beyond the chart.

Therefore, the estimate value of B = 0.5 m cannot be accepted

If $f_f$ varies by more than about 10% from assumed value, a new estimate for B is made and calculation repeated.
Figure A1. Convex Wall Yield Locus with $\delta=50^\circ$ for The Hyl III Reactor.
Figure A2. Flow Factor For Delta ($\delta$)=50° [Arnold, 1992].
After several times of trial and error the estimate of \( B = 0.3 \text{ m} \) result in 
\( \sigma_1 = 221.74 \text{ Pa} \) and therefore \( \sigma_1 = 266 \text{ Pa} \). with estimate flow factor as 1.2.

Using the Convex Wall Yield Locus chart where estimated effective angle of 
friction \( (\delta) = 60^\circ \) and \( \sigma_1 = 266 \text{ Pa} \), it is found that \( (\phi) = 29^\circ \) (see Figure A3).

Using Flow Factor chart, for \( (\delta) = 60^\circ \), \( (\phi) = 29^\circ \), and \( \alpha = 16^\circ \) (see Figure A4) it is 
found that flow factor was 1.3

The difference of \( \text{ff} = \frac{(\text{ff}_{\text{calculate}} - \text{ff}_{\text{estimate}}) \times 100\%}{\text{ff}_{\text{estimate}}} \)

\[ = \frac{(1.3 - 1.2) \times 100\%}{1.2} = 8.3\% \]

Since the difference of 8.3% is lower than 10%, it is assumed that the value of \( B = 0.3 \text{ m} \) can be accepted.

Figure A3. Convex Wall Yield Locus With \( \delta = 60^\circ \) for The Hyl III Reactor.
Figure A4. Flow Factor For Delta (δ) = 60° [Arnold, 1992]
Appendix

B. Calculation of Arching in the Hyl III Model Reactor

Arching in the Hyl III model reactor is calculated in the same manner with the calculation in Appendix A. The bulk density (\( \gamma \)) is 11250 kg/m\(^3\), hopper half angle (\( \alpha \)) = 16°.

After several times of trial and error, the estimate of B = 0.03 m result in \( \sigma_1 = 16.3 \) Pa and therefore \( \sigma_1 = 21.19 \) Pa. with estimate flow factor as 1.3.

Using the wall yield locus chart where estimated effective angle of friction (\( \delta \)) = 60°, \( \sigma_1 = 21.19 \) Pa, it is found that (\( \phi \)) = 28° (see Figure A5).

Using flow factor chart, for \( \delta = 60^\circ \), (\( \phi \)) = 28°, and \( \alpha = 16^\circ \) (see Figure A6) it is found that flow factor was 1.2.

The difference of ff = ((ff \(_{\text{calculate}}\) - ff \(_{\text{estimate}}\)) \times 100%) / ff \(_{\text{estimate}}\)

= ((1.3 - 1.2) \times 100%) / 1.2 = 8.3 %

Since the difference of 8.3 % is lower than 10 %, it is assumed that the value of B = 0.03 m can be accepted.
Figure A5. Convex Wall Yield Locus with $\delta=60^0$ for The Hyl III Reactor Model.
Figure A6. Flow Factor For Delta (δ)=60°[Arnold,1992].
C. Calculation of Fluidization in the Hyl III Reactor.

Average cooling gas inlet capacity is 50,615 m$^3$/h at a pressure of 5.27 kg/cm$^2$ and that at the outlet is 47,520 m$^3$/h at a pressure of 4.17 kg/cm$^2$.

Voidage for sponge from measurement is 0.35, solid density of sponge iron with bulk density 1,700 kg/m$^3$ (measurement) is 2,615 kg/m$^3$ and the distance between the cooling gas inlet and outlet is 8 m.

Calculated minimum pressure drop for fluidization according to equation (2.11b) is

$$\Delta P = \rho_s (1 - \varepsilon) g \Delta h$$

$$= 2,615 (1 - 0.35) (8) (9.8) \text{ kg/m}^3$$

$$= 133,280 \text{ Pa.}$$

D. Calculation of Fluidization in the Hyl III model reactor

D1. For Feroxide

Feroxide has a bulk density (from measurement) of 3,000 kg/m$^3$ and its voidage is 0.35 so solid density is 4,615.4 kg/m$^3$. The distance between air inlet and outlet is 0.62 m.

From equation (2.11b) pressure drop for fluidization is

$$\Delta P = \rho_s (1 - \varepsilon) g \Delta h$$

$$= 4615.4 (1 - 0.35) (9.8) (0.62)$$

$$= 18,228 \text{ Pa or 185.9 cm H}_2\text{O}$$
Superficial velocity at minimum fluidizing velocity for small particles of feroxide, from equation (2.8) is

\[ U_{mf} = \frac{d^2 (\rho_s - \rho_g)g}{1650 \mu} \]

\[ = \frac{(6.74 \times 10^{-4})^2 (4615.4 - 1.186) (9.8)}{(1650 (1.9 \times 10^{-5}))} \]

\[ = 0.655 \text{ m/s} \]

Linear Velocity

\[ V = \frac{U_{mf}}{\varepsilon} = \frac{0.655}{0.35} = 1.871 \text{ m/s} \]

Cross Section area

\[ A = \pi \frac{d^2}{4} = \pi (0.343)^2 / 4 = 9.24 \times 10^{-2} \text{ m}^2 \]

Inlet Gas Flow Rate

\[ VA = 1.871 \times 9.24 \times 10^{-2} = 0.17292 \text{ m}^3/\text{s} \]

\[ = 622.5 \text{ m}^3/\text{h} \]

**D2. For Sand**

Sand has a bulk density of 1,250 kg/m\(^3\) which is lower than Feroxide. Voidage from measurement 0.401 and mean diameter for size distribution 1 is 6.43 \times 10^{-4} \text{ m}, similar calculations as in Appendix D1,

\[ U_{mf} = \frac{d^2 (\rho_s - \rho_g)g}{1650 \mu} \]

\[ = \frac{(6.43 \times 10^{-4})^2 (2,086 - 1.186) (9.8)}{(1650 (1.9 \times 10^{-5}))} \]

\[ = 0.264 \text{ m/s} \]

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Linear velocity

\[ V = \frac{U_{mf}}{e} = \frac{0.264}{0.401} = 0.6719 \text{ m/s} \]

Cross section area is \( 9.24 \times 10^{-2} \text{ m}^2 \)

Inlet Gas Flow Rate

\[ V.A. = (0.6719) (9.24 \times 10^{-2}) (3600) \]
\[ = 223.51 \text{ m}^3/\text{h} \]

Superficial minimum fluidizing velocity for small sand particles with size distribution 2 with dp mean is 0.536.

\[ U_{mf} = \frac{d^2 (\rho_s - \rho_g) g}{1650 \mu} \]
\[ = \frac{(5.36 \times 10^{-4})^2 (2,086 - 1.186) (9.8)}{(1,650) (1.9 \times 10^{-5})} \]
\[ = 0.187 \text{ m/s} \]

E. Calculation of Pressure Drops

Example in Run Number 1 (A stick-slip flow condition)

Calculation of pressure drop with velocity of gas \((U_0/e)\) is 0.13 m/s and the velocity of solid is 0.01 m/s. The linear velocity of gas relative to solid \((\Delta U)\) calculated with equation \((2.11a)\):

\[ \Delta U = (U_0 / e) + U_s \]
\[ \Delta U = (0.13) + 0.01 \]
\[ \Delta U = 0.14 \text{ m/s} \]

Minimum velocity of fluidization \((U_{mf})\) from Appendix D2 is 0.264 m/s so
\( \Delta U \) is lower than \( U_{mf} \) which is a stick-slip flow condition (Section 2.4).

Calculating pressure drop with equation (2.10) and on replacing given values we find experimental pressure drop as

\[
\Delta P = \frac{150(1-0.401)^2(1.9E-5)(0.14)}{(0.401)^3(0.8*6.43E-4)^2} + \frac{1.75(1-0.401)(1.186)(0.14)^3}{(0.401)^3(0.8*6.43E-4)}
\]

\( \Delta P = 5809 \text{ Pa or } 54.6 \text{ cm H}_2\text{O} \)

\( L = \text{distance inlet and outlet air to measure pressure drop}=0.62 \text{ m} \)
\( \varepsilon = \text{voidage from measurement}=0.401 \)
\( \mu = \text{air viscosity}=1.9E-5 \text{ N.s/m}^2 \)
\( \rho_g = \text{density of gas}=1.186 \text{ kg/m}^3 \)
\( \phi_s = \text{sphericity of particle from measurement}=0.8 \)
\( d_p = \text{mean diameter particle}=6.43E-4 \text{ m} \)

Example in Run Number 4 (Aerated flow condition)

Calculation of pressure drop with velocity of gas \( (U_0/\varepsilon) \) is 0.26 m/s and the velocity of solid is 0.02 m/s. The linear velocity of gas relative to solid \( (\Delta U) \) calculated with equation (2.11a):

\[
\Delta U = \left( \frac{U_0}{\varepsilon} \right) + U_s
\]

\( \Delta U = (0.26) + 0.02 \)

\( \Delta U = 0.28 \text{ m/s} \)

Minimum fluidization velocity \( (U_{mf}) \) from Appendix D2 is 0.264 m/s so \( \Delta U \) is higher than \( U_{mf} \) which is aerated flow of solids (Section 2.4). Calculating pressure drop with equation (2.12) and on replacing given values we find experimental pressure drop as
\[ \Delta P = 2086 \times (1 - 0.401) \times (9.8) \times (0.62) \pm |\Delta P|_{\text{friction}} \]

with

\[ \mu = \text{apparent viscosity} = 1.9 \times 10^{-5} \text{ N.s/m}^2 \]

\[ U_s = \text{the mean velocity of solid} = 0.02 \]

\[ d_t = \text{effective diameter} = 0.34 \text{ m} \]

\[ L = \text{distance example measurement pressure drop} = 0.62 \text{ m} \]

\[ \rho_s = \text{density of solid} = 2,086 \text{ kg/m}^3 \]

\[ g = \text{gravity} = 9.8 \text{ m/s}^2 \]

\[ \Delta H = \text{same as L} \]

Calculating \(|\Delta P|_{\text{friction}}|\) with equation (2.14) and replacing given values we find

\[ |\Delta P|_{\text{friction}}| = (32) (1.9 \times 10^{-5}) (0.02) (0.62)/ (0.34)^2 \]

\[ |\Delta P|_{\text{friction}}| = 0.000065 \text{ Pa}. \]

\[ \Delta P = (7592.04 + 0.000065) \text{ Pa} \]

\[ = 7592.040065 \text{ Pa or} \]

\[ = 77.43 \text{ cm H}_2\text{O} \]

**G. Calculation Main Effect and Interactions**

From Table 4.1 calculated main effect and interactions

Gas Flow Effect (G)

\[ G(+) = 1/4 (71.4 + 71.8 + 81.2 + 83.6) = 77 \]

\[ G(-) = 1/4 (54.6 + 61.8 + 62.6 + 68) = 61.15 \]

\[ G = G(+) - G(-) = 77 - 61.15 = 15.85 \]
Solid Flow effect (S)
\[ S(+) = \frac{1}{4}(61.8+71.8+68+83.6) = 70.7 \]
\[ S(-) = \frac{1}{4}(54.6+71.4+62.6+81.2) = 67.45 \]
\[ S = S(+) - S(-) = 70.7 - 67.45 = 3.25 \]

Size Distribution Effect (D)
\[ D(+) = \frac{1}{4}(54.6+71.4+61.8+71.8) = 64.3 \]
\[ D(-) = \frac{1}{4}(62.6+81.2+68+83.6) = 73.85 \]
\[ D = D(+) - D(-) = 64.3 - 73.85 = -9.55 \]

Two Factor Interaction

G*D interaction:
\[ = \frac{1}{4}((Y1+Y3+Y6+Y8) - (Y2+Y4+Y5+Y7)) \]
\[ = \frac{1}{4}((54.6+61.8+81.2+83.6) - (71.4+71.8+62.6+68)) \]
\[ = 1.25 \]

G*S interaction
\[ = \frac{1}{4}((Y1+Y4+Y5+Y8) - (Y2+Y3+Y6+Y7)) \]
\[ = \frac{1}{4}((54.6+71.8+62.6+83.6) - (71.4+61.8+81.2+68)) \]
\[ = -1.85 \]

S*D interaction
\[ = \frac{1}{4}((Y1+Y2+Y7+Y8) - (Y3+Y4+Y5+Y6)) \]
\[ = \frac{1}{4}((54.6+71.4+68+83.4) - (61.8+71.8+62.6+81.2)) \]
\[ = 0.65 \]

Three Factor Interaction

G*S*D = \[ \frac{1}{4}((Y2+Y3+Y5+Y8) - (Y1+Y4+Y6+Y7)) \]
\[ = \frac{1}{4}((71.4+61.8+62.6+83.6) - (54.6+71.8+68+81.2)) \]
\[ = 0.35 \]
F. Calculation of Fluidization for Agglomerates in the Hyl III Model Reactor

Agglomerate has a bulk density (from measurement) of 993 kg/m$^3$ and solid density is 1947 kg/m$^3$, calculate voidage (see Appendix I) is 0.49. The distance between air inlet and outlet is 0.62 m.

Superficial minimum fluidizing velocity for large particles of agglomerate, from equation (2.9) is

$$Um^f = \frac{dp (\rho_s - \rho_g)g}{(24.5 \times \rho_g)}$$

$$= \frac{(0.02) (1947 - 1.186) (9.8)}{(24.5 \times 1.186)}$$

$$= 13.12 \text{ m}^2/\text{s}^2$$

$$Um^f = 3.62 \text{ m/s}$$

Linear Velocity

$$V = \frac{Um^f}{\varepsilon} = 3.62 / 0.49 = 7.39 \text{ m/s}$$

Cross Section Area

$$A = \pi \frac{d^2}{4} = \pi \left(0.343\right)^2 / 4 = 9.24 \times 10^{-2} \text{ m}^2$$

Inlet Gas Flow Rate

$$VA = 7.39 \times 9.24 \times 10^{-2} = 0.68 \text{ m}^3/\text{s}$$

$$= 2,458.2 \text{ m}^3/\text{h}$$
H. An Example Calculation of Average Diameter of Sand Mixture:

From equation (2.17):

\[
dp = \frac{1}{\sum (x/dp)_{i}}
\]

\[
dp = \frac{1}{[0.888+0.444+0.222]}
\]

\[
dp = 0.643 \text{ mm or } 6.43 \times 10^{-4} \text{ m}
\]

I. Calculated Packed Bed Porosity

From equation (2.20):

\[
\gamma = \rho_s (1-\varepsilon)
\]

If sand with size distribution 1 and mean diameter (dp) = 6.43 \times 10^{-4} m;

\[
\rho_s = 2,086 \text{ kg/m}^3 \text{ and } \gamma = 1,250 \text{ kg/m}^3
\]

\[
1,250 = 2,086 (1-\varepsilon)
\]

\[
\varepsilon = 0.401
\]
Appendix

J. Reproducibility of The Result and An Example

Calculation Reproducibility

Table J.1. Reproducibility of The Result

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<th>Series3</th>
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An Example Calculation Reproducibility of The Result.

From equation (3.1) for run number 3:

\[ R = \frac{0.397306}{5} \times 100\% \]

\[ = 7.9\% \]