Use of a Baffle to Enhance Distribution of a Liquid Sprayed into a Gas-Solid Fluidized Bed

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Abstract

Bubbles in gas-solid fluidized beds are generally beneficial as they promote solids mixing, heat transfer and mass transfer. In most fluidized beds, the local bubble gas flux varies greatly over the cross-section. A novel triboelectric method is developed to measure the bubble gas distribution in a fluidized bed. A correlation relates the local bubble gas flux to the triboelectric signal generated by the impact of the gas bubbles on a triboprobe. Several signal analysis tools, such as power spectrum, cycle analysis and signal moments, were used to determine the best experimental fit for the profile of the bubble gas flux. The triboelectric method is used to study the impact of baffle and fluxtube on the distribution of the gas bubbles.

Efficient and uniform liquid feed distribution in Fluid Cokers™ enhances the yield of valuable products and the coker operability by reducing the formation of wet agglomerates. A promising method to improve liquid distribution could be the modification of bed hydrodynamics and mixing characteristics using simulated ring baffles with and without fluxtubes. In small scaled-down fluidized bed, such a baffle changed the fluidized bed hydrodynamics by redirecting gas bubbles above the baffle region, directed towards the jet spray, which improved liquid distribution by reducing agglomerate formation. The experimental results show that the best liquid distribution is obtained when the tip of the liquid injection nozzle is aligned just above the baffle tip. As long as the baffle angle with the vertical is less than 45°, this will also prevent the formation of any deposit on the baffle.

**Key words:** Triboelectricity, Bubble, Baffle, Power Spectrum, Triboprobe, Bubble Flux, Fluidized Bed, Agglomeration, Fluxtube, Bubble Distribution
## Co-Authorship Statement

### Chapter 2

**Article Title:**
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**Authors:**
Majid Jahanmiri, Cedric Briens, Franco Berruti, Jennifer McMillan, Francisco Javier Sanchez

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Majid Jahanmiri conducted all the experimental work, analyzed the data and wrote the manuscript. This work was jointly supervised by Cedric Briens and Franco Berruti. Various drafts of the paper were reviewed by Cedric Briens, Jennifer McMillan and Franco Berruti. Jennifer McMillan provided technical advice throughout the project progress, ensuring the study is relevant to Fluid Coking and reviewed the final draft of the manuscript. F. Sanchez helped with data acquisition system setup.

### Chapter 3

**Article Title:**
Effect of a baffle on gas bubbles flow patterns and the distribution of liquid injected into gas-solid fluidized beds

**Authors:**
Majid Jahanmiri, Cedric Briens, Franco Berruti, Jennifer McMillan, Francisco J. Sanchez Careaga

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### Chapter 4

**Article Title:**
Effect of a baffle with fluxtube on gas bubbles flow patterns and distribution of liquid feed injected into gas-solid fluidized beds

**Authors:**
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Majid Jahanmiri conducted all the experimental work, analyzed the data and wrote the manuscript. This work was jointly supervised by Cedric Briens and Franco Berruti. Various drafts of the paper were reviewed by Cedric Briens, Jennifer McMillan and Franco Berruti. Jennifer McMillan provided technical advice throughout the project progress and reviewed the final draft of the manuscript.
Dedication

I would like to dedicate this thesis to the loving memory of my uncle Professor Mehdi Jahan-Miri who was my main source of inspiration to pursue my academic career after a big 16 year gap. His words and encouragement kept my hopes alive during the hard and long process of obtaining admission to this research study. His sudden death during the period of my research came as a shock to me, but remembering his words and his image as a special, true and unique human being, kept me motivated in completing my studies and research.

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Nomenclature

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\( d_B \)  
Bubble diameter (m)

\( g \)  
Gravitational acceleration (m/s\(^2\))

\( K \)  
Velocity coefficient

\( U_B \)  
Bubble rising velocity (m/s)

Chapter 2

\( B_i \)  
Undefined signal characteristic

\( C_i \)  
Undefined signal characteristic

\( D_i \)  
Undefined signal characteristic

\( F \)  
Full regression model

\( f_i \)  
Average frequency of local tribo signal (Hertz)

\( P_i \)  
Power of local tribo signal (V\(^2\))

\( q_b \)  
Bubble gas flux (m/s)

\( q_{b,i} \)  
Local bubble gas flux (m/s)

\( \overline{q_{b,c}} \)  
Calculated Cross-sectional average bubble flux (m/s)

\( \overline{q_{b,e}} \)  
Experimental Cross-sectional average bubble flux (m/s)

\( R \)  
Reduced regression model
Fluidization velocity (m/s)

Minimum fluidization velocity (m/s)

Injection nozzle horizontal penetration length into the fluidized bed (m)

Tribo rod horizontal penetration length inside the fluidized bed (m)

Vertical distance from distributor to triboelectric probe (m)

Empirical calibration constant

Empirical calibration constant

Empirical calibration constant

Signal cycle time

Agglomerate diameter (µm)

Typical injection nozzle tip, hole diameter, used in commercial fluid coker (mm)

Average frequency of local tribo signal (Hertz)

Conductance of fluidized bed (Siemens)

Power of local tribo signal (V²)

Bubble gas flux (m/s)

Local bubble gas flux (m/s)
\( \overline{q}_{b,c} \) Calculated Cross-sectional average bubble flux (m/s)

\( R_{bed} \) Fluidized bed resistance (Ω)

\( R_m \) Measurement resistance (Ω)

\( V_f \) Fluidization velocity (m/s)

\( V_m \) Applied Voltage (V)

\( V_{mf} \) Voltage drop across measurement resistor (V)

\( X_n \) Injection nozzle horizontal penetration length into the fluidized bed (m)

\( X_t \) Tribo rod horizontal penetration length inside the fluidized bed (m)

\( z \) Vertical distance from distributor to triboelectric probe (m)

\( \gamma \) Empirical calibration constant

\( \beta \) Empirical calibration constant

\( \gamma \) Empirical calibration constant

\( \tau \) Agglomerate moisture release time constant (s)

**Appendix B**

\( M_{<600\mu m} \) Total mass of bed after recovery of macro-agglomerates (g)

\( M_{dye} \) Mass of blue dye (g)

\( M_{GA} \) Mass of Gum Arabic binder (g)
<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
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<tbody>
<tr>
<td>$m_p$</td>
<td>Mass of particles trapped in agglomerates (g)</td>
</tr>
<tr>
<td>$m_R$</td>
<td>Representative sample taken from $m_{&lt;600\mu m}$ (g)</td>
</tr>
<tr>
<td>$M_s$</td>
<td>Sand mass (g)</td>
</tr>
<tr>
<td>$M_{\mu aggl,si}$</td>
<td>Mass of micro-agglomerates for an individual size cut in sample (g)</td>
</tr>
<tr>
<td>$M_{\mu aggl,i}$</td>
<td>Total mass micro-agglomerates in the bed for a given size cut (g)</td>
</tr>
<tr>
<td>$X_f$</td>
<td>Weight fraction of fines (g)</td>
</tr>
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<td>$X_{f,bed}$</td>
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Chapter 1

1 Introduction and literature review

The current research work presented in this thesis, studies the impact of a baffle to enhance the distribution of a liquid that is sprayed into a gas-solid fluidized bed. Liquid injection is the basis of several industrial processes utilizing fluidized beds with a vital goal of improving production yield and lowering agglomeration. The motivation inspiring this research work is to enhance the productivity of the present fluid coking processes.

This chapter covers a short summary of notable commercial processes involving liquid injection with concise details about Fluid Coking™ which is the main focus of this research work. The review of previous academic research on bubble characterization in a gas-solid fluidized bed follows along with the few studies of baffles in fluidized beds. Finally, the research objectives are outlined.

1.1 Introduction

Canada possesses large fields of oil sands that have proven to be a powerful boost to the Canadian economy. These oil sands are significant due to the fact that light oil reserves are diminishing worldwide [1]. Canada's oil sands are the third largest reserves of crude oil in the world, with 166 billion barrels of recoverable oil. The oil sands are located in different regions within the province of Alberta (Western Canada) and cover an area over 142,000 square kilometers. Oil production from oil sands has been conducted commercially for almost five decades. Initially oil sands were primarily accessed through large open pit mining operations. Since the mid-1980s and especially over the last decade, in situ technologies have played a growing role in oil sands production.

Oil sands are a mixture of 85% sand and clay, 5% water and 10% bitumen. Bitumen is a black substance with a high carbon to hydrogen ratio that is too heavy or viscous to flow or be pumped without being diluted or heated. The bitumen contains 50-60 weight % of vacuum residue, i.e., components that cannot be distilled, which must be converted to
distillable fractions [2]. The vacuum residue has a high boiling point of above 525 °C [3] which is due to the existence of long hydrocarbon chains in the bitumen and as a result, oil refineries are unable to process bitumen in its raw form. Therefore, this bitumen requires extensive processing and upgrading in order to produce transportation fuel and petrochemical products that can be marketed.

Conventional commercial processes employed to convert bitumen into lighter valuable hydrocarbon fractions which are considered carbon rejection processes are Delayed Coking, Flexicoking™ and Fluid Coking™ that involve the production of carbon rich coke solid, distillable liquid and light ends [2]. This research work aims at enhancing the operation and production process of Fluid Cokers and thus focuses on Fluid Coking™.

Even though Fluid Coking is a leading process for bitumen conversion in Canada, there are many issues that hinder the smooth operation of Fluid Cokers. Ideally, efficient and uniform liquid feed distribution would boost the Coker liquid yield and enhances its operability span by reducing the formation of large agglomerates [4][5]. A new study is the utilization of baffles inside Fluid Cokers connected downwardly and radially inwards from the reactor wall to improve liquid distribution and avoid fouling [6][7]. This research aims to study the impact of a baffle on the formation of large bubbles altering the bed hydrodynamics, and also its effect on the interaction of injected liquid and solid particles in a gas-solid fluidized bed.

A brief introduction to the major processes that involve injection of liquid into the reactor is provided in the following sections:

1.2 LLDPE Process

The production of linear low density polyethylene which is a hydrogenation process involves an exothermic reaction and the heat is removed by injection of a liquid which has a neutral behavior in the reaction. In a LLDPE process licensed by BP (British Petroleum), where a Ziegler-Natta catalyst is used, a fluidized bed reactor is utilized to process the polyethylene reaction at a temperature ranging between 75 °C - 85 °C and a pressure of 2 MPa. The polymerization reaction is exothermic and the heat from the reaction is removed
by cooling the hot recycle gas at the reactor outlet and compressing it back into the reactor. The reaction loop for the production of linear low density Polyethylene is shown in Figure 1.1.

![LLDPE reaction loop diagram](image)

**Figure 1.1: LLDPE reaction loop**

The issue with this process which limits the production rate, is restriction of the rate at which the reaction heat is removed using heat exchangers in the gas recycle loop. So an ideal solution to this issue that increases reactor productivity, is the injection of an inert condensable hydrocarbon with a high heat capacity, such as Pentane, which instantly removes the reaction heat by evaporation and allows for increased production rate. In order to control the reaction at a desired temperature, an optimum injection rate of pentane should be considered, since over-injection of pentane would not allow the pentane to evaporate in a timely manner and could lead to plugging of heat exchanger and reactor gas distributor.
This could cause uneven fluidization and lead to issues such as channeling, agglomeration and bogging of the bed that would eventually shut down the reactor.

An optimum jet penetration length and jet angle should be considered for each injection nozzle to achieve a uniform distribution and atomization of the liquid droplets, which helps in quick vaporization of pentane and as a result, good control of the reaction temperature. The nozzle tip geometry, nozzle penetration and elevation from the distributor are other important factors in maintaining an efficient liquid distribution. The solids flow pattern and the liquid penetration pattern in a typical LLDPE reactor are shown in Figure 1.2.

Figure 1.2: Top & side view of liquid injection pattern in a LLDPE fluidized bed reactor
1.3 Fluidized Catalytic Cracking (FCC)

Powdered catalyst types introduction lead to the development of fluidized bed catalytic cracking (FCC) in 1942 (United States) and has since been one of the most significant and widely used processes in the petroleum refineries for the production of gasoline and diesel from heavy distillates. FCC is also called the heart of a modern refinery destined for maximum production of gasoline [8].

In the FCC process, hot (500 °C) heavy hydrocarbon and dispersion steam are injected into the riser of reactor and are mixed with the hot reactivated zeolite catalyst flowing down the regenerator standpipe. The feed is vaporized by the hot catalyst and the mixture is lifted up the riser into the reactor, where the catalyst is segregated from the vapors. The cracking reactions usually start in the riser and are completed in the reactor in less than 3 seconds. The zeolite catalysts are designed to have extended surface area, creating room for extra active sites which aid in enhancing the cracking reaction. The cracking reaction being endothermic obtains its heat from the hot zeolite catalyst. A disadvantage of this process is the zeolite catalyst being poisoned by heavy hydrocarbons containing high level of impurities, which restricts the application of heavy hydrocarbons [9].

A key feature of the FCC process which impacts the product composition greatly is the liquid injection [10] which according to Newton et al. [11] should form small droplets with minimum variation in drop size, wide spray coverage along the riser flow area, and efficient droplet mixing with catalyst. Larger droplets tend to heat up slowly and lead to undesired coking reactions and possible agglomeration, whereas too small droplets are likely to overcrack and result in dry gas.

1.4 Fluid Coking™ Process

The bitumen extracted from sand oil is too viscous and contains high levels of impurities such as heavy metals, sulfur, nitrogen and oxygen and therefore needs to be upgraded before being used as the feed for Vacuum & Atmospheric units in refineries to
produce light, valuable hydrocarbons. A common upgrading process established by ExxonMobil is the Fluid Coking™ Process.

Introduction of the Fluid Coking™ process has the advantage of continuous operation, which avoids the use of alternate coke drums as in Delayed Coking process. Fluid Coking™ is a continuous pyrolysis process designed to convert heavy, non-vaporizing oils i.e. high molecular weight hydrocarbons into valuable light distillates with lower boiling points with the product called synthetic crude oil, which can be further processed in refineries. The light hydrocarbon or synthetic crude oil is considered, feed for vacuum and atmospheric columns existing in many refineries, which produce valuable products like gasoline, kerosene, diesel fuel and naphtha (feedstock for petrochemical olefin units). The process involves thermal cracking at high temperatures (500 – 550 °C). This process develops in a fluidized bed where hot Coke particles introduced in the freeboard region, come into contact with dispersed bitumen droplets, resulting in an endothermic reaction [12]. The coke particles carry the heat for reaction and solid by-products, called “new coke”, deposit on the coke surface. The Coker typically operates in a turbulent fluidized regime with the bulk of solids moving downwards and the bulk of the vapours having an upward motion, resulting in a core-annulus regime behavior with the dense downflowing solid particles surrounding the upflowing lean vapor region in the center [13].

The coking process is initiated by preheating the bitumen at 350 °C which lowers its viscosity and saves the amount of heat required, for coking and thermal cracking reactions in the Fluid Coker. Preheating the bitumen is also beneficial in enhancing its flowability through injection nozzles into the Coker. Before injection into the reactor, bitumen is diluted with atomizing steam in a mixing chamber to form a two-phase bubbly [14] fluid which is transferred to a series of injection nozzles located at different elevations on the periphery of the reactor. The aim of utilizing a series of uniformly located injection nozzles on the reactor is to enhance liquid distribution in the fluidized bed Coker and avoid over moistening of coke particles in any region of the reactor. Atomizing steam is efficient in dispersing bitumen into small droplets, once injected through the spray nozzle, which allows the bitumen scatters to reach the interior region of the reactor where they enclose
the hot coke particles at 510 – 0.550 °C [15] and a pressure of 34.5 to 103.4 kPa [1]. Steam is also used at the bottom of the Coker to attrite the large coke particles and fluidize the smaller coke particles [15]. The heat transferred from coke particles to bitumen droplets triggers the endothermic thermal cracking reaction resulting in the formation of volatile hydrocarbon vapors at the bitumen-coke border.

Figure 1.3: Simplified flow diagram of Fluid Coker (Adapted from Prociw [14])

The Fluid Coker (Figure 1.3) is divided into the following 5 sections of reactor, scrubber, stripper, fractionator and burner which are described as follows:

Bitumen is injected into the reaction zone by a series of injection nozzles, where the uniformly distributed bitumen droplets react with the fluidized hot coke particles. The resulting vapor from the reaction is carried by the fluidization gas containing entrained coke particles, into the upstream cyclones. The cyclone utilizes centrifugal force to redirect the separated solids from the vapors and gas, back to the reaction zone by gravity, the
vapors flow to the scrubbing section and the gas leaves the reactor as the product gas. The vapor droplets are further separated in the fractionator zone of the scrubber, considering their vapor pressure. The lighter vapors which do not condense in the scrubber, leave the Coker from the top section for further condensation and upgrading. The heavier vapors that condense and descend over the scrubber sheds [16], act as a coolant to the fresher vapors entering the scrubber [14].

The heavy vapors that travel down to the stripping zone, release the hydrocarbon trapped on the coke surface. The coke particles that are stripped are transferred to the burner via a pneumatic system, where they are partially combusted in the presence of air to produce heat. The hot coke in the burner is introduced back to the reactor to provide part of the heat required for the thermal cracking reaction.

1.4.1 Scrubber

The hydrocarbon vapor product combines with steam to travel upward into the lean phase (freeboard) of the reactor, where the superficial gas velocity is 1 – 2 m/s and as a result some fine coke particles also get entrained along with the vapor-steam mixture into the cyclones. The centrifugal force created by the cyclone separates the entrained fine coke particles from the mixture and redirects them to the reaction zone by gravity, through the cyclone diplegs. The vapor products exit from the top of the cyclones into the scrubbing zone, where the vapors are quenched to 370 – 400 °C as they ascend towards the inverted U or V-shaped internal sheds and come in contact with the circulating oil in this region. The light vapors travel to the top of the scrubber, while the heavy vapor is recycled to the reaction zone for further reaction. The internal sheds is the point where the descending quench oil comes in contact with the ascending hot vapors. A de-entrainment grid assembled above the sheds is responsible for further elimination of impurities like fine coke particles and quench oil from the vapor product [17].

Venturi-effect impacts the vapors passing out of the cyclone to the scrubbing zone by increasing the product vapors velocity and transporting them to wall region where the quench oil is descending, which results in the oil droplets being carried above the internal
sheds and the de-entrainment grid. The fine coke present in this area reacts with the oil to form deposits on the sheds and the grid, leading to partial blockage and fouling of the grid. This has an impact on restricting the area of vapor flow which results in low contact time with the quench oil above the scrubber and further plugging of the de-entrainment grid, resulting in poor product gas quality, containing high levels of impurities which could subsequently affect downstream units like Hydrotreating unit (catalyst poisoning) (Prociw et al., 2014) [14].

A solution provided by Bulbuc et al. ([17] to rectify this issue was to assemble baffles at the cyclone discharge to neutralize the venturi-effect on the outflowing vapors by restricting their velocity increase to create a uniform distribution of the vapors across the scrubber cross-section which results in reduction of coke deposit on the de-entrainment grid above the scrubbing zone.

1.4.2 Stripper

The stripping zone located at the bottom of the Fluid Coker is responsible for recovering the hydrocarbon vapor trapped in the heavy coke particles falling into this region, by injection of stripping steam from a sparger. This action increases the reaction yield and avoids unwanted vapor formation in the downstream burner which could lead to plugging and operation complications (Davuluri et al. 2012)[16]. The stripping section contains inverted v-shaped baffles better known as “sheds” to help distribute the steam flow uniformly and as a result, increase the contact surface between the upflowing steam and the falling coke particles.

A common issue in the stripping zone is the formation of coke deposits on the shed surface. These deposits restrict the area available for solids and gas flows, which may lead to premature shut-down. Grace et al. (2005)[13] addressed this issue by suggesting v-shaped sheds with different angles which was helpful in minimizing agglomeration and enhancing the contact surface area between the fluidizing gas and the heavy coke particles. Sanchez et al. (2013) [7] studied the effect of sheds with 3 different configurations and showed that the “Mesh-Shed” was the most effective in terms of least amount of vapors
reaching the stripping zone and having the lowest percentage of liquid lost (entering) to the burner which was an improvement compared to the normal sheds. It was also proven that with the normal sheds, an opening surface area of 60 – 70 % would be ideal to minimize valuable hydrocarbons entering the burner.

1.4.3 Reaction Zone

Based on the design of Pfeiffer et al. (1959) [18] a typical Fluid Coker’s reaction zone (coking section) is comprised of the following subzones and operating conditions:

**Table 1.1: Dimensions of the Fluid Coker’s reaction zone designed by Pfeiffer et al. [18]**

<table>
<thead>
<tr>
<th>Zone</th>
<th>Diameter (m)</th>
<th>Height (m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 Disengaging Zone (Top straight side)</td>
<td>2.7</td>
<td>6.1</td>
</tr>
<tr>
<td>2 Dense Bed (Wide Diameter)</td>
<td>3.4</td>
<td>4.9</td>
</tr>
<tr>
<td>3 Middle zone cone</td>
<td>1.2 (narrow end), 3.4 (wide end)</td>
<td>10.4</td>
</tr>
<tr>
<td>4 Stripping Zone</td>
<td>1.2</td>
<td>3</td>
</tr>
</tbody>
</table>

Today’s conventional Fluid Cokers have the same geometry as the design established by Pfeiffer et al. [18], however the exact dimensions might not be the same as in Table 1.1. In order to tackle the increase in the vapor product and maintain the fluidization velocity in the reaction zone, the Coker’s diameter is increased in a conic pattern, from top of the stripping zone to the disengagement section, above the reaction zone. A sudden decrease in the diameter of the disengagement zone is to surge the gas-solid mixture into the cyclones in the fractionation zone, and also avoid over-cracking of the hydrocarbon vapors in the reaction zone by reducing their residence time.
Table 1.2: Operating conditions of a typical Fluid Coker (Pfeiffer et al. 1959) [18]

<table>
<thead>
<tr>
<th>Operation Conditions</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>1. Temperature (°C)</td>
<td>550</td>
</tr>
<tr>
<td>2. Pressure (kPag)</td>
<td>75</td>
</tr>
<tr>
<td>3. Mass of bed solids (kg)</td>
<td>64000</td>
</tr>
<tr>
<td>4. Fluidization gas velocity (m/s)</td>
<td>0.3 (Bottom) – 1.1 (Top)</td>
</tr>
</tbody>
</table>

The reaction zone requires high temperatures of 510 - 566 °C [19] to enable thermal cracking of the long chained heavy hydrocarbons. The reaction pressure is at a relatively low value of 75 kPag to ensure the formation of vapor hydrocarbons, however the pressure changes from 34.5 kPa to 103.4 kPa [1] from top to bottom respectively. The height of solids in the fluidized bed approached 18 m which are constantly circulated by vapors and steam. A research by Song et al. [20] on hydrodynamics of Fluid Cokers, showed that the dense phase of coke solids is formed towards the wall region of the reactor and the lean phase of coke solids develops towards the center of the bed generating a core-annulus flow pattern. The solids tend to flow down along the wall region and are fluidized upward towards the center. The solids accumulated in the wall region are entrained into the spray jet, formed at the tip of injection nozzles. The effective uniform distribution of bitumen is significant in its interaction with coke particles and enhancing the product yield.

1.4.4 Burner (Furnace)

The heat required for the continuous endothermic coking reaction in the Fluid Coker is generated in the Burner. Coke particles from the reaction zone, descending through the stripper, are transferred to the furnace vessel utilizing a pneumatic transport system and are kept fluidized by air. The coke particles partially combust with air introduced to the burner to raise the temperature of the coke particle to a range of 540 – 820 °C and the vessel is operated at this temperature by controlling the air flowrate into the burner. Some of the hot coke particles are pneumatically transferred to the reaction zone of the Coker and the surplus coke in the burner is sent to storage (Worley et al.) [21]. The flux of coke particles
transferred to the Coker depends on the temperature gradient between these two vessels (Pfeiffer et al.) [18], where a greater temperature gradient would demand a higher flux of hot coke particles into the reaction zone.

### 1.5 Flexicoking™

The Flexicoking process is a carbon rejection process which has a reactor operation very similar to Fluid Coking. This process is an expansion of the Fluid Coking process where the net coke yield, after gasification of most of the coke produced in the coker, is only 2 wt. % compared to the 20 wt. % for Fluid Coking, which means that the majority of the coke that remains after gasification is utilized in heating the feed [19]. A gasifier vessel with a fluidized bed is added to the process which is used to gasify the coke by injection steam and air to produce a low BTU synthesis gas which contains hydrogen, carbon monoxide, nitrogen, and hydrogen sulphide [22].

### 1.6 Review of local bubble flux characterization in gas-solid fluidized beds

Formation of bubbles in gas-solid fluidized beds is generally beneficial in enhancing good mixing of the solid particles and developing efficient heat and mass transfer. Bubbles have a remarkable impact on the hydrodynamics of the fluidized bed which is a pivotal factor in operating fluidized beds. Variation of the local bubble flux over the fluidized bed cross section plays an important role in promoting these features. Over the past decades, several researchers have utilized various methods to study the distribution of gas bubbles across the fluidized bed cross section. Some researchers also aimed at modifying the radial distribution of gas bubbles by implementing novel methods.

A suitable and convenient method to determine the bubble gas distribution in a fluidized bed is direct visualization. Lim et al. used the principles of digital image analysis [23] to determine the bubble size distribution at varying heights and fluidization velocities. However this technique was incapable of measuring the bubble size distribution at lower heights where smaller bubbles were predominant because detection of very small bubbles,
where excessive solid interference occurs, produces problems as a consequence of the light attenuation in the emulsion phase [23]. This method is reliable for measurement of bubbles at higher bed heights where large bubbles are predominant.

Horio et al. visualized three dimensional (3D) structures of dilute suspensions in the freeboard of a bubbling fluidized bed, utilizing the laser sheet technique [24]. The eddies ejected from the bed surface into the freeboard as a result of bubble eruption, do not disappear immediately but drift upwards in the form of gas pockets which are called ghost bubbles. With this technique the eruption of bubbles in the freeboard can be visualized in an arbitrary cross section and by applying multiple laser sheets horizontally and vertically, the 3D structure of the ghost bubbles formed in the freeboard can be investigated [24].

However, although the above mentioned methods and other techniques dealing with direct visualization have the advantage of performing a non-intrusive measurement of bubble properties [25] but the issue is that their application is restricted to direct visualization of the outer section of dense three-dimensional fluidized beds which are against the wall, pseudo two-dimensional beds and very lean gas-solid beds [26]. Therefore these methods cannot be used to study bubble characteristics in the bed interior [25].

Among researchers who used x-ray to study bubble gas motion in gas-solid fluidized beds, Rowe et al. [27] were pioneers. Using this technique they were able to define a spherical shape for undisturbed bubbles with an indented base occupying about one-quarter of the sphere volume which is referred to as the bubble wake [27]. The size of wake tends to increase as the particle size decreases and as the particle shape changes from spherical to angular. They determined that the rising velocity of bubble increases with its diameter and can be defined as:

$$U_B = K \sqrt{(g \cdot d_B)/2}$$

Where K varies based on the nature of solid particles in fluidized bed from 0.8 to 1.2. They also found that with increase in bubble concentration, the frequency of coalescence increases as well. Rowe et al. enhanced their research by developing a theory to show how
bubble shape and frequency changes with height in a fluidized bed [28]. Yates et al. employed x-ray images to evaluate voidage distribution around bubbles coalescing in a gas-solid fluidized bed [29]. They found that in fluidized bed of powders in group A and B of Geldart’s classification, bubbles formed are enclosed by an expanded “shell” of gas and particles in which the voidage tends to decrease exponentially as compared to that of the emulsion phase, remote from the bubble. With a simple model of coalescence they showed that the volume size of a “visible” bubble grows with the inflow of gas from the shell region [29].

The drawback of using x-ray images to characterize bubbles are their weakness in resolving multiple, simultaneous bubbles [26]. X-ray is not designed to study multiple bubbles because it can only visualize 2D projection of 3D objects and therefore tomography should be implemented to obtain an enhanced observation of multiple bubbles [26].

Several researchers employed x-ray tomography to study bubbles in fluidized beds. Mudde used a double X-ray tomographic scanner [30] to measure solid distribution in a fluidized bed and was able to determine bubble characteristics such as size, volume and velocity, for bubbles which are greater than 2.5 cm in size. In this method the vertical dimensions of bubbles were obtained from the bubble velocity which made it possible to measure the volume of each bubble. Brouwer et al. utilized fast X-ray tomography [31] to study fluidized beds since this method was reliable in reconstructing bubble volumes and bubble velocities. Their research showed that at high pressures there is a clear decrease in the average bubble size which will typically enhance the fluidized bed performance. Their results also showed that increasing the fine content of the bed will decrease the size of bubbles [31]. Verma et al. used ultrafast electron beam X-ray tomography [32] to study the impact of parameters such as bed inlet gas velocity, initial particle bed height and bed material on the equivalent bubble size distribution, porosity distribution, bubble diameter and the bubble rise velocity. For particles like alumina and glass, their results for bubble rise velocity were in good agreement with the Hilligardt and Werther correlation [32]. With the aid of X-ray tomography, Maurer et al. [33] were able to obtain high resolution measurements of bubble hold-up in the fluidized bed with and without vertical internals. This method proves to be a reliable source in the design of bubbling fluidized bed reactors.
However, the common issue with X-ray tomography is its inability to detect small bubbles. X-ray tomography shows improvement in image reconstruction by providing better spatial resolution than electrical capacitance tomography (ECT) but the temporal resolution of images is poor [26].

Li et al. successfully utilized electrical capacitance tomography (ECT) as a non-invasive measurement technique to determine the averaged bubble rising velocity in single bubbling regime [34]. Chandrasekara et al. made development using ECT by implementing a technique to improve the spatial resolution, which resulted in capturing accurate size of gas bubbles in fluidized beds [35]. The downside of ECT application is that this method is restricted to small scale units, since in larger units the image resolution decreases by applying ECT.

Optical probes can be employed to study the hydrodynamics of gas-solid fluidized beds. Mainland et al. introduced optical probes for gas-solid fluidized beds operating at high temperatures which was able to determine bubble properties such as bubble frequency, local bubble residence time, bubble velocity, pierced length, characteristic bubble size and visible bubble flow [36]. These optical probes had the benefit of being used in applications where visual observation is not feasible [36]. Rüdisüli et al. were able to use optical probes to effectively determine changes in bubble size with gas velocity, however the technique failed to provide a clear trend for the bubble rise velocity [37].

Capacitance probes were early used by Werther and Molerus to study the spatial distribution of bubbles for determining fluidization regime transition in gas fluidized beds [38]. They were the first to employ a needle-type capacitance probe which would have less impact on the bubble flow [39]. However, in fluidized beds with a large number of smaller bubbles, the capacitance measured is lower than expected causing a fraction of the emulsion to be accounted as bubbles [25]. In an attempt to analyze bubbling and turbulent regimes in a gas-solid fluidized bed using optical and capacitance probes, Farag et al. [40] found that the size and geometry of the capacitance probes impact the free motion of bubbles causing decay in the capacitance response and thus leading to underestimation of bubble frequency. They also found that high temperatures of 150 °C, increased
measurement discrepancy between optical and capacitance probes for both bubbling and turbulent regimes.

1.7 Application of baffle in fluidized beds

In a typical fluidized bed with uniform gas distribution, bubbles tend to coalesce and migrate towards the bed center, based on Darton’s theory [41]. The distribution of gas bubbles over the bed cross-section can be modified using baffles. Jiang et al. investigated the effect of ring baffles in the performance of a circulating fluidized bed reactor [42], assembled at different heights of the riser and found improvement in the solid holdup and ozone conversion in the gas phase of the riser, while compared to the no baffle condition. They also showed that enhanced solid and radial gas mixing can be achieved by utilizing baffles. Sanchez used radioactive particle tracking to study the impact of ring baffles on the motion of wet agglomerates in a fluidized bed [7][43]. The study aimed at improving the performance of the stripper of a commercial Fluid Coker™ and it was found that the baffles reduce fouling on the sheds of the stripper section by increasing the residence time of wet agglomerates above the baffle and ultimately reducing the undesired vapors leading to fouling [7]. However, the addition of fluxtubes to these baffles curtails the baffle performance. Wyatt et al. [6] designed a ring of frusto-conical baffles with fluxtubes (down-comers), to be assembled in the periphery of the reaction section of circulating fluidized bed reactors like Fluid Cokers. They employed computational fluid dynamic (CFD) modelling to show that the presence of baffles reduced fouling of the stripper internals [6]. Modelling also showed that baffle enhances the yield of C₅⁺ hydrocarbons and lowers coke make.

Kaza utilized baffles in bubbling fluidized beds to study the impact of shape of baffles in the bed hydrodynamics. He used 3 baffles i.e., triangular, square and circular shaped baffles to determine different hydrodynamic parameters such as minimum fluidization velocity, bed expansion, pressure drop across the bed, fluctuation ratio and expansion ratio in a bubbling fluidized bed [44]. By introducing a new parameter called blockage ratio, he found that the bed hydrodynamics are a function of the blockage ratio and independent of the baffle shape [44].
The impact of baffle on gas backmixing in a two dimensional turbulent fluidized bed of FCC catalyst was studied by Zhang et al. [45]. They found that adding a louver baffle can greatly reduce solid backmixing flux across the baffle and also retain a highly efficient gas-solid contact [45].

Issangya et al. investigated the impact of horizontal baffles [46] in circulating fluidized beds on reduction or elimination of gas bypassing (Jet streaming) which is a defect in circulating fluidized beds causing solids flow and solid/gas contacting issues in the return legs. They found that for a given operating condition the ability to eliminate jet streaming depended on baffle spacing and their axial location in the CFB [46].

In another study, Yang et al. treated baffles as a perforated gas distributor [47] in a bubbling fluidized bed to simulate the hydrodynamics of the bed using Geldart A particles. Simulation results using CFD were in rational agreement with the experimental data [47].

Rossbach et al. implemented CFD to study the impact of ring baffles on gas-solid flow in circulating fluidized bed risers [48]. They found that with the best configuration, air-foil shaped baffles improves solid distribution with a 45% decrease in the solid dispersion coefficient in comparison to the case without ring baffles.

Baffles were used by Zhang et al. to analyze the control of mean residence time difference for particles with wide size distribution in fluidized beds [49]. They found that with baffles in the fluidized bed, the mean residence time difference for different particles sizes increases by adding more baffles. However adjusting the mean residence time difference for different particles is a function of both fluidization velocity and number of baffles and it would be difficult to effectively adjust the MRT difference with just one variable. Under high fluidization velocity \( \left( \frac{V_d}{U_{mf}} = 23 \right) \) in the fluidized bed with four baffles, the MRT ratio of coarse particles to fine particles reached 5.5 [49].

Bachman et al. recently used vertical baffles in a fluidized bed to study the residence time distribution of particles [50]. They were able to derive new empirical equations for different baffle configurations in the bed, as a measure for particle transport, based on the correlation for the Bodenstein number in the absence of internal baffles [51].
1.8 Research objectives

The objective of this study was to develop a measurement method to study the distribution of gas bubbles in a fluidized bed, including the impact of a baffle with and without fluxtube.

The first step was to develop a reliable, practical method to determine the local bubble flux. To resemble conditions in a commercial Fluid Coker™, our experiments demanded high fluidization velocities of 0.45 m/s. Furthermore, the planned study of an internal to concentrate and redirect gas bubbles would produce bubbles and associated shear forces typically called “gulf streams” that would damage standard optical or capacitance probes. Therefore, the use of much sturdier triboprobes, which consist of a metal rod on which surface, triboelectric current is generated by the bubble motion, was investigated. The copper triboprobe has the advantage of being affordable without any maintenance requirement. Frictional contact between two materials that exchange electrons results in the triboelectric effect. S. Soo et al. [52] showed that when a metal is inserted into a gas-solid fluidized bed, the electric current transferred from the metal to the electrical ground is caused by the particles colliding with the metal. Matsusaka et al. [53] demonstrated that triboelectric charging is influenced by several factors such as the gas properties and the physical, chemical and electrical properties of the collision surface. The intensity of the generated electric current will rely on many factors like the fluidization velocity, the bubble size and the particle size [54][55]. Tribo-electric probes have been used to detect fines in a fluidized bed [55], measure moisture content of solids in fluidized beds [56] and monitor solid flow in fluidized beds [57], in injection jets, in fluidized beds [58], and in cyclones [59].

In the second step, the impact of simulated ring baffles on liquid distribution and bubble flow patterns in a fluidized bed was investigated. The liquid distribution was characterized by two distinct methods of Conductance and Gum Arabic. The triboelectric method developed was used to associate the liquid distribution results with the bubble flow patterns in the presence of a baffle.
In the final step, the impact of adding a fluxtube to the baffle was studied. The improvement achieved in liquid distribution using a baffle and previous research studies on the addition of fluxtube to baffle [6][7], prompted this research work.

1.9 References


Chapter 2

2 Development of a triboelectric method to measure the local bubble gas flux in a fluidized bed

2.1 Abstract

Bubbles in gas-solid fluidized beds are generally beneficial as they promote solids mixing, heat transfer and mass transfer. In most fluidized beds, the local bubble gas flux varies greatly over the cross-section. This chapter presents a novel triboelectric method to measure the bubble gas distribution in a fluidized bed. A correlation was developed that relates the local bubble gas flux to the triboelectric signal generated by the impact of the gas bubbles on a triboprobe. Several signal analysis tools, such as power spectrum, cycle analysis and signal moments, were used to analyze the triboelectric signals to determine the best experimental fit for the profile of the bubble gas flux. These triboelectric signals can be used to detect fluidization regime transitions in gas-solid fluidized beds. In this study, this new triboelectric method is used to study the impact of baffles on the distribution of the gas bubbles.

2.2 Introduction

Motion of bubbles in gas-solid fluidized beds are generally beneficial as they enhance solids mixing, heat transfer and mass transfer. Fluidized beds in general, are featured with their local bubble gas flux changing greatly over the bed cross-section. Many researchers have studied the distribution of the gas bubbles over the column cross-section, using different methods. Other researchers have developed methods to modify the radial distribution of gas bubbles in fluidized beds.

Direct visualization is a convenient method to determine the bubble gas distribution in a fluidized bed. Lim et al. employed an image analysis technique [1] to determine the bubble size distribution at various heights and fluidization velocities in a fluidized gas-solid bed. Horio et al. [2] used laser sheet illumination to visualize gas pockets (ghost bubbles)
erupting from the bed surface of a bubbling gas-solid fluidized bed. However, application of these methods is restricted to direct visualization of the outer section of dense three-dimensional fluidized beds, pseudo two-dimensional beds and very lean gas-solid beds [3].

Rowe et al. were one of the pioneers in the use of x-ray images to study bubble motion in gas-solid fluidized beds [4]. They used x-ray images to study bubble splitting and coalescence. Rowe et al. expanded their research by developing a theory to show how bubble shape and frequency changes with height in a fluidized bed [5]. Yates et al. also employed x-ray images to examine bubble coalescence and the void space surrounding individual bubbles [6]. Ariyapadi et al. (2003) used a digital X-ray imaging technique [7] to show how liquid injection affects gas bubbles. The limitation of x-ray images are their inability to resolve multiple, simultaneous bubbles [3]. X-ray imaging is not designed to study multiple bubbles because it can only visualize a 2D projection of 3D objects and therefore tomography would be better suited to obtain an enhanced observation of multiple bubbles [3].

X-ray tomography was used by several researchers to study bubbles in fluidized beds. Mudde used a double X-ray tomographic scanner [8] to measure solids distribution in a fluidized bed and was able to determine bubble characteristics, of bubbles which were greater than 2.5 cm in size. Brouwer et al. also used the same technique [9] to study, bubble size changes under different pressures in the fluidized bed. Verma et al. used ultrafast electron beam X-ray tomography [10] to detect the bubble rise velocity in fluidized beds. Maurer et al. [11] also used X-ray tomography, recently, to look at bubble coalescence and hold up and bed expansion which aided in the design of fluidized bed reactors, but required an expensive X-ray tomographic scanner. The common issue with X-ray tomography is the failure to detect small bubbles. X-ray tomography shows improvement in image reconstruction by providing better spatial resolution than electrical capacitance tomography (ECT) but the temporal resolution of images is poor [3].

Li et al. used electrical capacitance tomography (ECT) as a non-invasive measurement technique to study the averaged bubble rising velocity in a gas-solid fluidized bed [12]. Chandrasekara et al. made developments using ECT by implementing a technique to
improve the spatial resolution, which resulted in enhancing bubble imaging to facilitate bubble size measurements [13]. However, ECT is only limited to small scale units and the image resolution becomes poor with larger units.

Mainland et al. used optical probes [14] to determine different bubble properties at high temperatures. These probes are used in applications where visual observation is impractical. Liu et al. employed a parallel, multifunctional, 3-fiber optical probe [15] to measure local solids concentration, particle velocity and instantaneous solid flux. Changes in bubble size with gas velocity were successfully determined using optical probes by Rüdisüli et al.[16], however the technique failed to provide a clear trend for the bubble rise velocity.

Werther and Molerus were among the first to utilize capacitance probes to study the spatial distribution of bubbles for determining fluidization regime transition in gas fluidized beds [17]. They used needle-type capacitance probes to minimize disruptions affecting the flow [18]. However in an attempt to analyze bubbling and turbulent regimes in a gas-solid fluidized bed using optical and capacitance probes, Farag et al. [19] found that the size and geometry of the capacitance probes impacted the free motion of bubbles causing decay in the capacitance response and thus leading to underestimation of bubble frequency. They also found that high temperatures of 150 °C, increased measurement discrepancy between optical and capacitance probes for both bubbling and turbulent regimes.

The triboelectric effect is the result of frictional contact between two materials that exchange electrons. S. Soo et al. [20] showed that when a metal is inserted into a gas-solid fluidized bed, the electric current transferred from the metal to the electrical ground is caused by the particles colliding with the metal. Matsusaka et al. [21] demonstrated that triboelectric charging is influenced by several factors such as the gas properties and the physical, chemical and electrical properties of the collision surface. The intensity of the generated electric current will rely on many factors such as the fluidization velocity, the bubble size and the particle size [22] [23]. Tribo-electric probes have been used to detect fines in a fluidized bed [23], measure moisture content of solids in fluidized beds [24] and
monitor solid flow in fluidized beds [25], in both the injection jets in fluidized bed [26] and in cyclones [27].

Based on Darton’s theory [28], in a typical fluidized bed with uniform gas distribution, bubbles tend to coalesce and migrate towards the bed center. The distribution of gas bubbles over the bed cross-section can be modified using baffles. Jiang et al. investigated the effect of ring baffles in the performance of a circulating fluidized bed reactor [29], assembled at different heights along the riser and found improvement in the solids holdup and ozone conversion in the gas phase of the riser, compared to the no baffle condition. They also showed that enhanced solids and radial gas mixing can be achieved by utilizing baffles. Sanchez used radioactive particle tracking to study the impact of ring baffles on the motion of wet agglomerates in a fluidized bed [30]. The study aimed at improving the performance of the stripper of a commercial Fluid Coker™ and it was found that the baffles reduce fouling on the sheds of the stripper section by increasing the residence time of wet agglomerates above the baffle and ultimately reducing the amount of undesired vapors in the region that lead to fouling. However, the addition of fluxtubes to these baffles curtails the baffle performance. Wyatt et al. [31] designed a ring of frusto-conical baffles with fluxtubes (down-comers), to be assembled in the periphery of the reaction section of circulating fluidized bed reactors like Fluid Cokers. They employed Computational fluid dynamic (CFD) modelling to show that the presence of baffles reduced fouling of the stripper internals [31]. Modelling also showed that the baffle enhances the yield of C₅⁺ hydrocarbons and lowers coke make.

The objective of this study was to develop a measurement method to study the distribution of gas bubbles in a fluidized bed, including the impact of a baffle with and without fluxtube. These experiments demanded high fluidization velocities of 0.45 m/s to resemble commercial Fluid Coker fluidization conditions. Moreover, the planned study of internals to concentrate and redirect gas bubbles would produce bubbles and associated shear forces that would damage standard optical or capacitance probes. Therefore, the use of much sturdier triboprobes, which consist of a metal rod on which surface triboelectric current is generated by the bubble motion, was, thus, investigated. The copper triboprobe has the advantage of being affordable without any maintenance requirements.
2.3 Experimental setup

A Plexiglas fluidized bed was used to perform the triboelectric experiments (Figure 2.1). The Plexiglas unit had a height of 0.92 m and a rectangular cross section of 0.5 m by 0.1 m. The rectangular baffle spanned the bed thickness (0.1 m), had an angle of 45\(^\circ\), and extended 0.16 m into the bed, blocking 33\% of the cross-sectional area (Figure 2.1). The baffle was sealed on both sides along its length with rubber gaskets to prevent bubbles from escaping through. The fluxtube (downcomer) was a vertical, aluminium cylinder with an inner diameter of 0.04 m attached to the center of the rectangular baffle.

Figure 2.1: Schematic of the fluidized bed system and the tribo signal measuring locations
The bed particles were silica sand (from Bell & Mackenzie, Hamilton, Canada) with a Sauter mean diameter of 190 μm and an apparent particle density of 2650 kg/m³. A porous gas distributor enabled uniform fluidization of the sand particles at all the gas superficial velocities used in this study, which ranged from 0.1 to 0.45 m/s. The fluidization gas was air and the entire system was operated at room temperature.

The lone triboelectric rod used in this study, shown in Figure 2.2, was a 0.61 m long copper rod covered with tygon coating making it 25 mm in diameter, with a 12 mm long naked metal needle attached to the copper rod tip which was used as the triboelectric sensor. This rod could be moved horizontally so that the probe could go from one end of the bed to the other. The horizontal planes (elevations) at which the triboprobe was inserted into the bed are indicated in Figure 2.1. The tygon coating was used to insulate the copper rod from direct contact with the fluidized silica sand in order to avoid generation of electricity over the length of the rod and to allow triboelectricity to occur only with the needle at the probe tip.

![0.61 m copper rod with tygon coating](image)

Figure 2.2: Schematic and image of the triboelectric probe

The triboelectric current that resulted from the friction of sand particles on the metal probe was converted to voltage and amplified using a multi-range amplifier from an input of 0 -
200 nA to an output of 0 - 10.4 V. A data acquisition system was used to record the raw triboelectric current (signal) for a period of 300 seconds at a sampling frequency of 1000 Hz (time frequency of 1 ms).

The experimental procedure was to move the probe in increments of 0.05 m, along the length of the rectangular column, in order to cover the whole section of the bed. These measurements were repeated at 3 different heights, shown in Figure 2.1, for 4 fluidization velocities. Therefore, in order to obtain the complete profile of the bubble gas flux for each horizontal (lateral position) plane of the fluidized bed, 36 runs were performed.

The bubble gas flux which is the volumetric bubble flux can be defined as follows:

\[
Volumetric \text{ Bubble Flux} = \frac{m^3 \text{ bubble gas flow/s}}{m^2 \text{ column cross section}} = U_b (\text{bubble velocity}) \times x_b (\text{fraction of column volume occupied by bubbles})
\]

The transparent Plexiglas bed allowed for observation of the bubble and sand movement and, during each run, a digital video camera was used to capture video of the bubble coalescence and motion in the fluidized bed.

### 2.4 Development of Signal Analysis Methods

Distinct features of the raw triboelectric signal were extracted using various signal analysis methods. These signal characteristics were used to analyze the raw triboelectric signal generated which were utilized in a correlation developed to achieve the local, time averaged, bubble gas flux. Excel solver would be used to obtain the best experimental fit for the bubble gas flux. Three signal analysis methods were implemented in this study:

1) **Moments of the signal** featuring four signal characteristics: (a) Signal Average, (b) Signal Standard Deviation, (c) Signal Absolute Deviation, (d) Signal Kurtosis
2) **Power Spectrum of a signal** which is defined as the power or energy contained at each frequency (Figure 2.3) with two key signal characteristics: (a) Power, (b) Average Frequency

3) **Cycle Analysis** presenting two major signal features: (a) Average Cycle Time, (b) Peak of V statistic. V statistic makes the detection of cycle time convenient by allowing a well-defined peak (Figure 2.4). In comparison with the power spectrum, the V statistic provides a clear cycle time whereas power is usually noisy and the peak cannot be clearly identified.

![Power Spectrum - V_f = 0.2 m/s](image)

**Figure 2.3**: Example of power spectrum analysis of a typical signal measured with baffle and fluxtube at $z = 0.4\ m$, $X_t = 0.3\ m$ and $V_f = 0.2\ m/s$
2.4.1 Single variable correlation

The raw signal was acquired at a time interval of 1 ms, with the aid of a data acquisition system and a Lab Windows program. Two typical raw signals measured at two extreme fluidization velocities of $V_f = 0.1 \text{ m/s}$ and $V_f = 0.45 \text{ m/s}$ for the experiments are compared in Figure 2.5. The signals were measured in voltage as a function of time and as it can be seen, through the value of voltage, over a period of 5 seconds, the impact of fluidization velocity is enormous (about 5-10 volts) on the generated triboelectric signal.
Figure 2.5: Example of two typical signals measured at 2 extreme fluidization velocities with baffle and fluxtube at \( z = 0.4 \text{ m} \) and \( X_t = 0.3 \text{ m} \)

In order to characterize this difference in signal strength, a general correlation was developed as follows:

\[
q_{bi} = \alpha \cdot B_i^\beta
\]  

(1)

Where \( q_{bi} \) represents the local volumetric flux of the bubble gas (with units of m/s) measured along the horizontal plane at equal intervals of 5 cm, and \( B_i \) represents the triboelectric signal characteristic used. The triboelectric current was measured at 3 different elevations of \( z = 0.08, 0.28 \) & \( 0.4 \text{ m} \) (Figure 2.1) and at 4 different fluidization velocities of \( V_f = 0.1, 0.2, 0.3 \) & \( 0.45 \text{ m/s} \). As a result, 36 values are measured at each elevation in the Plexiglas unit, which will be used to obtain the bubble gas profile for that particular elevation. The local bubble gas flux \( (q_{b,i}) \) is used to calculate the cross-sectional average.
of the bubble gas flux ($\overline{q_{b,c}}$) at each elevation and for each fluidization velocity, which is calculated using the following equation:

$$\overline{q_{b,c}} = \frac{1}{x_w} \int_0^{x_w} q_{b,i} \cdot dx$$  \hspace{1cm} (2)

where $x_w$ is the cross-sectional length of the Plexiglas bed.

The calculated cross-sectional average of the bubble flux ($\overline{q_{b,c}}$) is then compared to the experimental value of the cross-sectional average bubble flux ($\overline{q_{b,e}}$), which is determined as follows:

$$\overline{q_{b,e}} = V_f - V_{mf}$$  \hspace{1cm} (3)

Where $V_{mf}$ is the minimum fluidization velocity, which was measured as 0.04 m/s.

To achieve the best fit for the experimental values, different methods of signal analysis, already discussed, were used and the parameters obtained from these methods were applied in the above correlation (Equation 1). The Excel solver was used to obtain the values for $\alpha$ and $\beta$, which minimized the error that was calculated as follows:

$$\sum \left( \frac{q_{b,e}}{q_{b,c}} - 1 \right)^2$$  \hspace{1cm} (4)

The calculated cross-sectional average of the bubble flux, $\overline{q_{b,c}}$ was obtained from measurements at $X_t = 0.05, 0.1, 0.15, 0.2, 0.25, 0.3, 0.35, 0.4$ and $0.45$ m, where $X_t$ is the distance of the triboprobe from the column wall on the baffle side. The calculated cross-sectional average of the bubble flux, $\overline{q_{b,c}}$ was obtained at 3 vertical positions of $z = 0.08, 0.28$ and $0.4$ m from the gas distributor, for fluidization velocities of $V_t = 0.1, 0.2, 0.3$ and $0.45$ m/s, providing 12 terms in the above equation (3).
The two typical signals at two extreme fluidization velocities of $V_f = 0.1 \text{ m/s}$ and $V_f = 0.45 \text{ m/s}$, shown in Figure 2.5, are compared in Figure 2.6, using power spectrum. The power of the triboelectric signal is plotted on a logarithmic scale versus the average frequency over a range of 0 to 25 Hertz. There is a clear distinction between the signals but with the amount of existing noise captured by the power spectrum method, it is hard to identify clearly the signal peak. Consequently, instead of using the peak frequency, the average frequency and power of the signal were calculated over the 0-25 Hz range using the following equations (4) and (5):

\[
\text{Average Frequency } (f_{\text{mean}}) = \frac{\int_0^{25} P(f) \cdot f \, df}{\int_0^{25} P(f) \, df} \quad (5)
\]

\[
\text{Average Power } (P) = \frac{1}{25} \int_0^{25} P(f) \, df \quad (6)
\]

The average frequency, thus calculated, was 5.3 Hz for $V_f = 0.1 \text{ m/s}$ and 4.7 Hz for $V_f = 0.45 \text{ m/s}$. 
Figure 2.6: Example of two typical signals measured at 2 extreme fluidization velocities with baffle and fluxtube at z = 0.4 m and X_t = 0.3 m

Figure 2.7 shows that cycle analysis could be a useful tool in clearly identifying the signal peak, since it is less sensitive to signal noise. Light smoothing of the resulting curve of the $V_{\text{statistic}}$ vs. time was used to identify more clearly the cycle time $\tau_i$.

Again, using the logarithmic scale for the cycle analysis method by plotting $V_{\text{statistic}}$ versus time, by analyzing the same two typical signals, they can be clearly distinguished with the $V_{\text{max}-i}$ clearly defined as the signal peak at the corresponding cycle time, $\tau_i$ (Figure 2.7). As shown by Figure 2.7, the cycle time was 0.192 s for $V_f = 0.1$ m/s and 0.211 s for $V_f = 0.45$ m/s.
Figure 2.7: Signal Analysis comparison of signals at 2 extreme fluidization velocities with baffle and fluxtube at $z = 0.4$ m and $X_1 = 0.3$ m

The result of the triboelectric signal obtained from experiments with the baffle with fluxtube is displayed in table 2.1. It shows that the standard deviation of the triboelectric signal provides the lowest error, followed by the absolute deviation and the power.
Table 2.1: Local bubble flux analysis with single signal characteristic (experiment with fluxtube)

<table>
<thead>
<tr>
<th>Correlation</th>
<th>$q_{bi} = \alpha \cdot B_i$</th>
<th>No.</th>
<th>$B_i$</th>
<th>Error</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Standard Deviation</td>
<td>0.0799</td>
<td></td>
<td></td>
</tr>
<tr>
<td>2</td>
<td>Absolute Deviation</td>
<td>0.0814</td>
<td></td>
<td></td>
</tr>
<tr>
<td>3</td>
<td>Power</td>
<td>0.0962</td>
<td></td>
<td></td>
</tr>
<tr>
<td>4</td>
<td>Cycle Time</td>
<td>0.1274</td>
<td></td>
<td></td>
</tr>
<tr>
<td>5</td>
<td>Average</td>
<td>0.1339</td>
<td></td>
<td></td>
</tr>
<tr>
<td>6</td>
<td>Maximum of $V_{statistic}$</td>
<td>0.1417</td>
<td></td>
<td></td>
</tr>
<tr>
<td>7</td>
<td>Kurtosis</td>
<td>0.15299</td>
<td></td>
<td></td>
</tr>
<tr>
<td>8</td>
<td>Average Frequency</td>
<td>0.1736</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

The signal characteristics shown in table 2.1 were obtained from experiments with baffle containing a fluxtube. The measurements were made at lateral positions of $X_i = 0.05, 0.1, 0.15, 0.2, 0.25, 0.3, 0.35, 0.4$ and $0.45$ m, where $X_i$ is the distance of the triboprobe from the column wall on the baffle side and at fluidization velocities of $V_f = 0.1, 0.2, 0.3$ and $0.45$ m/s. The error was finally determined from the combination of measurements recorded at three vertical heights of $z = 0.08, 0.28$ and $0.4$ m from the gas distributor.

Figures 2.8, 2.9 & 2.10 show that the calculated bubble gas fluxes do not present a good match for the measured values, whichever signal parameter is used. Therefore, an adjustment should be made to the developed correlation by combining several parameters in order to achieve desirable results. The figures presenting the results for the remaining 5 signal characteristics in table 2.1, is shown in Appendix D.
Figure 2.8: Impact of Standard Deviation on calculating bubble gas flux
Figure 2.9: Impact of Absolute Deviation on calculating bubble gas flux
2.4.2 Double variable correlation

A new correlation was proposed by adding an unknown variable and coefficient to the initial developed correlation as follows:

\[ q_{bi} = \alpha B_i^\beta C_i^\gamma \]  \hspace{1cm} (7)

In this correlation, 2 signal characteristics are combined with the aim of obtaining a better fit for the measured bubble gas flux. 7 different combination were tested with the results shown in table 2.2.
Table 2.2: Local bubble flux analysis, combining two signal characteristics (experiment with fluxtube)

<table>
<thead>
<tr>
<th>No.</th>
<th>$B_i$</th>
<th>$C_i$</th>
<th>Error</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Power</td>
<td>Average Frequency</td>
<td>0.007395</td>
</tr>
<tr>
<td>2</td>
<td>Standard Deviation</td>
<td>Cycle Time</td>
<td>0.007813</td>
</tr>
<tr>
<td>3</td>
<td>Average</td>
<td>Standard Deviation</td>
<td>0.0079224</td>
</tr>
<tr>
<td>4</td>
<td>Kurtosis</td>
<td>Cycle Time</td>
<td>0.10768</td>
</tr>
<tr>
<td>5</td>
<td>Average</td>
<td>Kurtosis</td>
<td>0.11594</td>
</tr>
<tr>
<td>6</td>
<td>Average</td>
<td>Cycle Time</td>
<td>0.121641</td>
</tr>
<tr>
<td>7</td>
<td>Maximum of $V_{\text{statistic}}$</td>
<td>Cycle time</td>
<td>0.131596</td>
</tr>
</tbody>
</table>

It can be seen that the first 3 combination of signal characteristics in the correlation provide good results with low errors which proves to be a good fit for the measured bubble gas flux as shown in Figures 2.11, 2.12 and 2.13. Figures showing the results for the remaining 4 signal characteristic combination in the above table, is presented in Appendix D.
Figure 2.11: Impact of combining Power & Average Frequency on calculating bubble gas flux
Figure 2.12: Impact of combining Standard Deviation & Cycle Time on bubble gas flux
$q_{bl} = \alpha \text{Avg}_{i} \beta \text{SD}_{i}^{Y} (\text{fluxtube}) \text{(Average - Standard Deviation)}$

Figure 2.13: Impact of combining Signal Average & Standard Deviation on bubble gas flux

On achieving a good match between the measured and calculated bubble flux, these results were used to obtain the lateral profile of the bubble gas flux in the Plexiglas fluidized bed column. This was done by plotting the ratio of the local bubble flux ($q_{bl}$) to the cross-sectional average of the bubble gas flux ($\overline{q_{b,c}}$) against the distance of the triboprobe from the column wall on the baffle side ($X_t$).

Figure 2.14 shows a typical bubble profile of triboelectric measurements experiment utilizing baffle with fluxtube which was performed at $z = 0.28$ m for different fluidization velocities. The average power and frequency were used to obtain the local bubble fluxes.
To show the impact of the baffle on the bubble gas distribution, the average of the profiles obtained at the four fluidization velocities was plotted in Figure 2.15 for the different baffle configurations, for a height of $z = 0.28$ m from the grid, using the power and average frequency.

Power spectrum method has been used to plot the bubble profile for 3 different conditions (Figure 2.15) of without baffle, with baffle and baffle with fluxtube, where the profile represents the average of fluidization velocities at which the average bubble gas flux is measured. All the profiles of bubble gas flux were measured at a height of $z = 0.28$ m from the grid.
2.4.3 Triple Variable Correlation

The previous correlation used to calculate the bubble gas flux was further expanded by adding a third signal characteristic to the equation. This was done to check, further improvement for the calculated bubble gas flux, when compared to the measured values. The new proposed correlation is as follows:

\[ q_{bi} = \alpha \cdot B_i^\beta \cdot C_i^\gamma \cdot D_i^\delta \]  

(8)
With 3 variables and 4 coefficients which will be determined by the Excel solver to obtain the minimum error in matching measured bubble gas flux. The 3 variables can be a combination of the best signal characteristics from the same or different signal analysis methods already tested in a 2 variable combination. Table 2.3 displays 4 different combination of parameters sorted in the order of increase in error.

Based on the error calculated in determining the local bubble gas flux, it can be seen in table 2.3 that the first 3 combinations, demonstrate an improvement by the reduction in error in comparison to the 2 variable combinations.

Table 2.3: Local bubble flux analysis, combining three signal characteristics (experiment with fluxtube)

<table>
<thead>
<tr>
<th>No.</th>
<th>Bi</th>
<th>Ci</th>
<th>Di</th>
<th>Error</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Average</td>
<td>Standard Deviation</td>
<td>Average Frequency</td>
<td>0.004707</td>
</tr>
<tr>
<td>2</td>
<td>Average</td>
<td>Standard Deviation</td>
<td>Cycle Time</td>
<td>0.004976</td>
</tr>
<tr>
<td>3</td>
<td>Power</td>
<td>Average Frequency</td>
<td>Cycle Time</td>
<td>0.00534</td>
</tr>
<tr>
<td>4</td>
<td>Standard Deviation</td>
<td>Cycle Time</td>
<td>Average Frequency</td>
<td>0.00718</td>
</tr>
</tbody>
</table>

The graphs for the best 3 combination of signal characteristics, displaying the calculated fit for the measured values of the local bubble gas flux is shown in Figure 2.16, Figure 2.17 and Figure 2.18. Figure showing the result for the last signal characteristic combination in the above table, is presented in Appendix D.
Figure 2.16: Impact of combining Signal Average, Standard Deviation & Average Frequency on bubble gas flux
Figure 2.17: Impact of combining Signal Average, Standard Deviation & Cycle Time on bubble gas flux
Figure 2.18: Impact of combining Power, Average Frequency & Cycle Time on bubble gas flux

It is observed that, as expected, by applying 3 variables in the correlation the error does reduce as compared to the correlation with two variables. A statistical test was applied to determine whether this improvement was statistically significant.

2.4.4 Comparison of models

Kutner et al. [32] developed a statistical method, using the Fisher-Snedecor distribution, to compare two regression models by defining one model as “Full model” with a larger number of parameters and the other as “Reduced model” with a smaller number of parameters. Here, we consider the correlation with 3 variables as the Full model (F) and the 2 variables correlation as the reduced model (R). As already discussed, the errors for the Full model will always be smaller than the errors for the reduced model, the reason...
being, the more parameters that are in the model, the better one can fit the data, including measurement errors, and the smaller are the deviations around the fitted regression function [32]. The significance level ($\alpha$) value, which represents the probability of rejecting the Reduced model (R) when it is actually the better model is taken as 5%. We have to bear in mind that reducing the significance level ($\alpha$), increases the risk of accepting the reduced model when it is actually the poorer model. The values for F and R are determined by calculating the error sum of squares and if:

R > F – then Full model is the better model

F > R – then Reduced model is accepted

**Table 2.4: Statistical comparison of 3 variable correlation (Full model) and 2 variable correlation (Reduced model) in experiments with fluxtube**

<table>
<thead>
<tr>
<th>No.</th>
<th>$B_i$</th>
<th>$C_i$</th>
<th>$D_i$</th>
<th>Error</th>
<th>F</th>
<th>R</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Average</td>
<td>Standard Deviation</td>
<td>Average Frequency</td>
<td>0.004707</td>
<td>5.32</td>
<td>0.546</td>
</tr>
<tr>
<td>2</td>
<td>Average</td>
<td>Standard Deviation</td>
<td>Cycle Time</td>
<td>0.004976</td>
<td>5.32</td>
<td>0.457</td>
</tr>
<tr>
<td>3</td>
<td>Power</td>
<td>Average Frequency</td>
<td>Cycle Time</td>
<td>0.00534</td>
<td>5.32</td>
<td>2.584</td>
</tr>
<tr>
<td>4</td>
<td>Standard Deviation</td>
<td>Cycle Time</td>
<td>Average Frequency</td>
<td>0.00718</td>
<td>5.32</td>
<td>0.62</td>
</tr>
</tbody>
</table>

Table 2.4 shows the comparison of the best 4 models and based on the statistical analysis it can be observed that for all the 4 models, the reduced model (R) is easily the better model and hence the results of signal analysis shall be based on a 2 variable model (reduced) model.

A statistical comparison was also performed between the single variable correlation (equation 1) and 2 variable correlation (equation 6) to ensure that the equation 6 is also statistically the best correlation to determine the local bubble gas flux. Accordingly, the single variable model was considered the reduced model (R) and the 2 variable model was taken as the full model (F). The significance level ($\alpha$) was the same as to the previous comparison and equal to 5 %. The results for statistical comparison is shown in table 2.5:
Table 2.5: Statistical comparison of 2 variable correlation (Full model) and single variable correlation (Reduced model) in experiments with fluxtube

<table>
<thead>
<tr>
<th>No.</th>
<th>B&lt;sub&gt;i&lt;/sub&gt;</th>
<th>C&lt;sub&gt;i&lt;/sub&gt;</th>
<th>Error</th>
<th>F</th>
<th>R</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Power</td>
<td>Average Frequency</td>
<td>0.007395</td>
<td>5.12</td>
<td>88.198</td>
</tr>
<tr>
<td>2</td>
<td>Standard Deviation</td>
<td>Cycle Time</td>
<td>0.007813</td>
<td>5.12</td>
<td>81.729</td>
</tr>
<tr>
<td>3</td>
<td>Average</td>
<td>Standard Deviation</td>
<td>0.0079224</td>
<td>5.12</td>
<td>145.29</td>
</tr>
</tbody>
</table>

Based on the results achieved and the criteria determining the best statistical model, it can be concluded that the full model which is the 2 variable correlation is the best model.

2.4.5 Discussion

A statistical analysis of the results indicates that the best prediction of the local bubble gas flux requires two different characteristics of the triboelectric signal, such as the power and the average frequency, or the standard deviation and the cycle time (Table 2). Measures of the amplitude of the signal fluctuations, such as the power or the standard deviation, are not sufficient to obtain the bubble flux. This suggests the triboelectric signal is affected by the bubble flux and the size of the gas bubbles. The bubble size affects measures such as the bubble frequency or cycle time, which can thus provide information on the bubble size.

There are two major issues with any method based on triboelectric signals:

- Triboelectric signals are weak and sensitive to electrical noise. This is why, in this study, the power is preferred to the standard deviation, since most of the electrical noise can be eliminated by taking the power between 0 and 25 Hz.

- Triboelectric effects depend on the surface composition of the particles and metal probe, which can be affected by erosion, corrosion and adsorbed species, such as water molecules. Consequently, if separate calibration experiments were conducted, there would be a risk of significant changes in empirical calibration constants between the calibration experiments and the actual measurement.
experiments. This is the reason why this study uses self-calibration: the actual measurements are used to obtain the calibration constants.

To measure the bubble flux using the self-calibration method, it is important to vary both the fluidization velocity and the elevation to capture the effects on the triboelectric signals of the bubble flux and the size of the gas bubbles. Changing the fluidization velocity \(V_f\) changes the bubble flux. Since the bubble size varies with elevation \(z\), getting data at various elevations provides information on the impact of the bubble size.

### 2.5 Results

As shown in Table 2.2, the best method combined the average frequency and the power. Reasonable results were also obtained with the combination of standard deviation and cycle time, and with the combination of average and standard deviation: similar lateral profiles of the bubble gas flux were obtained with the top three combinations of Table 2, as shown in Appendix A. All the bubble gas flux results presented below were obtained by combining power and average frequency, as shown in Equation (9):

\[
q_{bi} = \alpha P_i^\beta f_i^\gamma
\]  

(9)

**Table 2.6: Power spectrum with two signal characteristics for 3 different fluidized bed configurations**

<table>
<thead>
<tr>
<th>No.</th>
<th>(q_{bi} = \alpha P_i^\beta f_i^\gamma)</th>
<th>(\alpha)</th>
<th>(\beta)</th>
<th>(\gamma)</th>
<th>Error</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Without Baffle</td>
<td>0.05983</td>
<td>0.90303</td>
<td>2.5995</td>
<td>0.006108</td>
</tr>
<tr>
<td>2</td>
<td>With Baffle</td>
<td>0.02808</td>
<td>0.8153</td>
<td>2.6732</td>
<td>0.00649</td>
</tr>
<tr>
<td>3</td>
<td>With Baffle &amp; Fluctube</td>
<td>0.04717</td>
<td>0.9001</td>
<td>2.7372</td>
<td>0.007395</td>
</tr>
</tbody>
</table>
It can be seen in table 2.6 that the error for fitting the calculated bubble gas flux with the measured value, for all the 3 different fluidized bed conditions (baffle configurations) is negligible and power spectrum proves to be a good choice for our bubble gas flux calculations and bubble profile demonstration. The self-calibrated constants, which are exponents of power and average frequency, are in good agreement for different baffle configurations. The differences in the coefficient values, and especially the $\alpha$ coefficient, can be attributed to minor changes in fluidization gas humidity, impurities in bed solids, and probe erosion/corrosion, since runs with the different baffle configurations were conducted months apart. All results below were obtained with the values shown in Table 2.6.

The bubble flux profile for 3 different fluidized bed configurations measured at $z = 0.08$ m (Figure 2.19) suggests, small gas bubbles tend to coalesce towards the center of the bed which confirms the research of Darton et al. [28] on the growth of bubbles in a fluidized bed. Even though the power spectrum method provides a good indication that coalescence occurs towards the bed center for all the 3 configurations, minor variations are noticed in the case of “Baffle” and “No baffle”, where the bubble coalescence is stronger towards the center as compared to the “Fluxtube” case.
Figure 2.19: Bubble profile with average $V_f$ at $z = 0.08$ m for 3 configurations – Power Spectrum (using power and average frequency as signal characteristics in correlation (6) to calculate bubble flux)

At $z = 0.28$ m, when no baffle was used, the lateral profile was nearly symmetrical with a peak near the center, as expected. With a baffle without fluxtube, the maximum bubble flux was shifted and was just above the baffle tip: the gas from the bubbles rising below the baffle accumulate in the gas pocket below the baffle and large bubbles are released along the baffle tip. When a fluxtube was used with the baffle, the maximum bubble flux was shifted further, as gas bubbles were released from the gas pocket not only along the baffle tip but also through the fluxtube (Figure 2.15).
Figure 2.20: Bubble profile with average $V_f$ at $z = 0.4$ m for 3 configurations — Power Spectrum (using power and average frequency as signal characteristics in correlation (6) to calculate bubble flux)

Triboelectric measurements at $z = 0.4$ m (Figure 2.20) demonstrate the impact of baffle and fluxtube in shifting the gas bubbles above the baffle and fluxtube region, whereas without baffle, the bubble gas continue to grow towards the bed center.
Figure 2.21: Map showing bubble distribution using power spectrum for the baffle with fluxtube configuration

Figure 2.21 shows the bubble distribution for the fluidized bed measured at 3 different elevations, for the baffle with fluxtube configuration. Identical values from the ratio of local bubble flux to the average bubble gas flux obtained from power spectrum analysis, have been used to plot a map, where lighter gray levels represent regions of high bubble gas flux and darker gray levels represent the emulsion phase (solids accumulation) with low bubble gas flux. Figure 2.21 shows the majority of bubbles are concentrated above the baffle and fluxtube region. The map provides a more clear perspective of the bubbles accumulation and growth inside the bed, using the power and average frequency. It is evident that the bubbles are concentrated above the fluxtube at the elevation of \( z = 0.28 \) m, and have moved above the baffle tip region at the elevation of \( z = 0.4 \) m.
2.6 Discussion

The baffle concentrates bubbles to one region of the bed. As they rise, these large bubbles carry bed particles in their wake. This likely results in solid and bubble gas circulation phenomena referred to as “gulf streaming” [33][34][35]. Other studies usually achieved strong gulf streaming through uneven distribution of gas at the bottom of the bed [36][37], providing similar results to that obtained with a baffle in this study. Strong downward solids currents in other regions of the bed pull down small bubbles that are thus diverted to the baffle region, enhancing the bubble gas flow in the baffle region and strengthening the gulf streaming currents.

Triboprobe experiments were performed in a dry bed, without any injection. There were 3 levels: a low level to show that the baffle had a negligible impact in the lower region of the bed, below the baffle; a mid-level to characterize the flow just between the baffle and the level of the nozzle; and a high level where we could compare the triboprobe results with the videos.

2.7 Conclusion

Triboproses provide a sturdy and reliable way to measure the bubble gas distribution in a fluidized bed. The local bubble flux can be determined from easily determined signal characteristics, such as the power and average frequency.

A correlation has been developed that relates the local bubble gas flux to the triboelectric signal generated by the impact of the gas bubbles on a triboprobe. Several signal analysis tools, such as power spectrum, cycle analysis and signal moments, were used in a power law providing the local bubble gas flux. A self-calibration procedure was developed to use the data obtained with the triboprobes at different heights and different fluidization velocities to determine the empirical parameters in the power law.

The triboelectric method was used to study the impact of baffles on the distribution of the gas bubbles. Because large bubbles are released from the baffle tip, the triboprobes were
able to perform measurements in conditions under which less sturdy probes, such as optical fiber probes, would have failed.

The baffles redirected gas bubbles by trapping gas bubbles in gas pockets below the baffle, from which gas escaped along the baffle edge. With the baffle with fluxtube, part of the accumulated gas also escaped through the fluxtube.

2.8 References


Chapter 3

3 Effect of a baffle on gas bubbles flow patterns and the distribution of liquid injected into gas-solid fluidized beds

3.1 Abstract

In Fluid Cokers™, heavy oil is injected into a fluidized bed of hot coke particles where it undergoes thermal cracking. Efficient and uniform liquid feed distribution enhances the yield of valuable products and the coker operability span by reducing the formation of wet agglomerates. A promising method to improve liquid distribution could be the modification of the bed hydrodynamics and mixing characteristics using baffles. The impact of a ring baffle on bed hydrodynamics and liquid distribution was studied in a small, cold flow, rectangular Plexiglas unit with a fluidized bed of sand. The baffle changes the fluidized bed hydrodynamics by redirecting gas bubbles above the baffle and towards the spray jet region, which has a beneficial effect on liquid distribution and reduces the formation of agglomerates. The experimental results show that the best liquid distribution is obtained when the tip of the liquid injection nozzle is located just above the tip of the baffle. As long as the baffle angle with the vertical is less than $45^\circ$, this will also prevent the formation of any deposit on the baffle.

3.2 Introduction

Liquid injection into a fluidized bed has been used in many industrially significant processes including fluid catalytic cracking risers (injection of heavier cuts of crude oil), fluidized polyethylene reactors (injection of pentane to absorb reaction heat), and Fluid Cokers™ (heavy oil or bitumen injection). The intrinsically good mixing and heat transfer characteristics favor liquid distribution injection and reaction efficiency. The major drawback of the mentioned industrial processes is the possible formation of wet agglomerates, as liquid trapped within agglomerates will react or evaporate much more slowly: in Fluid Cokers, for example, agglomeration needs to be minimized in order to achieve smooth reactor
operation and maximize the yield of valuable products [1][2]. Agglomeration results in incomplete reaction of the feed in the reactor zone, which leads to lower yield and fouling of the stripper zone. Ariyapadi et al. [3] took X-ray movies of radio-opaque liquid sprayed into a gas-solid fluidized bed and discovered that wet agglomerates were immediately formed at the tip of the spray jet cavity and then mixed through the bed by gas bubbles. Bruhns and Werther [4] studied liquid injection into a fluidized bed with thermocouples and inferred from their results that agglomeration occurs near the injection nozzle outlet.

It is vital to investigate new methods to curtail agglomerate formation when liquid is sprayed into a fluidized bed. One such new finding would be to place a baffle [5] that could positively influence the bed hydrodynamics and mixing characteristics in the vicinity of the injection nozzles (Figure 3.1). Limited studies has been reported in the literature on the effect of baffles in bubbling gas-solid fluidized beds. Sanchez Careaga (2013) [6] used radioactive particle tracking to study the effect of ring baffles on the motion of wet agglomerates: his results suggest that ring baffles affect bed hydrodynamics by redirecting gas bubbles above the baffle region as a result of gulf streaming [7][8][9] where a low pressure region created by baffles can create gas expansion and drag large bubbles above the baffle. Mohagheghi et al. [10][11][12] have found that the interaction of gas bubbles with the spray jet cavity can greatly affect the distribution of the sprayed liquid.
Figure 3.1: Baffle & Injection nozzle configuration in a Fluid Coker (Wyatt, Jr. et al) [5]

The research objective in this chapter is to discuss the impact of simulated ring baffles on bubble flow patterns and liquid distribution in a fluidized bed. This required the implementation of suitable methods to detect the bubble flow patterns and characterize the liquid distribution in the fluidized bed.

Several methods have been proposed in the literature to measure bubble flow patterns in fluidized beds, ranging from optical probes [13] to X-ray [14] and capacitance tomography [15] methods. A method that has been found to resist the shear forces caused by the strong and violent local bubble flows encountered near baffles is the use of sturdy tribo-probes (Jahanmiri et. al)[16]. Bed particles projected by gas bubbles on the surface of the tribo-probe cause a triboelectric current that can be measured and analyzed.
Several measurement methods have been used to determine the quality of the distribution of a liquid sprayed into a fluidized bed. Bruhns and Werther \cite{4} and Fan et. al \cite{17} measured the temperature drops at various bed locations when an evaporative liquid was sprayed. The drawback of this method is that it cannot directly provide the fraction of injected liquid that is trapped in agglomerates and the liquid concentration of the wet agglomerates. An electrical conductance technique was initially developed by Leach et al. \cite{18} and improved by Portoghese et al. (2008) \cite{19} to study the distribution of liquid into a gas-solid fluidized bed, due to the fact that presence of water would increase the fluidized bed conductance. Several researchers \cite{20} \cite{21} \cite{22} \cite{23} \cite{24} were, thus, able to determine the fraction of injected liquid that is trapped in agglomerates by monitoring the change in the electrical conductivity when water was injected into a fluidized bed of sand particles.

Morales et al. \cite{25} developed an experimental method that can simulate at 68 °C the processes that occur in the Fluid Coking reactors, when bitumen is injected in a bed of coke particles at 550 °C. This method provides information not only on the fraction of the injected liquid that is trapped within agglomerates, but also on the size and liquid concentration of the wet agglomerates. A major issue with this method is that it employs flammable solvents requiring the use of nitrogen as fluidization gas, which becomes prohibitively expensive at high fluidization velocities. A Gum Arabic method developed by Pardo \cite{26} provides similar benefits, but it can be used with compressed air as fluidization gas. Its limitation is that it requires a relatively high bed temperature of about 130 °C.

This chapter will, thus, use triboprobes to determine the impact of simulated ring baffles on bubble flow patterns. It will then adapt the Gum Arabic method to determine the impact of simulated ring baffles on liquid distribution in a fluidized bed.
3.3 Experimental Setup & Methodology

All the experiments were performed in the rectangular Plexiglas fluidized bed unit shown in Figure 3.2. The height of the unit was 0.92 m with a rectangular cross-section of 0.5 m x 0.1 m. The baffle was a rectangular slab made of aluminum with dimensions of 0.205 m x 0.1 m. The baffle was sealed on both sides along its length with rubber gaskets to prevent bubbles from escaping through. The injection nozzle was positioned 0.36 m above the gas distributor, which is 0.03 m above the end of the baffle connected to the unit wall (Figure 3.2). The bed particles were silica sand (from Bell & Mackenzie, Hamilton, Canada) with a Sauter mean diameter of 190 microns and an apparent particle density of 2650 kg/m$^3$. The Plexiglas unit had a porous gas distributor to enable uniform fluidization of the sand particles at all the gas superficial velocities used in this study, which ranged from 0.1 to 0.45 m/s. The fluidization gas was air and its flowrate was controlled and monitored with a bank of calibrated sonic orifices. The entire system was operated at room temperature.
Figure 3.2: Experimental unit showing the ideal nozzle penetration with the maximum jet length. The triboprobe orientation in the unit has been illustrated.
3.3.1 Triboelectric Method

The triboelectric method [16] was implemented to study the effect of the baffle on gas bubbles. The setup used to perform the triboelectric experiments was the same as in previous studies [27]. To have an accurate measurement of the bubble gas motion and avoid any possible false signal generation, the injection nozzle was removed from the unit. To measure the current generated by the particles hitting a metallic conductor due to bubble motion, a copper rod was used, which was 0.61 m long and 25 mm in diameter as shown in Figure 2.2 in the previous chapter. A 12 mm brass needle attached to the copper rod was used as the localized measurement probe. The copper rod was completely coated with tygon tubing to avoid generation of signals when the bubbles come in contact with the rod itself. This would allow the triboelectric current to be generated only when the bubbles come in contact with the needle triboprobe locally. The triboelectric rod was inserted into the bed at 3 different elevations above the distributor as illustrated in Figure 3.3. A plastic fitting was used at each entry port of the triboelectric rod into the bed, so that the rod could be easily moved through the sand along the cross section from one end of the bed to the other.
Figure 3.3: Orientation and lateral position of tribo rod in the fluidized bed

Triboelectric current is generated from the friction with the triboprobe of sand particles moving at high velocity in the bubble wakes. This triboelectric current was amplified and measured using a data acquisition system for a period of 300 seconds at a sampling frequency of 1000 Hz. The local, time-averaged, bubble gas flux was obtained through signal analysis of the triboelectric current [27]. The following correlation was used to calculate the local bubble gas flux:

\[ q_{bi} = \alpha \cdot P_i^\beta \cdot f_i^\gamma \]  \hspace{1cm} (10)

where \( P_i \) is the power of the tribo signal and \( f_i \) denotes the average frequency, measured over a frequency range of 0 – 25 Hz, where the bulk of the signal was generated. The cross-sectional average bubble flux was measured from the following equation:

\[ \overline{q_{b,e}} = V_f - V_{mf} \]  \hspace{1cm} (11)
where $V_f$ is the superficial gas velocity and $V_{mf}$ is the minimum fluidization velocity. The empirical constants $\alpha$, $\beta$ and $\gamma$ are obtained by matching the calculated cross-sectional average bubble flux $\left(q_{b,c}\right)$, obtained from the signal analysis data at various heights and superficial gas velocities, with the value obtained from the equation (10) [27]. The lateral profile of the bubble gas flux is obtained by plotting the ratio of the local bubble flux to the cross-sectional average bubble flux $\left(q_{b,i}/\overline{q_{b,c}}\right)$ against the penetration distance of the triboprobe inside the fluidized bed ($X_t$).

### 3.3.2 Injection Nozzle Configuration

A gum Arabic solution or water was injected into the Plexiglas fluidized bed (Figure 3.4) with nitrogen as atomization gas, to break the liquid into fine droplets which allows for enhanced and efficient jet-bed interaction.

![Figure 3.4: Injection Nozzle configuration with stable spray](image-url)
The atomization gas was injected via a 0.4 mm sonic nozzle, upstream a ¼” pre-mixer. This pre-mixer is a bilateral flow conditioner (BFC) developed by McCracken et al. [28], which was used in our experiments to efficiently mix the atomizing nitrogen gas with the pressurized liquid from the blow tank, upstream the injection nozzle. A blow tank was used to pressurize the liquid with nitrogen before its injection through a 0.61 mm reducer upstream of the BFC into the fluidized bed (Figure 3.4). The atomization gas and the blow tank had separate nitrogen sources with a specific regulator to adjust the pressure that was measured with a calibrated transducer. This was done to maintain a desired Gas-to-Liquid Ratio, known as the GLR. A typical value of 2% was used for the majority of the experiments, to simulate the atomization steam typically used in an industrial Fluid Coker™ bitumen injection system while accounting for differences in gas density and temperature [29]; the atomization steam flowrate used in Fluid Cokers represent a compromise between the atomization quality and the need to minimize steam consumption.

Figure 3.5: TEB injection nozzle geometry
The spray nozzle used in all the experiments was a TEB nozzle patented by Base et al. [30] with a tip hole diameter of 1.16 mm (Figure 3.5). The purpose of this nozzle design is to create a sudden expansion in the atomization gas within the nozzle which shatters the liquid into minute droplets and propels the mixture towards the center of the fluidized bed. An increase in the GLR would enhance the droplet dispersion at the nozzle tip. The tip hole size of 1.16 mm was chosen to achieve a jet penetration depth of 29 cm [31] which is the downscaled version of the spray nozzle (d_h = 12.7 mm) used in industrial Fluid Coker™ operating with a mass liquid flowrate of 23 USGPM (1500 g/s). The primary goal of choosing 29 cm as the jet penetration depth was to avoid the jet impacting the opposite wall while allowing the nozzle to be penetrated up to 19 cm inside the fluidized bed. The atomization nitrogen pressure associated with a GLR of 2 %, was 150 psig (1034 kPa) yielding a flowrate of 0.25 g/s and the blow tank was pressurized up to 235 psig (1620 kPa) to achieve a liquid flowrate of 12.42 g/s, corresponding to the industrial TEB injection nozzle operating at a liquid flowrate of 1500 g/s. In experiments with GLR = 1.5%, the nitrogen gas flowrate was reduced to 0.19 g/s.

Before running the experiments, open air runs were performed with the injection nozzle to check for consistent and stable nozzle sprays. With the initial nozzle configuration, which used a pre-mixer with a 30° Tee, the spray was pulsating constantly, which was undesired in terms of achieving a uniform liquid distribution in the fluidized bed. With the new nozzle configuration, the size of the bilateral flow conditioner (BFC) (pre-mixer), the restriction downstream of the blow tank and the nozzle tip diameter were reduced and as a result, a more stable spray was attained. The new nozzle configuration yielding a stable jet is shown in Figure 3.4. Portoghese et al. [29] found that, geometry of pre-mixers with a 90° Tee, as shown in Figure 3.6, provides good mixing and the most stable jet: this was verified, in this study, with the video analysis of the spray in open air (Appendix C) [32].
Figure 3.6: Pre-mixer with 2 geometry
3.3.3 Conductance Technique

For this method, a series of conductance plates attached to the Plexiglas wall was used to measure the electrical conductance of the fluidized bed at different heights, as shown in Figure 3.7.

Figure 3.7: Plexiglas unit with the electrodes for “Conductance” method
Each conductance plate was connected to the signal generator that provided an AC sinusoidal, 100 Hz current with a constant root mean square (RMS) voltage of 7.8 V ($V_{app}$) as shown in Figure 3.8. The free moisture that is not trapped within wet agglomerates can be determined from the bed conductance [19][21][22]. Liquid in the fluidized bed is in 3 forms: trapped in agglomerates, free moisture and vapour [23]. Farkhondehkavaki et al. [22] discovered that only the free moisture within the fluidized bed or released from the breakup of agglomerates can be detected by electrodes and therefore the liquid trapped in agglomerates goes undetected.

![Figure 3.8: Plexiglas unit circuit to measure conductance](image)

To perform the Conductance experiments, the Plexiglas bed was loaded with 55 Kg of silica sand and the blow tank was filled with 100 g of water. For these preliminary experiments, the bed of sand was then fluidized using air at a superficial velocity of 0.15 m/s. The liquid was injected for 8 s and the subsequent variation in fluidized bed conductance was measured.
Using a method developed by Prociw et al. [24], the variation with time of the measured bed conductance, following a pulse injection of liquid, was used to calculate the agglomerate breakup constant, \( \tau \), defined as the time required for 62.5% of the moisture initially trapped in agglomerates to be released as free moisture in the fluidized bed. In order to obtain the agglomerate breakup time constant, 4 replicate runs were required for each condition (with and without baffle).

### 3.3.4 Gum Arabic Method

The Gum Arabic method developed by Pardo [26] was adapted so that experiments could be performed at room temperature, by combining it with a procedure initially used by House [33]. A Gum Arabic solution was injected into the fluidized bed and the bed instantly defluidized. The bed was then dried at a velocity below the minimum fluidization velocity \( (V_{mf} = 0.04 \text{ m/s}) \), to dry the wet agglomerates without disruption; this allowed the agglomerates to slowly solidify, without changing the properties of the agglomerates that were immediately formed after the injection.

The whole bed was sieved to obtain the size distribution and mass of the agglomerates. To determine the dye concentration and amount of liquid initially trapped in the agglomerates, they were dissolved in a water solution. The binder (GA) solution used in all experiments had a composition of 6 wt. % of Gum Arabic (binder) and 2 wt. % of blue dye dissolved in water. To obtain agglomerates with features similar to those produced in the Fluid Coker, concentrated hydrochloric acid was added to the binder solution to adjust the pH, lowering it from 7 to 3 and reduce the solution viscosity to the same value as the bitumen injected in Cokers [26].

The same Plexiglas unit employed for tribo-electric and conductance experiments was used for the Gum Arabic experiments. In order to achieve higher fluidization velocities, the mass of solids used in the fluidized bed was lowered to 40 kg to prevent overpressure of the Plexiglas unit. The total mass of injected liquid was reduced accordingly to 75 g to avoid bed bogging, which is defined as: poor fluidization or defluidization of the bed due to high liquid to solid concentration [34][35].
During each experiment, the bed was initially fluidized at a velocity of 0.3 or 0.45 m/s. The prepared GA (binder) solution was injected for a period of 6 seconds and, immediately after the injection, the fluidization velocity was lowered below the minimum fluidization velocity to slump the bed and leave it to dry under ambient conditions until the following day. The dried bed was emptied completely and the agglomerates were recovered by sieving. The size distribution of the agglomerates, and the amount of water trapped in the agglomerates for each size cut, were measured with the methods developed by Pardo [26]. More details maybe found in Appendix B.

3.4 Results

3.4.1 Deposit formation on baffle

In order to choose a baffle angle for the experiments, a slide angle test was run with solids of variable moisture (water) concentration ranging from 0 – 20 wt. %. The plate used in the test was made of the same aluminum as used for the baffle. Based on the results obtained, the slide angle with horizontal is 30° for dry sand and varies from 33° to 70° for wet sand, depending on the wet sand moisture concentration.
Figure 3.9: Effect of the moisture content on the slide angle of wet sand

Figure 3.9 shows that, as expected, the slide angle increases with increasing moisture. Although wet agglomerates with moistures as high as 20 % or even 25 % have been obtained near the spray tip (Appendix B, Figure B-3), a baffle angle of 70° would be difficult to accommodate in a commercial, full-scale Coker. A baffle angle of 45° was, therefore selected: it was expected that even agglomerates with very high initial moisture would have the moisture of their surface layers, which contact the baffle surface, reduced to below 10% by the time they reached the baffle surface.
Wet agglomerate deposits were formed on the baffle when the injection nozzle was positioned 0.02 m inside the bed (Figure 3.10), due to the stagnant region caused by the lack of bubble gas movement above the baffle, required to fluidize the wet sand formed instantly after liquid injection. This issue was not observed at nozzle penetrations of $X_n = 0.1$ m and above, inside the bed (Fig. 3.11).
3.4.2 Effect of baffle on the lateral profile of the bubble gas flux

The triboelectric method was initially used to study the profile of the bubble gas flux with and without a baffle in the fluidized bed. Based on these experiments, the ratio of the local bubble gas flux to the cross-sectional average of the bubble flux, called dimensionless bubble flux, was plotted against the distance of the probe from the wall. Figure 3.12 shows the lateral profile at three elevations from the distributor of \( z = 0.08, 0.28 \) and \( 0.4 \) m in the absence of a baffle. The height of \( z = 0.08 \) m was selected to determine whether the baffle affected the bubble distribution below the baffle. The height of \( z = 0.28 \) m was selected to measure the lateral distribution of the bubbles just below the level of the jet cavity. Finally, measurements were performed at \( z = 0.4 \) m to compare the triboprobe results with the video observations. These measurements were performed without the spray jet.
In the absence of a baffle, Figure 3.12 shows that the profile is symmetrical and typical of fluidized beds: bubble coalescence results in the bubble gas flux peaking in the bed central region for all three heights. This corresponds to what has been observed by Darton [36] and other researchers [15] [37] [38] [39].

Figure 3.12: Lateral profile of the dimensionless bubble gas flowrate at different elevations from the grid (average of measurements at fluidization velocities ranging from 0.1 to 0.45 m/s)
Snapshots taken through the Plexiglas wall show that bubbles concentrate near the bed center in the absence of baffle, as shown by Figure 3.13.

Figure 3.13: Snapshots of bubble coalescence towards the bed center in the absence of baffle
The baffle affects the bubble distribution, resulting in profiles that are no longer symmetrical. Figure 3.14 shows that, with the baffle, at a height of 0.4 m above the distributor, i.e. well above the baffle, the bubbles have migrated towards regions above the baffle. This results from the accumulation of bubble gas in the gas pocket that forms below the baffle and from which gas escapes along the baffle edge. Because the lateral profile of the dimensionless bubble flux did not greatly change with the fluidization velocity, Figure 3.14 shows the average for the 8 fluidization velocities (0.1, 0.15, 0.2, 0.25, 0.3, 0.35, 0.4 and 0.45 m/s), at the height of 0.4 m above the distributor. For a confirmation of these findings, a video of each run was taken from outside the transparent Plexiglas bed.

**Figure 3.14: Impact of baffle on lateral profile of bubble gas flowrate 0.4 m above grid (z = 0.4 m)**
Figure 3.15 shows the picture of large bubbles formed above the baffle region, captured for a typical run at $V_f = 0.45$ m/s.

Figure 3.15: Pictures confirming bubble motion above the baffle region
Figure 3.16 shows that, as expected, the baffle did not have much of an impact on the lateral profile of the bubble gas flux at locations well below the baffle (0.08 m above the gas distributor in Figure 3.16) within the entire range of gas superficial velocities used.

Figure 3.16: Impact of baffle on the lateral profile of the bubble gas flowrate 0.08 m above the grid (z = 0.08 m) – (Average of measurements at 4 fluidization velocities)
Figure 3.17 shows the lateral profile of the dimensionless bubble gas flowrate for $z = 0.28$ m, i.e. 0.11 m above the height of the baffle tip (see also Figure 3.3). This figure shows that the baffle completely changed the flow pattern of the gas bubbles. The maximum bubble gas flux is found just above the baffle tip: this confirms that gas bubbles escape from the gas pocket below the baffle, along the baffle tip.

Figure 3.17: Impact of baffle on bubble gas flowrate, 0.28 m above the distributor and the probe laterally facing the baffle ($z = 0.28$ m) – (Average of measurements at 8 fluidization velocities)
3.4.3 Impact of baffle on liquid distribution (Conductance Method)

Figure 3.18, which uses the conductance technique described in section 3.3.3., shows that the baffle greatly improved the liquid distribution. With the baffle, weaker agglomerates were formed and the time constant of agglomerate breakage was reduced by 43%. It was observed that under identical conditions without a baffle, a stronger conductance signal was generated and consequently a longer time is consumed for the bed to dry. For these experiments, the fluidization velocity was 0.15 m/s and the tip of the nozzle was located just above the baffle tip at $X_n = 0.16$ m (Figure 3.19 c).

![Figure 3.18: Comparison of the effect of baffle on conductance and agglomerate breakup](image)

$X_n = 0.16$ m, $V_f = 0.15$ m/s

$\tau = 154$ s

$\tau = 88$ s
Figure 3.19: Nozzle penetration orientation related to baffle for Gum Arabic method
3.4.4 Impact of nozzle penetration depth on liquid distribution
(Gum Arabic Method)

Results from the gum Arabic experiments show that the baffle greatly enhanced liquid distribution into a fluidized bed, with varying nozzle penetrations at high fluidization velocities. In analyzing the resulting agglomerates from several runs, the impact of baffle is evident from the reduction in the total mass of agglomerates (Figure 3.20). The maximum relative reduction in the mass of agglomerates is reached when the nozzle tip is located just above the baffle tip, and reaches 62 % and 79 %, respectively at $V_f = 0.3 \, \text{m/s}$ and $V_f = 0.45 \, \text{m/s}$ (Figure 3.20). Figure 3.21 further asserts the significance of baffle by showing similar results for the liquid trapped in agglomerates, with relative reductions 51 % and 58 %, respectively at $V_f = 0.3 \, \text{m/s}$ and $V_f = 0.45 \, \text{m/s}$. Figures 3.20 and 3.21 indicate that the optimum lateral position of the nozzle is above the baffle tip ($X_n = 0.16 \, \text{m}$).

![Figure 3.20: Impact of baffle on relative reduction in total mass of wet agglomerates compared to “No baffle” at $V_f = 0.3 \, \text{m/s}$ & $0.45 \, \text{m/s}$](image-url)
Figure 3.21: Impact of baffle on relative reduction in total amount of liquid trapped in agglomerates compared to “No baffle” at $V_f = 0.3 \text{ m/s}$ and $0.45 \text{ m/s}$

Figure 3.22 and Figure 3.23 show that the baffle reduced the liquid content of all agglomerate sizes for $V_f = 0.3 \text{ m/s}$ and $0.45 \text{ m/s}$, respectively. They also confirm that the optimum lateral position of the nozzle is above the baffle tip ($X_n = 0.16 \text{ m}$). Nozzle penetration beyond $X_n = 0.19 \text{ m}$ was not tested due to the limitation of the jet hitting the opposite wall of the bed.
Figure 3.22: Effect of Baffle and fluidization velocity on liquid trapped in agglomerates for different nozzle penetrations at $V_f = 0.3$ m/s & GLR = 2%. All data is with baffle unless otherwise indicated.
Figure 3.23: Effect of Baffle and fluidization velocity on liquid trapped in agglomerates for different nozzle penetrations at $V_f = 0.45$ m/s & GLR = 2%. All data is with baffle unless otherwise indicated.
Figure 3.24 shows that increasing the fluidization velocity by 50%, decreases the fraction of injected water that is trapped in agglomerates for all agglomerate sizes.
Figure 3.25: Reduction in percentage of liquid trapped in agglomerates by increasing fluidization velocity from 0.3 m/s to 0.45 m/s

Figure 3.25 shows that, for all agglomerate size cuts, the reduction in water trapped in agglomerates that is achieved by increasing the fluidization velocity from 0.3 to 0.45 m/s, is larger when a baffle is present. The beneficial effect of the baffle is more pronounced for the agglomerates larger than 425 μm, which are more likely to create problems in commercial Fluid Cokers.

With the baffle, the reduction in water trapped in agglomerates that is achieved by increasing the fluidization velocity from 0.3 to 0.45 m/s, is nearly independent of the agglomerate size, with the exception of the smallest agglomerates (smaller than 425 μm). In contrast, without the baffle, the reduction in water trapped in agglomerates that is achieved by increasing the fluidization velocity from 0.3 to 0.45 m/s, mostly decreases with increasing agglomerate size, as shown in Figure 3.25.
With a superficial velocity of 0.3 m/s elsewhere in the bed, the fluidization velocity in the space between the tip of the baffle and the opposite wall is 0.45 m/s (Figure 3.26). Figure 3.27 shows that liquid distribution is, then, better than with a fluidization velocity of 0.45 m/s everywhere in the bed without a baffle, especially for larger agglomerates. This shows that the impact of baffle is not only to increase the gas velocity in the baffle region, but mostly to change the fluidized bed hydrodynamics by concentrating larger bubbles above the baffle, i.e. the region where liquid is injected at $X_n = 0.16$ m.

**Figure 3.26: Change of fluidization velocity at the baffle tip**

With a superficial velocity of 0.3 m/s elsewhere in the bed, the fluidization velocity in the space between the tip of the baffle and the opposite wall is 0.45 m/s (Figure 3.26). Figure 3.27 shows that liquid distribution is, then, better than with a fluidization velocity of 0.45 m/s everywhere in the bed without a baffle, especially for larger agglomerates. This shows that the impact of baffle is not only to increase the gas velocity in the baffle region, but mostly to change the fluidized bed hydrodynamics by concentrating larger bubbles above the baffle, i.e. the region where liquid is injected at $X_n = 0.16$ m.
Figure 3.27: Impact of baffle on cumulative reduction of water trapped in agglomerates
3.4.5 Impact of GLR reduction on liquid distribution

The effect of lower GLR on liquid distribution into the fluidized bed was assessed and it was observed (Figure 3.28) that at GLR = 1.5 %, the fraction of water trapped in agglomerates in the presence of the baffle is much lower than without the baffle at GLR = 2 %, but not as good as with the baffle and GLR = 2 %. This suggests that, in Fluid Cokers, it might be possible to reduce the required flowrate of atomization steam by modifying the bed hydrodynamics. The fluidization velocity for these experiments was $V_f = 0.45$ m/s.

![Figure 3.28: Impact of GLR and Baffle on cumulative reduction of water trapped in agglomerates at $V_f = 0.45$ m/s (the GLR is the ratio of the atomization gas mass flowrate to the mass flowrate of injected liquid).](image-url)
3.5 Discussion

The baffle used in the fluidized bed greatly enhanced the distribution of liquid sprayed into the bed. The results suggest that the baffle had two complementary effects on liquid distribution:

1) The increased turbulence in the region between baffle tip and opposite wall has a high shear effect on the wet agglomerates that are formed when liquid is injected into a fluidized bed. Most of the injected liquid is initially trapped in wet agglomerates [3][40]. Over the range of fluidization velocities used in Fluid Cokers and in this study, agglomerate breakup, results from fragmentation resulting from the shear forces induced by gas bubbles in the fluidized bed [41][42]. This was verified with agglomerates of coke and bitumen at both reacting and non-reacting temperatures [43]. Past studies have shown that fragmentation is greatly increased when increasing the fluidization velocity [41][42][43] [44].

2) The baffle redirected the gas bubbles to the right location on the jet cavity. This is confirmed by the great effect of the position of the nozzle tip relative to the baffle tip on the beneficial impact of the baffle on liquid distribution, as shown in this study. Mohagheghi [9] showed that adding additional gas at the right location below the jet cavity greatly improved liquid distribution. Mohagheghi developed a model [12] that shows that adding extra gas to the jet cavity accelerates the expansion-contraction cycle of the jet cavity, mixing the injected liquid with a larger mass of bed particles and resulting in drier agglomerates that break up more easily [41][42][45][46].

3.6 Conclusions

Several conclusions were drawn from this study:

- A ring baffle completely changes how bubbles rise through the bed. Instead of concentrating at the center, bubbles are first concentrated at the tip of the baffle and this flow pattern is retained in regions well above the baffle.
• The baffle can greatly improve the distribution of liquid sprayed into the fluidized bed. This was confirmed with two independent measurement methods: a Conductance method that measures the breakage rate of wet agglomerates and a Gum Arabic method that provides the mass and liquid content of agglomerates.

• The ideal injection nozzle position is so that its tip is located above the tip of the baffle \(X_n = 0.16\) m to achieve an optimal liquid distribution and avoid deposit formation on the baffle.

3.7 References


[32] M. Jahanmiri, Use of a baffle to enhance the distribution of a liquid sprayed into a gas-solid fluidized bed. Master’s diss., Western University (Canada), 2017.


Chapter 4

4 Effect of a baffle with fluxtube on gas bubbles flow patterns and distribution of liquid feed injected into gas-solid fluidized beds

4.1 Abstract

Bitumen is injected into the reaction section of a Fluid Coker™ where it interacts with a fluidized bed of hot coke particles and undergoes thermal cracking to form hydrocarbon vapors. Efficient and uniform dispersion and atomization of the liquid feed enhances the yield of valuable products and the coker operability by reducing the formation of wet agglomerates. A fitting method would be to utilize baffles with fluxtubes to improve liquid distribution in the fluidized bed. The effect of a ring baffle with fluxtube on bed hydrodynamics and liquid distribution was studied in a small, cold flow, rectangular Plexiglas unit with a fluidized bed of sand. The baffle with fluxtube influences the fluidized bed hydrodynamics by redirecting gas bubbles above the baffle and fluxtube region directed towards the spray jet. This has a valuable impact by improving the initial liquid distribution upon injection, which results in reduced agglomerate formation in the fluidized bed. Experimental results show that the ideal liquid injection nozzle position to achieve an optimal liquid distribution is such that its tip is located just above the tip of the baffle with fluxtube. The injection nozzle must penetrate past a ¼ of the baffle width to avoid any potential wet solid deposit formation on the baffle.

4.2 Introduction

Many notable industrial processes inject liquid into a gas-solid fluidized bed which includes; fluid catalytic cracking risers (injection of heavier cuts of crude oil), fluidized polyethylene reactors (injection of pentane to absorb reaction heat), and Fluid Cokers™ (bitumen injection). The innate nature of fluidized beds favours good mixing and heat transfer, which enhances liquid distribution and reaction efficiency. A considerable
downside of the mentioned industrial processes is the possible formation of wet agglomerates, which needs to be minimized in order to achieve reliable reactor operation and maximize the yield of valuable products. The liquid trapped in agglomerates is slow to evaporate or react and hence, growth and accumulation of agglomerates leads to fouling, operation disruptions and eventually shuts down the reactor. For example, in a Fluid Coker™, to attain a reliable operation and enhance the yield of light, valuable hydrocarbons, agglomeration needs to be minimized [1][2]. In the presence of agglomerates, the feed does not react completely with coke particles in the reaction zone, reducing reaction yield and resulting in stripper zone fouling. X-ray movies of radio-opaque liquid sprayed into a gas-solid fluidized bed were analyzed by Ariyapadi et al. [3] and they discovered that wet agglomerates were immediately formed at the tip of the spray jet cavity and then mixed through the bed by gas bubbles. Bruhns and Werther [4] studied liquid injection into a fluidized bed with thermocouples and concluded from their results that agglomerates are formed near the injection nozzle outlet.

Investigation of new methods to restrict agglomerate formation when liquid feed is sprayed into a fluidized bed, is essential. One method would be to place a baffle with a fluxtube [5] into the fluidized bed that could positively impact the bed hydrodynamics and mixing characteristics in the vicinity of the injection nozzles (Figure 3.1). Few studies have been reported in the literature on the effect of baffles in fluidized beds. Jiang et al. investigated the effect of ring baffles on the performance of a circulating fluidized bed reactor [6], which were assembled at different heights of the riser. They found that solids holdup and ozone conversion in the gas phase of the riser improved, while compared to the no baffle condition. They were also able to show that utilizing baffles can result in enhanced solid and radial gas mixing. Only recently Sanchez Careaga (2013) [7] used radioactive particle tracking to study the effect of ring baffles and baffles with downcomers (fluxtubes) on the motion of wet agglomerates in a circulating fluidized bed. His results suggest that ring baffles redirect gas bubbles and affect the bed hydrodynamic behaviour. Mohagheghi et al. [8][9][10] have found that the interaction of gas bubbles with the spray jet cavity can greatly affect the distribution of the sprayed liquid.
In the study conducted in the previous chapter the author showed that a ring baffle, without a fluxtube, can greatly improve liquid distribution into the fluidized bed by changing the bubble rise pattern and modifying the bed hydrodynamics. They found that the injection nozzle should be positioned in a way that its tip is located above the baffle tip, to achieve the optimum liquid distribution.

In a patent work, Wyatt et al. [5] used CFD modelling to show that conical ring baffles reduce the possibility of stripper fouling, based on results confirming reduction of water concentration in solids, over the stripper cross section. Similar CFD modelling results revealed that the application of fluxtubes (downcomers) enhance the performance of a baffle [5]. These results were confirmed by a model developed to evaluate the yield of C5+ liquid and coke. These findings lead to the idea of experimenting the impact of a baffle with a fluxtube on liquid distribution in a fluidized bed.

Several methods have been proposed in the literature to measure bubble flow patterns in fluidized beds, ranging from optical probes [12] to X-ray [13] and capacitance tomography [14] methods. A method that has been found to resist the shear forces caused by the strong and violent local bubble flows encountered near baffles with fluxtubes is the use of sturdy tribo-probes [11][15]. Bed particles projected by gas bubbles on the surface of the tribo-probe cause a triboelectric current that can be measured and analyzed.

Several measurement methods have been used to determine the quality of the distribution of a liquid sprayed into a fluidized bed [4] [16] [17][18][19][20][21][22]. A detailed review of these methods may be found in [23]. Pardo [24] developed an experimental method that can simulate at low temperature (130 °C) the processes that occur in the Fluid Coking reactors, when bitumen is injected in a bed of coke particles at 550 °C. This method provides information not only on the fraction of the injected liquid that is trapped within agglomerates, but also on the size and liquid concentration of the wet agglomerates.

The objective of this study is to investigate the impact of simulated ring baffles with fluxtubes on bubble flow patterns and liquid distribution in a fluidized bed. Implementation of appropriate methods was needed to detect the bubble flow patterns and characterize the liquid distribution.
4.3 Experimental Setup & Methodology

All experiments were conducted in the rectangular Plexiglas fluidized bed unit shown in Figure 4.1. The height of the unit was 0.92 m with a rectangular cross-section of 0.5 m x
0.1 m. The baffle was a rectangular slab made of aluminum with dimensions of 0.205 m x 0.1 m. The baffle is sealed on both sides along its length with rubber gaskets to prevent bubbles from escaping through. The fluxtube (downcomer) was an aluminum cylinder with a diameter of 0.04 m attached from the bottom to the center of the rectangular slab. The injection nozzle was positioned 0.36 m above the gas distributor, which is 0.03 m above the end of the baffle connected to the unit wall (Figure 4.1).

The bed particles were silica sand (from Bell & Mackenzie, Hamilton, Canada) with a Sauter mean diameter of 190 microns and an apparent particle density of 2650 kg/m$^3$. The bed mass was 40 kg. The Plexiglas unit had a porous gas distributor to enable uniform fluidization of the sand particles at all the gas superficial velocities used in this study ranging from 0.1 to 0.45 m/s. The fluidization gas was air and the entire system was operated at room temperature.

The triboelectric method was implemented to study the effect of the baffle with fluxtube on gas bubbles [15]. The setup used to perform the triboelectric experiments was the same as in previous studies [Chapter 3 - Section 3.3.1]. To measure the current generated by the particles hitting a metallic conductor due to bubble motion, a copper rod was used, which was 0.61 m long and 25 mm in diameter as shown in Figure 2.2 (Chapter 2). A 12 mm brass needle attached to the copper rod was utilized as the localized measurement probe. The triboelectric rod was inserted into the bed at 3 different elevations above the distributor as illustrated in Figure 3.2, and could be easily moved along the cross section from one end of the bed to the other.

A gum Arabic solution was injected into the Plexiglas fluidized bed (Figure 4.1) with nitrogen as atomization gas. The objective was to break the liquid into droplets and increase their surface area to allow for enhanced and efficient jet bed interaction. The atomization gas was injected via a 0.4 mm sonic nozzle, upstream a ¼” pre-mixer [25]. A blow tank was used to pressurize the liquid with nitrogen before its injection through a 0.61 mm reducer upstream of the BFC into the fluidized bed (Figure 3.4 – Section 3.3.2). The atomization gas and the blow tank had separate nitrogen sources with a specific regulator to adjust the pressure that was measured with a calibrated transducer. This was done to
maintain a desired Gas-to-Liquid ratio, known as the GLR. A typical value of 2% was used for the majority of the experiments, to simulate the atomization steam typically used in an industrial Fluid Coker™ bitumen injection system while accounting for differences in gas density and temperature [26]; the atomization steam flowrate used in Fluid Cokers represent a compromise between the atomization quality and the need to minimize steam consumption.

The spray nozzle used in all the experiments was a TEB nozzle patented by Base et al. [27] with a tip hole diameter of 1.16 mm (Figure 3.5). The purpose of this nozzle design is to create a sudden expansion in the atomization gas within the nozzle that shatters the liquid into minute droplets. The tip hole size of 1.16 mm was chosen to achieve a jet penetration depth of 29 cm [28], and thus avoid the jet impacting the opposite wall while allowing the nozzle to penetrate up to 19 cm inside the fluidized bed. The atomization nitrogen pressure associated with a GLR of 2% was 150 psig (1034 kPa) yielding a gas flowrate of 0.25 g/s and the blow tank was pressurized up to 235 psig (1620 kPa) to achieve a liquid flowrate of 12.42 g/s through the injection nozzle, corresponding to the commercial TEB injection nozzle operating at a liquid flowrate of 1500 g/s. In experiments with GLR = 1.5%, the nitrogen gas flowrate was reduced to 0.19 g/s. Before running the experiments, open air runs were performed with the injection nozzle to ensure a consistent and stable spray (Section 3.3.2 & Appendix C). To perform experiments at room temperature, the Gum Arabic method developed by Pardo [24] was adapted, by combining it with a procedure initially used by House [29]. After injecting a Gum Arabic solution into the fluidized bed, the bed was instantly defluidized, allowing the bed to dry at a velocity below the minimum fluidization velocity \( U_{mf} = 0.04 \text{ m/s} \), to dry the wet agglomerates without disruption; this allowed the agglomerates to slowly solidify, without changing the properties of the agglomerates that were immediately formed after the injection.

The bed was completely sieved to obtain the size distribution and mass of the agglomerates that are formed immediately after injection. The binder (GA) solution used in all experiments had a composition of 6 wt. % of Gum Arabic (binder) and 2 wt. % of blue dye dissolved in water. To obtain agglomerates with features similar to those produced in the Fluid Coker, concentrated hydrochloric acid was added to the binder solution to adjust the
pH, lowering it from 7 to 3 and reduce the solution viscosity to the same value as the bitumen injected in Cokers [24]. The total mass of injected liquid was 75 g to avoid bed bogging, which took about 6 seconds to be injected into the fluidized bed. To determine the dye concentration and amount of liquid trapped in the agglomerates, the agglomerates were dissolved in a water solution. Detailed experimental procedures may be found in [24] and Appendix B.

4.4 Results & Discussion

4.4.1 Effect of baffle with fluxtube on the lateral profile of the bubble gas flux

The triboelectric measurement technique provides the ratio of the local bubble gas flux to the cross-sectional average of the bubble flux, which is defined as the dimensionless bubble flux. The lateral profile of bubble gas flux in the absence of a baffle at 3 elevations from the distributor of \( z = 0.08, 0.28 \) and 0.4 m has been determined in an earlier study with the same equipment [23]. Figure 4.2 shows that, as expected, the baffle with fluxtube did not have much of an impact on the lateral profile of the bubble gas flux at locations well below the baffle and fluxtube (0.08 m above the gas distributor in Figure 4.1). The profile is symmetrical and typical of fluidized beds: bubble coalescence results in the bubble gas flux peaking at the central region of the bed for all three heights. This is in agreement with what has been observed by other researchers [14][30][31][32][33].
Figure 4.2: Impact of baffle with fluxtube on the lateral profile of the bubble gas flowrate 0.08 m above the grid (z = 0.08 m) – (Average of measurements at 4 fluidization velocities)
Adding a baffle with fluxtube affects the bubble distribution, resulting in profiles that are no longer symmetrical. Figure 4.3 shows that, with the baffle and fluxtube, at a height of $z = 0.4$ m above the distributor, i.e. well above the baffle with fluxtube (Figure 4.1), the bubbles have concentrated towards regions above the baffle and fluxtube. A comparison of the results with the baffle with and without fluxtube shows that the fluxtube results in a shift of the bubbles towards regions above the fluxtube (Figure 4.3). This results from the accumulation of bubble gas in the gas pocket that forms below the baffle and fluxtube from which gas escapes through the fluxtube and along the baffle edge. Because the lateral profile of this dimensionless bubble flux did not greatly change with the fluidization velocity, Figure 4.3 shows the average for the 4 tested fluidization velocities (0.1, 0.2, 0.3 and 0.45 m/s), at the height of 0.4 m above the distributor.

Figure 4.3: Impact of baffle with fluxtube on lateral profile of bubble gas flowrate at 0.4 m above grid ($z = 0.4$ m)
For a confirmation of these findings, photographs were taken from outside the transparent Plexiglas bed: Figure 4a confirms that large bubbles are concentrated near the bed center in the absence of baffle (Figure 4a) and the bubble concentration shifts above the baffle with the baffle and fluxtube configuration (Figure 4b).

Figure 4.4: Snapshots comparing bubble coalescence towards the bed center in the absence of baffle with bubble accumulation above the baffle and fluxtube region at $V_f = 0.45$ m/s. (a) without baffle; (b) with baffle and fluxtube
Figure 4.5 shows the lateral profile of the dimensionless bubble gas flowrate for \( z = 0.28 \) m, i.e. 0.11 m above the height of the baffle tip (see also Figure 4.1) for all the 3 baffle configurations. This figure shows that the baffle with fluxtube completely changed the flow pattern of the gas bubbles. The maximum bubble gas flux is found above the fluxtube region: this confirms that gas bubbles escape from the gas pocket below the baffle both through the fluxtube and along the baffle edge.

**Figure 4.5: Impact of baffle with fluxtube on bubble gas flowrate at 0.28 m above the distributor and the probe laterally facing the baffle (\( z = 0.28 \) m) – (Average of measurements at 4 fluidization velocities)**
4.4.2 Impact of nozzle penetration depth on liquid distribution

Figure 4.6 shows the gum Arabic experiments that were conducted to study the impact of nozzle penetration on liquid distribution with the baffle with fluxtube. The injection nozzle penetrations experimented were \( X_n = 0.05, 0.08, 0.13, 0.16 \) and \( 0.19 \) m from the bed wall on the baffle end.

![Diagram showing nozzle penetration orientation related to baffle with fluxtube for Gum Arabic method](image)

**Figure 4.6**: Nozzle penetration orientation related to baffle with fluxtube for Gum Arabic method
Figure 4.7 shows the reduction in the total mass of agglomerates that can be achieved by introducing a baffle with fluxtube. According to Figure 4.7, the baffle with fluxtube reduces the total mass of agglomerates for both fluidization velocities and all tested nozzle penetrations. The maximum relative reduction in the mass of agglomerates, when compared to the results obtained without a baffle, is achieved when the nozzle tip is positioned just above the nozzle tip ($X_n = 0.16$ m), and reaches 32 % and 84 %, respectively at fluidization velocities of 0.3 m/s and 0.45 m/s.

Figure 4.7: Impact of baffle with fluxtube on relative reduction in total mass of wet agglomerates compared to “No baffle” at $V_f = 0.3$ m/s & 0.45 m/s
Figure 4.8 shows that the baffle with fluxtube reduces the mass of all agglomerates, whatever their size, for a fluidization velocity of 0.45 m/s, and similar results were obtained for a fluidization velocity of 0.3 m/s (Figure 4.9).

Figure 4.8: Impact of baffle with fluxtube on mass of agglomerates for different nozzle penetrations at $V_f = 0.45$ m/s and GLR = 2 %. All data with the baffle and fluxtube, unless otherwise indicated.
Figure 4.9: Impact of baffle with fluxtube on mass of agglomerates for different nozzle penetrations at $V_f = 0.3$ m/s and GLR = 2 %. All data with the baffle and fluxtube, unless otherwise indicated.
Figure 4.10: Impact of baffle with fluxtube on relative reduction in total amount of liquid trapped in agglomerates compared to “No baffle” at $V_f = 0.3 \text{ m/s}$ and $0.45 \text{ m/s}$

The baffle with fluxtube also reduces the amount of injected liquid that is trapped in agglomerates, as shown by Figure 4.10. As with the mass of agglomerates (Figures 4.8 and 4.9), the best nozzle tip position is just above the tip of the baffle (Figure 4.10), corresponding to Figure 4.6 d. For the best nozzle position, when compared to the case without baffle, the amount of liquid trapped in agglomerates is reduced by 57 % and 65 %, respectively, at fluidization velocities of $0.3 \text{ m/s}$ and $0.45 \text{ m/s}$. 
Figure 4.11: Effect of Baffle with Fluxtube on the proportion of the injected liquid that is trapped in agglomerates for different nozzle penetrations at $V_f = 0.3$ m/s & GLR = 2%. All data is with the baffle and fluxtube, unless otherwise indicated.

Figure 4.11 and 4.12 show that, for both fluidization velocities, the baffle with fluxtube reduces the amount of water trapped in all agglomerate sizes. In Figure 4.11, it is observed that there is an exception to these findings for agglomerates in the range of 4000 to 9500 µm while experimented with a nozzle penetration of $X_n = 0.13$ m.
Figure 4.12: Effect of Baffle with Fluxtube on the proportion of the injected liquid that is trapped in agglomerates for different nozzle penetrations at $V_f = 0.45$ m/s & GLR = 2%. All data shown is with baffle and fluxtube, unless otherwise indicated.

4.4.3 Impact of reverse nozzle penetration on liquid distribution

The objective of injecting liquid in the reverse direction (the side of the bed opposite to the baffle) and exactly at the same elevation as the normal injection nozzle penetration was to determine whether this would improve the liquid distribution into the fluidized bed. The experiments were performed with the injection nozzle positioned at an elevation of $z = 0.36$ m above the grid, exactly opposite to the normal injection position. The nozzle penetrations tested with this configuration were $X_n = 0.05, 0.065, 0.08, 0.1$ and $0.15$ m relative to the
Plexiglas unit wall, as shown in Figure 4.13. The experiments were performed at a GLR = 2% and a fluidization velocity of $V_f = 0.45$ m/s.

Figure 4.13: Reverse injection nozzle penetration with baffle and fluxtube configuration for Gum Arabic method. $X_n$ represents the distance from the wall, where the injection nozzle is penetrated.
In reverse injection experiments, as shown in Figure 4.14, the maximum relative reduction in the mass of agglomerates was 74 %, when compared to the results obtained without a baffle, and was achieved at $X_n = 0.08$ m from the unit wall (Figure 4.13 c).

Figure 4.14: Impact of baffle with fluxtube while injecting reverse on relative reduction in total mass of wet agglomerates compared to “No baffle” at $V_f = 0.45$ m/s
Figure 4.15 shows that the maximum reduction in the percentage of liquid trapped in agglomerates was 51%, and was also achieved at the same nozzle penetration ($X_n = 0.08$ m from unit wall).

Figure 4.15: Impact of baffle with fluxtube while injecting reverse, on relative reduction in total amount of liquid trapped in agglomerates compared to “No baffle” at $V_f = 0.45$ m/s
Figure 4.16 shows that the baffle with fluxtube reduces the mass of all agglomerates, whatever their size, while Figure 4.17 shows that the baffle with fluxtube reduces the amount of water trapped in all agglomerate sizes.

**Figure 4.16**: Impact of baffle with fluxtube on reduction of mass of agglomerates for different reverse nozzle penetration at $V_f = 0.45$ m/s and GLR = 2 %. $X_n$ is the distance of nozzle from the wall where the injection nozzle is penetrated. (ⓒ Reverse injection with baffle and fluxtube) (ⓞ Normal injection with baffle and fluxtube)
Figure 4.17: Impact of Baffle with Fluxtube on liquid trapped in agglomerates for different reverse nozzle penetrations at $V_t = 0.45$ m/s & GLR = 2%. $X_n$ is the distance of nozzle from the wall where the injection nozzle is penetrated. (● Reverse injection with baffle and fluxtube) (-sizing Normal injection with baffle and fluxtube)

However, the results also show that none of the reverse nozzle configurations tested were as good as the best nozzle penetration, for the nozzle located above the baffle, with the nozzle tip aligned with the baffle tip (see section 4.4.2).
4.4.4 Effect of baffle with fluxtube in improving liquid distribution at various fluidization velocities

Figure 4.18 shows that, for all agglomerate size cuts, the reduction in liquid trapped in agglomerates that is achieved by increasing the fluidization velocity from 0.3 to 0.45 m/s is larger when a baffle with fluxtube is utilized. The beneficial effect of the baffle with fluxtube is more pronounced for the larger agglomerates, which are more likely to create problems in commercial Fluid Cokers.

Figure 4.18 also shows that the fluxtube enhances the baffle performance. The reduction in liquid trapped in agglomerates that is achieved by increasing the fluidization velocity is also greater than the baffle configuration for all agglomerates in the range of 600 μm – 4000 μm.

Figure 4.18: Reduction in mass of liquid trapped in agglomerates by increasing fluidization velocity from 0.3 m/s to 0.45 m/s for different baffle configurations
4.4.5 Impact of reduction in atomization gas on liquid distribution

In the absence of a baffle, previous studies [18][19][34][35] have shown that reducing the flowrate of atomization gas from 2% to 1.5% has a detrimental effect on the distribution of injected liquid on bed particles. Figure 4.19 confirms that similar results are obtained when a baffle with fluxtube is used.

Figure 4.19: Effect of GLR and Baffle with Fluxtube on the proportion of the injected liquid that is trapped in agglomerates for $V_f = 0.45$ m/s & $X_n = 0.16$ m (the GLR is the ratio of the atomization gas mass flowrate to the mass flowrate of injected liquid).

Figure 4.19 also suggests that using a baffle with fluxtube would allow the atomization gas flowrate to be reduced by 25% with no penalty in liquid distribution performance. Additional experiments, however, would need to be conducted to verify that wet solids do not deposit on the baffle at lower atomization gas flowrates.
4.4.6 Discussion

A baffle has two main impacts: it redirects the gas bubbles above the baffle region, greatly changing the lateral bubble gas flux profile, and it creates a region, between the lip of the baffle and the opposite wall, where there is a higher fluidization velocity and more intense fluidization. Earlier studies, which used a baffle without fluxtube, [23] showed that the first effect, the redirection of gas bubbles, is predominant. These studies [23] also showed that the best position of the nozzle tip was above the baffle lip, in the region of maximum bubble gas flux. The proposed interpretation of these results, based on experiments and a model from Mohagheghi [10], is that bubbles that enter the spray jet cavity near its base greatly accelerate the jet expansion-contraction cycle, which helps distribute the sprayed liquid over a larger number of bed particles, resulting in wet agglomerates that break up more quickly [23][36][37][38][39][40][41][42].

Experiments in the present study, with a spray nozzle introduced from the wall opposite the baffle, confirmed this interpretation. Although the baffle with fluxtube still improves the liquid distribution with this nozzle configuration, it is not as effective as when the nozzle is introduced above the baffle and sprays away from the baffle. This confirms that it is more effective to redirect the gas bubbles to the base of the jet cavity, near the nozzle tip, than to the tip of the jet cavity.

An unexpected result, however, was that, with the normal nozzle configuration, a baffle with fluxtube is more effective than a regular baffle and the optimum position of the nozzle tip is still just above the baffle lip. This means that a significant portion of the redirected bubbles flow on the side of the nozzle tip, away from the jet cavity. This suggests that there is an additional benefit in agitating this region. This could be explained by earlier studies [43][44][45] that showed that a significant flux of solids flows from the nozzle tip region into the jet cavity, ensuring that more solids are mixed with the sprayed liquid. This was explained by the relatively low pressure in the jet cavity, just downstream of the nozzle tip [43]. These earlier studies also showed that gas is entrained from the region near the
nozzle tip to the jet cavity, resulting in poor fluidization or, even defluidization. A possible explanation of the beneficial effect of the fluxtube observed in the present study is that the additional bubbles redirected to this poorly fluidized region improve the mobility of the particles in this region and facilitate their beneficial flow into the jet cavity. Some of the gas from these bubbles may also be entrained into the jet cavity, contributing to the acceleration of the jet expansion-contraction cycle.

4.5 Conclusions

Conclusions drawn from this study are as follows:

- A new triboelectric technique shows that a ring baffle with fluxtube, thoroughly changes the bubble flux pattern through the bed. Instead of concentrating at the center, bubbles are first concentrated above the fluxtube and then the flow pattern shifts above the baffle tip region as they rise in the bed.

- The baffle with fluxtube can greatly improve the liquid distribution. This was confirmed with a Gum Arabic method that provides the mass and liquid content of agglomerates formed when spraying a liquid through a feed nozzle located above the baffle.

- The ideal injection nozzle position is when its tip is located just above the tip of the baffle with fluxtube ($X_n = 0.16$ m) where an optimal liquid distribution is obtained.

- Addition of a fluxtube to the baffle showed improvement in liquid distribution into the fluidized bed with lower amount of agglomerates formed. The improvement with fluxtube was minor compared to baffle without fluxtube but, very significant when compared to “no baffle” results.
4.6 References


Chapter 5

5 Conclusion and Recommendation

5.1 Conclusions

The findings from the present research work can be summarized as follows:

- A reliable solution to measure the bubble gas distribution in fluidized beds with high fluidization velocities, is the implementation of sturdy triboprobes. They can be used to conveniently measure signal characteristics such as power and average frequency which would help determine the fluidized bed local bubble flux. The signal analysis tools applied were power spectrum, cycle analysis and signal moments. The sturdy triboprobes were able to measure large and turbulent bubbles released from the baffle tip and the fluxtube, which under such conditions would have damaged less sturdy probes such as optical fiber probes.

- A baffle redirected gas bubbles in gas pockets below the baffle, from which gas escaped along the baffle edge. By adding a fluxtube to the baffle, part of the accumulated gas also escaped through the fluxtube. Instead of concentrating at the center, bubbles are first concentrated at the tip of the baffle and this flow pattern is retained in regions well above the baffle.

- The distribution of liquid sprayed into the fluidized bed was greatly improved with the baffle. This was confirmed with two independent measurement methods: a conductance method that measures the breakage rate of wet agglomerates and a Gum Arabic method that provides the mass and liquid content of agglomerates.

- To achieve an optimal liquid distribution in the presence of a baffle, the ideal nozzle penetration would be to have the nozzle tip aligned with the baffle tip.
• Injecting liquid opposite to the baffle with fluxtube do not present results as good as the normal injection from above the baffle.

• Adding a fluxtube to the baffle slightly improved the liquid distribution into the fluidized bed with a lower amount of agglomerates being formed.

5.2 Recommendations

• To confirm the beneficial impact of baffles on liquid distribution:
  - Experiments at high temperatures simulating the Fluid Coker
  - Inject liquid at higher elevations from the baffle at higher fluidization velocities

• The triboelectric measurements were made with a triboprobe which needed to be moved for every local measurement. To design a unit with multiple, fixed triboprobes would be of great advantage in saving time to perform experiments. It also helps in enhancing the accuracy of measurements, since all the data is recorded with a single run for each vertical position in the bed and can be analyzed together.

• Design and construction of a new taller unit would help measure the impact of elevation on bubble gas profiles and liquid distribution.

• A wider unit would be beneficial in injecting liquid for a wider nozzle penetrations to test the impact of baffle with and without fluxtube on liquid distribution. This will avoid the present injection limitation with respect to the jet impacting the opposite wall.

• The new unit can also be helpful in studying the impact of injecting liquid below the baffle.
• Experiments could be conducted to confirm the beneficial impact of a baffle on liquid distribution with a wide range of nozzle conditions (liquid and atomization gas fluxes) and geometries.

• As done in other studies [1], the bed conductivity should be measured during the injection of gum Arabic solution to confirm that the bed fluidization properties are not affected by the liquid injection before the bed is defluidized and dried.

• The new unit could be equipped with electrode plates across the unit wall, to develop the conductance experiments as an alternative.

• Experiments with X-ray videos to determine:
  
  o Impact of baffle on gas bubbles
  
  o Impact of baffle on liquid distribution (with radio-opaque liquid)
  
  o Impact of liquid spray on gas bubbles

5.3 Reference

Appendices

Appendix A: Comparison of the best signal characteristics combination

The bubble flux profile for the best 3 combination of signal characteristics shown in Table 2.2 and applied in Equation (6)(Chapter 2), is compared to verify the resulting bubble profile pattern. This comparison is made for all the 3 elevations of the bed discussed in chapter 2 (shown in Figure 2.1) which is $z = 0.08$, 0.28 and 0.4 m.

![Graph showing comparison of signal analysis methods for $z = 0.08$ m with fluxtube.](image)

**Figure A-1:** Comparison of 3 signal analysis methods for $z = 0.08$ m with fluxtube. (P denotes Power, f is frequency, S.D. stands for standard deviation, $\tau$ shows cycle time and Avg. is Signal Average)

Comparing the best 3 combination of signal characteristics at elevation of $z = 0.08$ m from the grid, where there is likely no impact of baffle and fluxtube on the bubble motion and the bed is shallow, all the 3 combinations follow the same pattern confirmed by Darton’s
theory of bubble coalescence towards the bed center as shown in Figure A-1. The combination of standard deviation and cycle time shows variation in the $\frac{q_{bi}}{\overline{q_b}}$ value compared to the other 2 combinations, however all the 3 patterns are similar.

Figure A-2: Comparison of 3 signal analysis methods for $z = 0.28$ m with fluxtube

Figure A-2 confirms that similar lateral profile of the bubble gas flux were obtained when using the best three combinations of table 2. However, the results using the power and average frequency seemed smoother and did not present the local bumps showed by the other two combinations, which are likely spurious. Also with results of combining power with average frequency, the bubble accumulation above the fluxtube region is more pronounced.
Figure A-3: Comparison of 3 signal analysis methods for z = 0.4 m with fluxtube

Similar to results at z = 0.08 m and z = 0.28 m, the bubble flux profile for all the 3 best signal characteristic combinations follow a similar pattern at z = 0.4 m, as shown in figure A-3. All 3 bubble profiles, display accumulation of bubbles above the baffle and fluxtube region. Similar to results at z = 0.28 m, variations in the $q_{bi}/\overline{q_b}$ values can be seen at local tribo penetrations of 0.1 m and 0.25 m from the bed wall end.
A map is used to plot and compare the beds’ combination of signal characteristics in equation 6 (Chapter 2), at three different elevations in the fluidized bed shown in Figure A-4. White and light gray levels represent regions with high bubble gas flux and darker gray levels show regions with low bubble gas flux. The pattern is a time-averaged measurement of the bubble gas flux. As discussed earlier, similarity of the pattern of different combination of signal characteristics is visible in the map (Figure A-4) at each elevation.
Appendix B: Characterization of agglomerates (Gum Arabic method)

Throughout each experiment, the bed was initially fluidized at a velocity of 0.3 or 0.45 m/s. The prepared GA (binder) solution was injected for a period of 6 seconds and immediately after the injection, the fluidization velocity was lowered below the minimum fluidization velocity to slump the bed and leave it to dry under ambient conditions until the following day. The dried bed was emptied completely and the agglomerates were recovered by sieving. The size distribution of the agglomerates, and the amount of water trapped in the agglomerates for each size cut, were measured with the methods developed by Pardo [24] as follows:

The agglomerates were classified into 9 different cuts:

1. Macro-agglomerates greater than 600 microns; 6 sieve sizes were used to classify them as follows: \( d_{\text{aggl}} \geq 9500 \, \mu m, \geq 4000 \, \mu m, \geq 2000 \, \mu m, \geq 1400 \, \mu m, \geq 850 \, \mu m, \text{ and } \geq 600 \, \mu m \)

2. Micro-agglomerates smaller than 600 \( \mu m \) and greater than 355 \( \mu m \); these were classified into 3 cuts: \( d_p \geq 500 \, \mu m, \geq 425 \, \mu m, \text{ and } \geq 355 \, \mu m \)

To recover the micro-agglomerates, a representative sample of 5 kg was taken with a chute splitter from the remaining individual bed particles, after the initial screening of the whole bed had recovered the macro-agglomerates. In order to obtain the amount of dye concentration for each size cut, the agglomerates of each size cut were weighed and dissolved in water with an approximate ratio of 1:3. The resulting solution was transferred to a vial and then centrifuged for 10 minutes at 4500 rpm using a Thermo Scientific Sorvall Legend XI Centrifuge (Thermo Fisher Scientific – Waltham, USA) to ensure there were no solid particles in the solution before measuring the dye absorbance. A spectrophotometer was used to determine the blue dye absorbance in the solution at a wavelength of \( \lambda_{\text{max}} = 630 \, \text{nm} \). The bed was dried in a manner to avoid agglomerate break up after each liquid injection and, therefore, the resulting liquid to solid ratio for each size cut of agglomerates corresponded to the initial amount of liquid trapped in agglomerates immediately following the injection.
Since the micro-agglomerates were in the size range of the bed particles, the amount of fines in the agglomerates was determined in order to obtain the actual mass of agglomerates, using a method developed by Pardo [24]. For the 3 size cuts of micro-agglomerates, a HELOS Particle Size Analyzer (PSA) (Sympatec – Clausthal-Zellerfeld, Germany) was used to detect the amount of fines in the representative sample of each size cut. Pardo [24] based this measurement method on the assumption that if there is no segregation of fine particles in the agglomerates, the size distribution of particles trapped in agglomerates would be the same as the size distribution of bed particles. Therefore, for any given size cut of micro-agglomerates, the mass of trapped particles were calculated by knowing the sand mass of this particular size cut (M_s) which was weighed after sieving the representative sample, and the fraction of fines in the sample (X_f) and in the initial bed mass (X_f_bed) :

\[ m_p = M_s \times \frac{X_f}{X_{f,\text{bed}}} \]

Then by knowing the mass of blue dye and the GA binder which have been obtained from the analysis to determine the mass of water trapped in agglomerates, the mass of agglomerates for each size cut can be calculated:

\[ M_{\text{agg},S_i} = m_p + m_{\text{dye}} + m_{\text{GA}} \]

Having known the mass of agglomerates for each size cut of the given sample, the total mass of micro-agglomerates for a given size cut in the bed \( M_{\text{agg},i} \) is calculated as follows:

\[ M_{\text{agg},i} = M_{\text{agg},S_i} \times \frac{m_{<600\mu m}}{m_R} \]

where \( m_{<600\mu m} \) is defined as the total mass of solids in the bed smaller than 600 µm and \( m_R \) is the representative sample taken from \( m_{<600\mu m} \).
Finally, the cumulative amount of water trapped in the agglomerates for each size cut could be determined, and based on the results for each test, the optimum nozzle penetration ($X_n$), GLR and fluidization velocity could be determined.

Considering the presence of fluxtube beneath the baffle, the nozzle penetrations experimented (Figure 4.6) for the Gum Arabic method were slightly different from experiments with just baffle. The nozzle penetrations tested for experiments with fluxtube were $X_n = 0.05, 0.08, 0.13, 0.16$ and $0.19 \text{ m}$.

The cumulative mass of agglomerates for experiments with baffle at two different fluidization velocities of $V_f = 0.3$ and $0.45 \text{ m/s}$ is shown in the following figures. Figure B-1 shows that the baffle reduces the mass of all agglomerates, whatever their size, for a fluidization velocity of $0.3 \text{ m/s}$ and according to Figure B-2, similar results were obtained at $V_f = 0.45 \text{ m/s}$. Both figures show that the best nozzle configuration for experiments with baffle is with the nozzle tip aligned above the baffle tip at $X_n = 0.16 \text{ m}$. 
Figure B-1: Impact of baffle on mass of agglomerates for different nozzle penetrations at $V_t = 0.3$ m/s and GLR = 2%. All data with the baffle, unless otherwise indicated.
Figure B-2: Impact of baffle on mass of agglomerates for different nozzle penetrations at $V_f = 0.45 \text{ m/s}$ and GLR = 2 %. All data with the baffle, unless otherwise indicated.
Figure B-3: Liquid to solid ratio for all agglomerate size cuts for two configurations: without baffle and with baffle and fluxtube with a nozzle penetration of $X_n = 0.16$ m and at a fluidization velocity of $V_f = 0.45$ m/s.

Results of gum Arabic experiments show that with baffle and fluxtube the ratio of liquid trapped in agglomerates reduces for all size cuts, when compared to the case without baffle at the optimum nozzle penetration of $X_n = 0.16$ m and $V_f = 0.45$ m/s.
Appendix C: Video of stable spray jet

Several nozzle configurations were experimented with open air jet sprays to achieve the stable jet shown in the video below. The stable nozzle configuration is shown in figure 3.4.

Stable jet spray.mp4

https://vimeo.com/221653209

Spray Jet Video
Appendix D: Bubble flux analysis

As discussed earlier in Chapter 2, 3 correlations were defined to determine the local bubble flux and eventually obtain the cross-sectional average bubble flux. The results were compared with the experimental values. The graphs showing these comparisons were shown in chapter 2. The graphs for the results shown in table 2.1, 2.12 and 2.4 which were not presented in chapter 2 are displayed as follows in the order of increasing error.

![Figure D - 1: Impact of Cycle Time on calculating bubble gas flux (Equation 1)](image)

\[ q_{bi} = \alpha (1/T)^{\beta} - \text{Cycle time - with fluxtube} \]

- baffle
- dipleg
- grid

\[ y = 0.7285x + 0.1147 \]
\[ R^2 = 0.9563 \]

<table>
<thead>
<tr>
<th>alpha</th>
<th>beta</th>
<th>error</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.029807</td>
<td>1.088977</td>
<td>0.12737</td>
</tr>
</tbody>
</table>
$q_{bi} = \alpha (\text{Avg.})^\beta - \text{Average - with fluxtube}$

\[
y = 0.6438x + 0.1358 \\
R^2 = 0.886
\]

Figure D - 2: Impact of Signal Average on calculating bubble gas flux (Equation 1)
Figure D - 3: Impact of $V_{statistic-max}$ on calculating bubble gas flux (Equation 1)
Figure D - 4: Impact of Kurtosis on calculating bubble gas flux (Equation 1)
Figure D - 5: Impact of Average Frequency on calculating bubble gas flux (Equation 1)
Figure D - 6: Impact of combining Kurtosis and Cycle Time on calculating bubble gas flux (Equation 6)
Figure D - 7: Impact of combining Signal Average and Kurtosis and on calculating bubble gas flux (Equation 6)
Figure D - 8: Impact of combining Signal Average and Cycle Time on calculating bubble gas flux (Equation 6)
Figure D - 9: Impact of combining $V_{\text{statistic-max}}$ and Cycle Time on calculating bubble gas flux (Equation 6)
Figure D - 10: Impact of combining Standard Deviation, Cycle Time and Average Frequency on calculating bubble gas flux (Equation 8)
Figure D-11: Impact of combining Power and Average Frequency to determine the bubble gas flux for the “No Baffle” Configuration

Figure D-11 shows the match between the calculated and measured cross-sectional average bubble flux indicated in table 2.6 using equation 2.9. With the signal characteristic combination shown, the least error was achieved for the “No Baffle” configuration and was thus determined as the best combination to measure the cross-sectional average bubble flux.
Figure D - 12: Impact of combining Power and Average Frequency to determine the bubble flux for the “Baffle” Configuration (Equation 9)

Similar to the “No Baffle” configuration and as shown in table 2.6, the combination of power and average frequency proves to be the best match for the “Baffle” configuration as well, in determining the average bubble gas flux.
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