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Effect of Spray Jet Interactions on the Liquid Distribution in a Fluidized Bed

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Graduate Program in Chemical and Biochemical Engineering

A thesis submitted in partial fulfillment of the requirements for the degree in Master of Engineering Science

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EFFECT OF SPRAY JET INTERACTIONS ON THE LIQUID DISTRIBUTION IN A FLUIDIZED BED

(Thesis format: Monograph)

by

Helal H. Elkolaly

Graduate Program in Chemical and Biochemical Engineering

A thesis submitted in partial fulfillment of the requirements for the degree of Masters of Engineering Science

The School of Graduate and Postdoctoral Studies
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Abstract

Many processes use spray nozzles to introduce liquid into a fluidized bed. One such process is the Fluid Coking Process™ which is used to upgrade bitumen. Poor liquid distribution in a Fluid Coker™ proves detrimental as agglomerates form which introduce mass and heat transfer limitations. The main objective of this thesis is to determine the effect of interacting spray jets on the liquid distribution in a Fluid Coker. This was accomplished by using a low temperature experimental model of Fluid Coking to estimate the liquid distribution. Preliminary screening of nozzle positions was accomplished by the use of conductance measurements. A binder solution was then utilized to further investigate the agglomerates produced for the most interesting nozzle positions. Three types of spray nozzle interactions were investigated to determine their effect on the liquid distribution in a Fluid Coker. Depending on the relative positions of the spray nozzles, improved or deteriorated performances were observed for each type of configuration.

Keywords:

Fluidized bed, Fluid Coking, Spray nozzle, Spray Jet, Spray nozzle interactions, Conductance measurements, Agglomerate formation, Liquid distribution
Co-Authorship Statement

Chapter 3

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Nomenclature

\( a \) Constant corresponding to the slope of the curve relating conductance to free moisture (S)

\( A_{\text{bed}} \) Cross sectional area of the fluidized bed (m\(^2\))

\( b \) Constant corresponding to the y-intercept of the curve relating conductance to free moisture (S)

\( C_{\text{blue}}^* \) Mass of blue dye in the agglomerate divided by the mass of blue dye in the original blue dye container (-)

\( C_{\text{yellow}}^* \) Mass of yellow dye in the agglomerate divided by the mass of yellow dye in the original yellow dye container (-)

\( D \) Diameter of a spray jet (m)

\( d_B \) Bubble diameter (m)

\( d_N \) Spray nozzle diameter (m)

\( E \) Ratio of evaporation rate of liquid from the fluidized bed to the mass of bed solids (s\(^{-1}\))

\( f_e \) Flowrate of evaporating liquid from the fluidized bed (kg/s)

\( F_L \) Flowrate of injected liquid (kg/s)

\( f_w \) Ratio of solids in the wake to bubble volume (-)

\( G \) Dry conductance subtracted from the actual conductance (S)

\( G_{\text{local}} \) Dry conductance subtracted from the local conductance using a mixing model (S)

\( \text{GLR} \) Gas to liquid ratio (-)

\( H^* \) Humidity of saturated vapour leaving the fluidized bed (kg water/kg air)

\( H_{\text{in}} \) Humidity of incoming fluidization air (kg water/kg air)

\( k \) Thermal conductivity of coke layer (W/m\(k\))

\( L \) Instantaneous penetration length of the spray jet (m)

\( L/S \) Liquid to solid ratio (-)

\( L_{\text{bc}} \) Bubble capture length of the spray jet (m)

\( L_{\text{jet}} \) Average penetration length of the spray jet (m)

\( L_{\text{max}} \) Maximum penetration length of the spray jet (m)

\( L_{\text{min}} \) Minimum penetration length of the spray jet (m)

\( M^* \) Normalized fraction of yellow dye in an agglomerate (-)
\( m_{<600\mu m} \) Mass of solids in the bed less than 600 \( \mu m \) (kg)
\( m_{\text{dye}} \) Mass of dye in the microagglomerates (kg)
\( m_e \) Mass of evaporated liquid (kg)
\( m_{\text{GA}} \) Mass of gum arabic in microagglomerates (kg)
\( M_L \) Mass of injected liquid (kg)
\( m_{Lr} \) Mass of liquid remaining in agglomerates (kg)
\( m_p \) Mass of particles trapped in a given size cut of microagglomerates in the sample (kg)
\( m_R \) Mass of an individual size cut of the microagglomerate sample (kg)
\( M_s \) Mass of solids in the fluidized bed (kg)
\( m_{\text{agg},i} \) Mass of microagglomerates in the bed (kg)
\( m_{\text{agg},Ri} \) Mass of microagglomerates in a given size cut in the sample (kg)
\( Q_a \) Atomization gas flowrate (m\(^3\)/s)
\( Q_{cb} \) Flowrate of bed bubbles entering the spray jet cavity (m\(^3\)/s)
\( Q_g \) Flowrate of gas entering the spray jet cavity (m\(^3\)/s)
\( R \) Radius of an agglomerate (m)
\( t \) Time (s)
\( t_0 \) Time at end of injection (s)
\( T_B \) Temperature of the bed (°C)
\( t_c \) Time for full conversion (s)
\( t_{ek} \) Time required to evaporate the liquid for an individual replicate (s)
\( t_{\text{exp}} \) Expansion cycle time of the spray jet (s)
\( t_i \) Initial time of integration of the conductance curve (s)
\( T_R \) Temperature of the reaction front where thermal cracking occurs (°C)
\( S \) Total mass of solids wetted in a single jet expansion (kg)
\( S_{R} \) Mass of solids trapped in the wake of a bubble (kg)
\( u_{mf} \) Minimum fluidization velocity (m/s)
\( v_B \) Volume of released bubble from the spray jet (m\(^3\))
\( v_g \) Superficial fluidization velocity (m/s)
\( X \) Free moisture in the fluidized bed (-)
\( X_0 \) Free moisture at the end of injection (-)
\( X_{\text{agg}} \) Free moisture released from agglomerates (-)
\( X_{agg0} \) Free moisture released from agglomerates at the end of injection (-)

\( X_{all} \) Fraction of injected liquid to solids mass (-)

\( x_f \) Fraction of fines in a size cut in the microagglomerate sample (-)

\( x_{ibed} \) Fraction of fines in the original bed mass (-)

\( X_i \) Free moisture in the fluidized bed at start time of integration (-)

\( y \) Conductance (S)

\( y_{dry} \) Conductance of the dry fluidized bed (s)

\( \alpha \) Angle of inclination of the spray nozzle from the vertical (°)

\( \beta \) Constant corresponding to the slope of the conductance curve multiplied by \( X_{all} \) (S)

\( \Gamma \) Constant used in the simplified thermal model (m\(^2\)/s)

\( \gamma_0 \) Fraction of injected water that is free moisture at the end of injection (-)

\( \Delta H \) Enthalpy change of the liquid (J/kg)

\( \Delta t_m \) Mixing time required for the injected liquid to reach the electrode (s)

\( \Delta x \) Predicted horizontal distance between jet tips (cm)

\( \Delta x_n \) Horizontal distance between spray nozzles (cm)

\( \Delta z \) Vertical distance between nozzle ports (cm)

\( \eta \) Normalized radial position of the reaction front (-)

\( \theta \) Half angle of expansion of a spray jet (°)

\( \lambda \) Constant used to simplify the jet expansion model (m)

\( \rho_{mf} \) Bulk density of solids at minimum fluidization (kg/m\(^3\))

\( \rho_s \) Density of the coke solids (kg/m\(^3\))

\( \tau \) Agglomerate breakup time constant (s)
1 Introduction

The proportion of crude oil produced worldwide that is heavy oil is steadily increasing due to depleting lighter oil reserves. Initially, many oil reserves consisted of very sweet crude oil that was easier to process and less expensive to produce. Many of these oil reserves have been depleted, and thus heavier crude oil has become one of the only options left for many countries. Thus newer processes are being implemented to upgrade heavier oils into more valuable lighter oils. Canada’s rich supply of bitumen from the oil sands has led to the use of the Fluid Coking Process™, which produces close to 20% of finished petroleum product exports (ExxonMobil n.d.).

Although the Fluid Coking Process produces a large portion of Canada’s oil, the operation of a Fluid Coker™ is not fully understood by the industry. The use of multiple interacting spray nozzles in a Fluid Coker has not been studied, although it might improve the performance of the Fluid Coker. Without understanding how multiple spray jets interact, the dynamics in the reactor cannot be appreciated since different hydrodynamics could result, which affect the liquid distribution and ultimately the liquid content in the agglomerates (House, P 2007).

The following section provides a brief overview of the major processes used to upgrade heavy oil: Fluid Catalytic Cracking and Fluid Coking™. Emphasis is placed on the Fluid Coking Process as the present work focuses on ways to optimize a Fluid Coker. The formation and properties of agglomerates in a Fluid Coker are then discussed to discover what research has been performed in this field. Finally, the objectives of this thesis are discussed.
1.1 Fluid Catalytic Cracking

One of the most popular catalytic processes that incorporates the use of a fluidized bed is the Fluid Catalytic Cracking (FCC) process. The FCC process is used to upgrade gas oil, which consists of high-boiling petroleum fractions, into more valuable products such as gasoline and diesel (Patel et al. 2012). The reactor consists of a riser where heated heavy oil is pumped through atomizing nozzles that distribute the gas oil onto zeolite catalysts. The design of zeolite catalysts is such that the surface area is maximized to allow for more active sites. The hot zeolite catalysts vapourize the oil and catalytically crack it into valuable products. As the reaction is endothermic, heat must be provided; in this case, the hot catalysts provide this heat. This reaction usually occurs within three seconds while the catalyst particles are lifted through the riser and mixed with atomized oil.

The catalyst is then separated from the vapours, usually by the use of cyclones. After the catalyst is separated from the vapours, steam is used to strip the catalyst of any valuable hydrocarbons that could be trapped in the catalyst. The catalyst is then sent to the regenerator where residual hydrocarbons on the catalyst are burned, using air, to produce heat for the cracking reaction. The regenerated catalyst is returned to the riser to complete the cycle. One disadvantage of this process is that heavy oils cannot be used as they would poison the zeolite catalyst due to the high content of impurities (Sadeghbeigi 2012).

1.2 The Fluid Coking Process™

Bitumen is a type of heavy oil found in Canada’s oil sands. A popular process used to upgrade bitumen from Canada’s oil sands is the Fluid Coking Process. This is similar to the FCC process with the main difference being the cracking process: thermal cracking is incorporated instead of catalytic cracking. In this process, heavy oils with an API gravity of 0 to 20° such as vacuum bottoms are preheated to 200-400°C to allow for a more flowable liquid and to reduce the amount of heat required in the Fluid Coker. The preheated oil is injected through nozzles into a
fluidized bed of coke particles at 500 to 600°C and a pressure between 34.5 and 103.4 kPa. Liquid bitumen is atomized using steam which is also injected through the nozzles into the fluidized bed. The nozzles used to inject bitumen and steam into the bed are arranged along the circumference of the vessel and staggered vertically. Since the flowrate of gas in the vessel increases vertically due to volatile gases rising, the vessel diameter increases vertically. As such, the fluidization velocity at the bottom of the vessel must be monitored to ensure a fluidization velocity between 0.30 to 0.91 m/s throughout the bed. When bitumen contacts the coke particles it is thermally cracked into smaller more volatile compounds, which are further refined downstream, and heavier compounds which form solid coke. In this continuous operation, the coke particles are transported to a furnace, where excess coke is removed and the remainder is reheated by combustion with air (Pfeiffer et al. 1959).

The thermal cracking reaction occurs in the liquid film formed on the surfaces of the hot coke particles. The high viscosity of the sprayed bitumen results in the formation of agglomerates as coke particles adhere to the viscous bitumen. This results in larger agglomerates which results in heat and mass transfer limitations, more coke particles being formed and bitumen being burned in the furnace instead of being cracked to form valuable products (House, P 2007). Thermal cracking is endothermic and the coke particles leave the Fluid Coker at a lower temperature than their inlet temperature.
Figure 1.1: Simplified process flow diagram of a Fluid Coker reactor (Prociw et al. 2014)

The coke particles are transported to a furnace where combustion of some of the coke reheat the coke particles before they are re-circulated to the Fluid Coker. Not only does this limit natural gas consumption in the furnace, steam and flue gas are also produced to be used in other parts of the plant, making the process sustainable (ExxonMobil n.d.).

In the stripping zone at the bottom of the vessel, steam is injected to displace hydrocarbon vapours from the coke produced. Placing the stripper section at the bottom of the vessel is advantageous since heavier coke particles will collect at the bottom of the vessel. In this way, larger coke particles can be reduced in size by injecting attrition steam at high velocities of about 200-3000 ft/s, thereby controlling the size of coke in the bed and ensuring the rate of thermal cracking as will be discussed below. To ensure turbulent conditions in the stripper so that the contact efficiency between coke and steam is increased, baffles are added (Pfeiffer et al 1959).
The vapours rise through the Fluid Coker until they exit to be refined in downstream processes. These vapours carry entrained solids so the bed surface should be below the top of the vessel to ensure most of the solids return to the bed. It is desirable however, to have some entrained hot coke particles in the disengaging space to maintain the temperature in this zone so that vapours do not condense and return to the bed where they would be further cracked. The vessel diameter decreases in the disengaging zone to increase the vapour velocity, thereby decreasing the vapour residence time, and to accelerate entrained coke particles so they can scour the walls of any adhered coke (Gray et al. 1994).

Separating entrained solids from vapours is accomplished by the use of cyclones. Fine coke particles from the disengaging zone enter the cyclone and are separated from vapours through centrifugal forces. The solids that are entrained and transported to the cyclones and scrubber should ideally be dry so that fouling of vessel walls can be minimized. The recovered solids are returned to the bed through diplegs so that fine dust particles are not released to the atmosphere. Since the coke particles can plug the dipleg, aeration gas is often used to ensure smooth flow of particles from the cyclone to the Fluid Coker.

Vapours from the cyclone rise through chimneys to the scrubber, which is placed above the Fluid Coker. Baffles are positioned above the entrance of the vapours into the scrubber and throughout the scrubber. The baffles serve two purposes: creating more turbulent conditions and collecting any condensate that forms. The vapours may condense along the walls or on the surface of the baffles due to a difference in temperature between the vapours and the vessel. Condensed vapours can return to the cyclones and Fluid Coker, which would interfere with the operation as agglomerates may form. To minimize the effects of condensing vapours in the scrubber, the baffles and the vessel wall should be well insulated. Another option is to use heat tracing, which can be in the form of superheated steam or flue gas coils. The temperature at the bottom of the scrubber is controlled to condense a portion of the vapour. This is done to remove heavy metals and other impurities that were present in the original feed. The condensate formed is removed through side streams and can be partly recycled to the Fluid Coker.
Uncondensed vapours rise through the scrubber to the fractionating section of the scrubber. Bubble caps are used to fractionate the vapours into products which are pumped through side streams for further downstream processing. Uncondensed vapours are cooled by the use of condensers and a portion is recycled to the fractionating part of the scrubber, while the rest is sent to a settling drum to separate vapours, gas oils, and water (Pfeiffer et al 1959).

The vessel dimensions and operating conditions of a typical Fluid Coker are listed in Table 1.1.

<table>
<thead>
<tr>
<th>Section</th>
<th>Diameter (m)</th>
<th>Length (m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Disengaging zone</td>
<td>2.7</td>
<td>6.1</td>
</tr>
<tr>
<td>Wide Diameter (Dense Bed)</td>
<td>3.4</td>
<td>4.9</td>
</tr>
<tr>
<td>Middle section cone</td>
<td>1.2 (narrow section), 3.4 (wide section)</td>
<td>10.4</td>
</tr>
<tr>
<td>Stripper section</td>
<td>1.2</td>
<td>3.0</td>
</tr>
<tr>
<td>Temperature (°C)</td>
<td>510</td>
<td></td>
</tr>
<tr>
<td>Outlet pressure (kPag)</td>
<td>75.8</td>
<td></td>
</tr>
<tr>
<td>Weight of solids in bed (kg)</td>
<td>64,000</td>
<td></td>
</tr>
<tr>
<td>Gas velocity (m/s)</td>
<td>0.3 (bottom), 1.1 (top)</td>
<td></td>
</tr>
</tbody>
</table>

1.3 Agglomerate Formation and Liquid Distribution

Agglomeration of particles occurs in most gas-liquid-solid fluidized beds. In the pharmaceutical industry, agglomerates are desirable as they allow for easier transport of powders and can reduce issues encountered due to fine particles (Weber et al. 2009). Processes that incorporate thermal or chemical reactions do not benefit from the formation of agglomerates due to the mass and heat transfer limitations that result. In thermal cracking processes such as the Fluid Coking Process, agglomerates limit the yield of lighter hydrocarbons which results in a loss of feed. The feed that is trapped in agglomerates is lost in the burner; however, prior to reaching the burner agglomerates can contribute to fouling of fluidized bed internals. Fouling of internals can affect the hydrodynamics in the Fluid Coker as well as create temperature gradients. The fouling
created by the wet agglomerates can completely plug smaller sections of the Fluid Coker, especially towards the bottom since wetter agglomerates are denser. This can lead to fluidization issues and plugging of downstream processes. Therefore, fouling is not desired as it may eventually cause unplanned shutdowns which are quite expensive (Sanchez et al. 2013).

Bruhns and Werther (2005) proposed a model to determine the mechanism of agglomerate formation in a fluidized bed. As liquid is injected in the fluidized bed, particles are entrained in the gas-liquid jet and form agglomerates. These agglomerates are then dispersed throughout the bed where they break due to encounters with other particles in the fluidized bed.

Similar to Bruhns and Werther (2005), Gray (2002) proposed a mechanism to describe agglomerate formation in a Fluid Coker. The mechanism was found to be composed of three main steps, with the first step involving the formation of a liquid-gas jet, with fine liquid droplets, which entrains solids. Then, agglomerates are formed by the wetting of solids. In the

---

**Figure 1.2: Mechanism of agglomerate formation (Bruhns and Werther, 2005).**
final stage, agglomerates break due to shear forces in the bed and evaporation of liquid resulting in drier, weaker agglomerates.

Ariyapadi et al. (2003) also studied the mechanism of agglomerate formation by injecting a radio opaque liquid tracer mixed with ethanol and analyzing the jet cavity using X-ray imaging. Agglomerates were found to form at the jet tip where the pressure was lower than near the nozzle tip. This was due to lower shear forces being present at the jet tip, which promoted the formation of agglomerates.

Schafer and Mathiesen (1996) studied the effect of viscosity on the formation of agglomerates in a shear mixer. Two mechanisms were proposed to explain the initial wetting of particles by liquid droplets. The first mechanism is wetting by distribution in which liquid droplets are distributed on the surface of surrounding particles and initial nuclei are formed. This mechanism was found to dominate when the particles and droplets were of similar size. In the Fluid Coking Process, the coke particles and bitumen droplets formed by the efficient spray nozzles have approximately the same Sauter mean diameter (House, P 2007). Therefore, wetting by distribution predominates in the Fluid Coking Process. Schafer and Mathiesen (1996) determined that the second possible mechanism for the formation of agglomerates is wetting by immersion in which initial nuclei are formed by large droplets capturing individual particles.

The destruction of agglomerates in a Fluid Coker is necessary to maintain proper operation. Salman et al. (2004) presented two different modes of breakage of agglomerates that depended on the impact velocity. Larger and more porous agglomerates were found to promote chipping of agglomerates. Subero and Ghadiri (2001) proposed two types of breakage: localized damage and distributed damage. This is similar to what Weber et al. (2006) found: at low gas velocities surface erosion of agglomerates predominates whereas at high gas velocities fragmentation predominates. Weber et al. (2009) found that for cases where surface erosion is the dominant destruction mechanism, larger and denser agglomerates are more stable than smaller, less dense agglomerates. It was also determined that increasing the liquid viscosity up to a value of 3 cP
results in significantly more stable agglomerates which is detrimental to the Fluid Coking Process.

Weber et al. (2006) and Dunlop et al. (1958) found that smaller particles form more stable agglomerates than larger ones. Dunlop et al. (1958) determined that particles larger than 70 μm form agglomerates that are broken apart due to shear forces in a fluidized bed. Parveen et al. (2011) concluded that higher liquid and bulk density in agglomerates creates more stable agglomerates. Similar to what Weber et al. (2006) and Dunlop et al. (1958) found, Parveen et al. (2011) determined that increasing the size of individual particles in an agglomerate reduces the agglomerate’s stability.

1.4 Methods to Characterize Liquid Distribution in a Fluidized Bed

1.4.1 Non-intrusive Methods

Mohagheghi et al. (2013) used a capacitance method to determine the rate at which liquid is released from agglomerates. Liquid in the fluidized bed was characterized as liquid trapped in agglomerates, free moisture, and vapour. As agglomerates break, entrapped liquid is released as free moisture where it evaporates. The liquid trapped in agglomerates had no impact on the capacitance measurements, so the capacitance was found to depend solely on the free moisture in the fluidized bed. The capacitance was correlated with the free moisture by using a special nozzle that did not create agglomerates so that a given amount of injected liquid would result in the same amount of free moisture. The rate at which the capacitance changes indicates the rate at which liquid is released from agglomerates through mass balance. Various experiments were also performed to determine the effect of the nozzle location and local fluidization velocity on the rate at which liquid is released from agglomerates.
Prociw et al. (2014) used the same method as Mohagheghi et al. (2013) but with conductance instead of capacitance to test the effect of erosion of commercial-scale spray nozzles on the rate at which liquid is released from agglomerates.

Zirgachian et al. (2013) also used conductance to characterize the effectiveness of commercial-scale Fluid Coker nozzles using square plate electrodes. These plate electrodes were found to be advantageous when compared to rod electrodes that had previously been used by Farkhondehkavaki (2012), as they did not interfere with the hydrodynamics of the fluidized bed and gave measurements that were less localized.

1.4.2 Binder Solutions

To accurately study the effect of changing process parameters on the agglomerates produced, binder solutions could be used. After the binder solution is injected in the fluidized bed, the liquid evaporates and the binder forms agglomerates. These agglomerates are then collected and their mass and size distribution are determined. The liquid content in the agglomerates can be determined by dissolving the agglomerates in a solvent and measuring its concentration in the new solution. House et al. (2008) used sugar as a binder dissolved in water to form agglomerates. After the solution was injected, the fluidized bed was defluidized and dried slowly to minimize the destruction of agglomerates after injection.

Morales (2013) used a solution of Plexiglas™ dissolved in pentane and acetone to obtain agglomerates with properties similar to those found in Fluid Cokers. The effects of the type of nozzle and the Gas to Liquid Ratio (GLR) were studied using the developed Plexiglas™ solution.

Pardo (2015) also produced agglomerates with properties similar to those found in Fluid Cokers. A solution of gum arabic dissolved in water was used to form agglomerates. A dye was added to the solution so that the initial liquid content of the agglomerates could easily be determined. The
effects of fluidization velocity and temperature on the liquid distribution in a fluidized bed were investigated.

1.5 Objectives of the Research

This thesis focuses on the understanding of interacting spray jets in a fluidized bed to improve the performance of an industrial Fluid Coker. Various interacting spray nozzle configurations were used to determine the effect of interacting spray jets on the liquid distribution in a fluidized bed. The effect of interactions on the performance of an industrial Fluid Coker was estimated by utilizing a method developed by Sanchez (2013) which combined the quantity of agglomerates produced with their initial liquid content to determine the mass of liquid that would remain in agglomerates, after a certain time under Fluid Coker conditions. To obtain these values, the conductivity of the fluidized bed was measured to quickly select the most promising configurations. An experimental method developed by Pardo (2015) to simulate the complex phenomena that occur in a Fluid Coker, was used to determine the mass of agglomerates produced and their initial liquid content. The interactions between spray jets were divided into three sections:

**Chapter 3**: The effects of interactions of two horizontally opposing spray jets on the liquid distribution in a Fluid Coker were determined by applying the abovementioned methods. The effect of varying the size of one of the spray nozzles on the liquid distribution in a Fluid Coker was also investigated.

**Chapter 4**: Given the fact that spray nozzles in a Fluid Coker are parallel and oriented above each other, the effects of varying the penetration distances of the spray nozzles on the liquid distribution were studied to propose a simple modification to the current setup used in a Fluid Coker.
Chapter 5: The effects of spray nozzle inclination of one of the spray nozzles on the liquid
distribution in a Fluid Coker were studied to determine whether an inclined spray nozzle could
improve the performance of a Fluid Coker. This was accomplished for two spray nozzles on the
same and opposite sides of the fluidized bed.
Chapter 2

2 Experimental Setup and Methodology

Experiments were performed in a large scale fluidized bed (Figure 2.1) to determine the effect of nozzle position on the liquid distribution in a Fluid Coker. The fluidized bed had a cross section of 1.22 m x 0.15 m with an expansion zone of a 1.22 m x 0.47 m cross section. Expansion zones are often employed in fluidized beds to disengage most of the solids entrained by the gas exiting the fluidized bed, by reducing the superficial gas velocity.

Figure 2.1: Fluidized bed schematic diagram incorporating an electrode
Liquid was injected into the fluidized bed with a spray nozzle that used nitrogen gas to atomize the liquid into small droplets. The spray nozzle was a scaled-down version of an industrial spray nozzle used in the Fluid Coking Process: it is a variation of the TEB spray nozzle, named after Terence Edward Base, who contributed to the original design (Base et al. 1999). Two sizes of TEB spray nozzles were employed in this study: 2.2 and 2.7 mm (Figure 2.2).

![Figure 2.2: Cross section of the TEB spray nozzles used in this study (Morales, C 2013)](image)

The TEB spray nozzle creates a jet of small droplets by incorporating a convergent zone, a gradual divergent zone, and another convergent zone. The first convergent zone accelerates the gas-liquid flow, which induces shear and elongation mechanisms resulting in droplet size reduction. The divergent zone decelerates the gas-liquid flow so it can be accelerated in the second convergent zone which further reduces the droplet size (Base et al. 1999). These nozzles are designed to be “rodable”, i.e. eventual plugging of the nozzle by coke deposits can be cleared with a metal rod.

The atomizing nitrogen gas and liquid were premixed upstream of the TEB spray nozzle by using a Bilateral Flow Conditioner (BFC) pre-mixer. The liquid contacts the atomizing gas at a
A detailed Piping and Instrumentation Diagram of the liquid injection assembly and fluidized bed are shown in Figure 2.4. 150 kg of silica sand with a density of 2650 kg/m$^3$ and Sauter mean diameter of 190 μm (Figure 2.5 shows the size distribution) were used as the solid phase of the fluidized bed. Air with a relative humidity of 12% was introduced in the windbox through two separate sonic nozzles, which ensured the flowrate was constant, to fluidize the bed. Most of the entrained sand was returned to the bed by two cyclones through a dipleg. Compressed nitrogen was used to pressurize the liquid tank and as atomization gas. A sonic nozzle was installed upstream of the BFC premixer to ensure the gas flow was sonic and constant. Another sonic nozzle was used upstream of the premixer to restrict the flow of liquid so that the required flowrate could be achieved. The two-phase mixture was introduced into the fluidized bed using either a 2.2 or 2.7 mm TEB spray nozzle. Three pressure transducers were used to set the required pressures so that a liquid flowrate of 30 g/s and a Gas to Liquid Ratio (GLR) of 2 wt% were obtained. Since two TEB spray nozzles were used in this study, the liquid injection setup was duplicated for a second spray nozzle.
Figure 2.4: Piping and Instrumentation diagram of the fluidized bed and TEB spray nozzle system for one spray nozzle
Figure 2.5: Particle size distribution of the silica sand used
2.1 Measurement Methods used to Characterize the Liquid-Solid Agglomerates

The liquid distribution in the fluidized bed was characterized with three methods:

1) Bed conductance measurements were performed for preliminary screening of experiments at room temperature.

2) A cold simulation model which incorporated the use of a binder solution to obtain actual agglomerates was used for the base and best cases (as determined by the conductance measurements) to study the properties of the agglomerates formed: size distribution and liquid content.

3) A simplified model used the results from the cold simulation model to predict the impact of initial wet agglomerate properties on the reaction time in a Fluid Coker. This model, developed by Sanchez (2013), showed the effect of varying spray nozzle configurations on the performance of a Fluid Coker.

2.1.1 Bed Conductance Measurements for Preliminary Screening

The use of bed conductance measurements relies on the principle that the conductivity of the fluidized bed increases with increasing liquid content in the fluidized bed. Figure 2.6 shows the circuit diagram for the conductance measurements. In this circuit, a function generator supplied an AC current to a 100 kΩ resistor which was wired to an electrode that was installed on the inside of the fluidized bed. The electrode was a stainless steel plate that was insulated from the fluidized bed wall since the metallic fluidized bed wall was grounded by the function generator and its anchors. The voltage across the resistor changed due to variations in fluidized bed conductivity and this signal was acquired using a Data Acquisition (DAQ) system.
Prociw et al. (2013) developed a method for analyzing the conductance of a fluidized bed and determining the liquid content. In their experimental setup, 24 electrodes were installed in a large pie-shaped fluidized bed where the signal from each electrode was acquired. Prociw et al. (2013) applied a method developed by Mohagheghi (2014) for capacitance measurements to conductance measurements. In this method, liquid is in one of three phases in the fluidized bed: trapped in agglomerates, free moisture, or vapour (Figure 2.7). Mohagheghi (2014) found that electrodes only detected free moisture since this liquid phase was distributed on the surface of bed particles and not trapped in agglomerates.
Mohagheghi (2014) found that most of the liquid was trapped in agglomerates at the end of liquid injection, with some liquid in the form of free moisture. After the end of liquid injection, agglomerates gradually broke so that water trapped in agglomerates gradually was freed in the form of free moisture. Some of the free moisture evaporated and exited the fluidized bed with the fluidizing gas.

The method developed by Prociw et al. (2013) required time-consuming calibrations that determined the relationship between the conductance and the free moisture. This was done by injecting a known quantity of liquid through a nozzle that used a high GLR to minimize the formation of agglomerates. In this way, the mass of injected liquid corresponded to the mass of free moisture in the fluidized bed. Since ZirGachian et al. (2013) showed that the bed conductance was linearly related to its free moisture, a calibration curve could be created to determine the free moisture, $X$, in the fluidized bed from the measured conductance, $y(t)$.

$$y(t) = aX + b$$  (2.1)

Prociw et al. (2013) then calculated the evaporation rate assuming the exiting vapour was saturated with liquid, which was verified by ZirGachian et al. (2013).
\[ f_e = \frac{dm_e}{dt} = v_g A_{\text{bed}} \left( H^* - H_{in} \right) \] \hfill (2.2)

where \( m_e \) is the mass of evaporated liquid, \( v_g \) is the superficial gas velocity, \( A_{\text{bed}} \) is the cross sectional area of the bed, and \( H^* \) and \( H_{in} \) are the humidity of the saturated vapour leaving the bed and the fluidization gas respectively.

The rate at which moisture was released from agglomerates was determined using a mass balance. Since free moisture was created by the evolution of liquid from agglomerates and was consumed by evaporation, equation 2.3 results:

\[ \frac{dX}{dt} = \frac{dX_{agg}}{dt} - E \] \hfill (2.3)

where \( E \) is the ratio of \( f_e \) to \( M_s \).

\[ \int \frac{dX_{agg}}{dt} = \int \left( \frac{dX}{dt} + E \right) dt \] \hfill (2.4)

Hamidi (2015), Mohagheghi (2014), and Prociw et al. (2013) all found that the cumulative amount of liquid released from agglomerates increased logarithmically with time. With this information, the gradual decrease with time of the amount of liquid trapped in agglomerates could be fitted by:

\[ \frac{M_s X_{agg}}{M_L} = 1 - e^{-\frac{\cdot}{\tau}} \] \hfill (2.5)
where $X_{agg}$ is the amount of liquid released from agglomerates divided by the bed mass, $M_L$ is the mass of injected liquid, and $\tau$ is the agglomerate breakup time constant. The agglomerate breakup time constant, $\tau$, is defined as the time required for 62.5% of the moisture initially trapped in agglomerates to be released in the fluidized bed as free moisture. A smaller value for $\tau$ means the agglomerates break faster and thus the overall performance is increased; therefore, a smaller value for $\tau$ is desired.

2.1.1.1 New Conductance Method

A new conductance measurement method was introduced to eliminate the time-consuming steps required for the method developed by Farkhondehkavaki (2012) and Prociw et al. (2013):

1) The response of each electrode has to be calibrated with separate experiments in which water is injected with special nozzles, with a very large atomization gas flowrate, to ensure that all the injected liquid is free moisture.

2) Before each experiment, the bed temperature had to be adjusted so that it was always the same.

The developed method is significantly simpler to use compared to other conductance methods, since it does not require separate calibration runs or bed temperature adjustments, as will be further discussed. Although the conductance equipment used in this thesis is relatively crude, since the conductance was only used for a rough preliminary selection of interesting cases, it is believed that this new method when applied to equipment with more sophisticated electrodes, as in the studies by Farkhondehkavaki (2012) and Prociw et al. (2013), will improve the accuracy of the conductance results.
In this measurement technique, the liquid used was water at room temperature with a flowrate of 30 g/s and a total mass of 150 g in each liquid tank, since Hamidi (2015) found that bogging occurred when the liquid content of the bed increased above 0.24 wt%. To add a safety margin, a total of 300 g of liquid were used (150 g in each tank) so that the ratio of the mass of liquid injected to the mass of bed solids was 0.2 wt%. The GLR of the atomization nitrogen used was 2 wt% and was not varied with experiments. As outlined in the Piping and Instrumentation Diagram in Figure 2.4, the temperature of the fluidized bed was acquired and for the purpose of the conductance measurements the bed temperature was always set at 25 °C, using the preheaters, at the start of liquid injection.

Prior to the liquid being injected, the conductance fluctuated around an average value; this is referred to as the dry conductance or $y_{dry}$. At the end of liquid injection, the conductance increased and a curve similar to that in Figure 2.8 resulted. The conductance increased due to the breakage of agglomerates and the formation of free liquid. Evaporation also occurred prior to the peak; however, the breakage of agglomerates dominated. After the peak was formed, evaporation dominated and thus the conductance decreased to the dry value.
Figure 2.8: Conductance vs. time of a typical water injection, bed temperature at start of injection: 25 °C

The developed method used the measured bed conductance to obtain the agglomerate breakup constant, $\tau$. The main assumptions of this method are:

1. The evaporation rate for individual experiments was constant since the bed temperature did not fluctuate dramatically, Figure 2.9.

2. The evaporating liquid was saturated and thus the relative humidity was 100% (ZirGachian et al. 2013; Prociw et al. 2013)
Figure 2.9: Temperature drop and recovery due to the injected liquid decreasing the bed temperature

Mohagheghi et al. (2013) showed that at the end of injection, \( t=0 \), some free moisture is present. Therefore, the fraction of moisture that can be freed from agglomerates corresponds to:

\[
X_{agg} = \frac{M_L}{M_x} - X_0
\]  

(2.6)

where evaporation during injection is neglected since most of the moisture is trapped in agglomerates. Therefore the cumulative fraction of moisture trapped within agglomerates can be represented by:
\[
\frac{X_{agg}}{X_{agg_0}} = 1 - e^{-\frac{-t}{\tau}} 
\]

(2.7)

The derivative of this equation is required to substitute into equation 2.4:

\[
\frac{dX_{agg}}{dt} = \frac{X_{agg_0}}{\tau} e^{-\frac{-t}{\tau}} 
\]

(2.8)

\[
\int_{t_i}^{t} \frac{dX}{dt} dt = \int_{t_i}^{t} \left( \frac{X_{agg_0}}{\tau} e^{-\frac{-t}{\tau}} - E \right) dt 
\]

(2.9)

After integration, the cumulative free moisture can be determined as:

\[
X - X_i = -X_{agg_0} \left( e^{-\frac{-t}{\tau}} - e^{-\frac{-t_i}{\tau}} \right) - E (t - t_i) 
\]

(2.10)

By choosing \(t = t_i = 0\), corresponding to the end of injection, the relation can be simplified.

\[
X = X_{agg_0} \left( 1 - e^{-\frac{-t}{\tau}} \right) - Et + X_0 
\]

(2.11)

ZirGachian et al. (2013) showed that the conductance is linearly related to the free moisture. Thus,

\[
\frac{y - b}{a} = X_{agg_0} \left( 1 - e^{-\frac{-t}{\tau}} \right) - Et + X_0 
\]

(2.12)
When there is no free moisture in the bed, \( X=0 \), \( b \) corresponds to the dry bed conductance, \( y_{\text{dry}} \). Therefore,

\[
G = y - y_{\text{dry}} = a \left[ X_{\text{agg}} \left( 1 - e^{-\frac{t}{\tau}} \right) - Et + X_0 \right]
\]  

(2.13)

which can be expanded to:

\[
G = a \left[ \left( \frac{M_L}{M_s} - X_0 \right) \left( 1 - e^{-\frac{t}{\tau}} \right) - Et + X_0 \right]
\]  

(2.14)

Replacing the ratio of \( M_L \) to \( M_s \) with \( X_{\text{all}} \) and defining the following variables:

\[
f_e = \frac{M_L}{t_{ek}}
\]  

(2.15)

\[
E = \frac{f_e}{M_s} = \frac{M_L}{M_s t_{ek}} = \frac{X_{\text{all}}}{t_{ek}}
\]  

(2.16)

\[
\gamma_0 = \frac{X_0}{X_{\text{all}}}
\]  

(2.17)

\[
\beta = a X_{\text{all}}
\]  

(2.18)

results in a more simplified equation.
\[ G = \beta \left[ 1 - (1 - \gamma_0) e^{-\frac{t}{t_{ek}}} - \frac{t}{t_{ek}} \right] \]  

Since the electrode primarily registers free moisture in its immediate area, as it is located in a certain region of the bed, a mixing factor must be added to account for this phenomenon as this local value is what is registered. To address this issue, a linear mixing model was applied that assumed there was a time delay for the free moisture to reach the electrode. Therefore, there are two cases: prior to the mixing time being achieved and subsequent to the mixing time being reached. This assumes that the electrode is segregated from the rest of the well-mixed bed.

\[ G_{local} = G \frac{t}{\Delta t_m} \]  

for \( t < \Delta t_m \)

\[ G_{local} = G \]  

for \( t \geq \Delta t_m \)

Other mixing models and their effectiveness are discussed in Appendix A.

To obtain the required parameters, four experiments were performed for each nozzle configuration that could have slightly fluctuating starting temperatures since the time required to evaporate the injected liquid \( t_{ek} \) was an adjusted parameter (Table 2.1).
Table 2.1: Summary of variables that were adjusted using the conductance method

<table>
<thead>
<tr>
<th>Variable</th>
<th>Run 1</th>
<th>Run 2</th>
<th>Run 3</th>
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Therefore the seven parameters in Table 2.1 were adjusted to minimize the total error between the four experiments of the predicted $G_{local}$ and the measured $G_{local}$ of each experiment.

Figure 2.10: Comparison between the actual $G_{local}$ and the predicted $G_{local}$ using the developed conductance model
Through preliminary experiments, it was found that $\gamma_0$ was essentially zero, indicating that the initial free moisture in the fluidized bed was negligible, which is why it was not an adjusted parameter.

Figure 2.11: Effect of increasing the number of replicates on the coefficient of variation

To determine the impact of the number of replicate runs on the accuracy of the estimated time constant $\tau$, 12 replicate runs were performed. For example, for 5 replicate runs, 1000 sets of 5 runs were selected at random from the 12 original runs, and the value of $\tau$ was calculated for each set; the 1000 values of $\tau$ were then used to calculate the corresponding coefficient of variation of $\tau$. This was repeated for different numbers of replicate runs.
Figure 2.11 shows that varying the number of replicate runs had a dramatic effect on the coefficient of variation as the coefficient of variation had a range from a negligible value to approximately 22%. Since the conductance measurements were meant to provide a preliminary analysis of the liquid distribution in the fluidized bed, four replicates were chosen, as this number was more reasonable. The most promising results obtained by the conductance measurements were then further investigated using the cold method discussed below.

2.1.2 Cold Simulation Method

Pardo (2015) developed a method to simulate agglomerate formation in a Fluid Coker to determine the size distribution and liquid content of the agglomerates in given size cuts. A binder solution of adjustable properties was injected into a fluidized bed of sand at 130 °C. A specific binder solution with fixed properties was employed for the purpose of this study. The liquid flowrate and GLR were the same as those used for the conductance measurement experiments: 30 g/s and 2 wt% respectively. A total mass of 1200 g of liquid was used with 600 g in each nozzle liquid tank. Two different solutions were used in each liquid tank so that the effect of each nozzle could be studied if needed. The first solution contained 6 wt% of gum arabic (the binder) and 2 wt% blue number 2 dye dissolved in water. The second solution was the same but with yellow number 5 dye being used at a concentration of 1.2 wt% instead of the blue dye. Both solutions were adjusted to a pH of 3.0 using hydrochloric acid.

At the end of injection, the fluidization velocity was reduced to minimum fluidization for 10 minutes so that the agglomerates could dry without breaking. After the bed was allowed to cool, the entire bed was sieved into nine size cuts to obtain the size distribution of the agglomerates. The agglomerates were characterized as either macroagglomerates or microagglomerates where macroagglomerates are large agglomerates that are larger than any bed particle (greater than 600 μm), whereas microagglomerates are smaller agglomerates that are within the size range of the larger bed particles (less than 600 μm). The agglomerates from each size cut were dissolved in
water and centrifuged to determine the quantity of dye in each size cut. This was done by using a spectrophotometer to determine the absorbance at 630 μm (wavelength of blue light) and 427 μm (wavelength of yellow light). Since the agglomerates were not allowed to break after the end of injection, the resulting Liquid to Solid (L/S) mass ratio corresponded to the initial value at the end of injection.

For the size cuts that had microagglomerates, the size of the agglomerates approached that of the larger particles of sand originally in the bed. Therefore a representative sample of the size cuts containing microagglomerates was taken, left in water for the gum to dissolve, and a HELOS Particle Size Analyzer (PSA) was used to determine what fraction of the size cuts were fines, from the original sand, so that the mass of actual agglomerates could be determined. This method is based on measurements from Pardo (2015) that demonstrated that size distribution of the bed particles trapped in agglomerates is the same as the size distribution of the bed particles. The mass of particles trapped in a given size cut of microagglomerates in the sample was determined by knowing the total mass of the sand in the fluidized bed, and the fraction of fines in the sample ($x_f$) and in the original bed mass ($x_{fbed}$):

$$m_p = M_s \frac{x_f}{x_{fbed}}$$

(2.22)

The mass of agglomerates in the sample was then determined for each size cut by knowing the mass of binder and dye, which were found from the analysis used to determine the initial L/S ratio:

$$m_{\mu agg,Ri} = m_p + m_{GA} + m_{dye}$$

(2.23)

Finally, the mass of microagglomerates in a given size cut in the bed $m_{\mu agg,i}$, was calculated given equation 2.24
\[ m_{\text{agg},j} = m_{\text{agg},R} \frac{m_{<600\mu m}}{m_R} \]  

(2.24)

where \( m_{<600\mu m} \) is the mass of solids in the bed less than 600 μm and \( m_R \) is the total mass of the representative sample taken.

### 2.1.3 Simplified Model for Impact of Initial Wet Agglomerate Properties on the Reaction Time in a Fluid Coker

A major issue with Fluid Cokers is that some wet agglomerates reach the stripper section, where they cause stripper fouling (Section 1.2). Therefore, it is important to predict how long an agglomerate of a given size and original liquid content would take to completely react and dry.

For the purpose of this study, the initial L/S ratio and agglomerate size were used to determine the time for full conversion of each size cut, using the model developed by Sanchez (2013). Crank’s equations regarding diffusion through a sphere were used to consider the limitations of the agglomerate size (Crank, J. 1975). The following assumptions were made by Sanchez (2013):

- Thermal cracking does not start until a critical temperature is reached, and then occurs instantaneously (this temperature is the temperature of the reaction front: Figure 2.12)
- External heat transfer is assumed to be non-rate limiting so that the surface temperature is essentially equal to the bed temperature
- Heat transfer through the particle is rate-limiting: thermal cracking reactions in the agglomerates are only limited by conduction heat transfer from the surface of the agglomerate to the reaction front
- Initially, the liquid is uniformly distributed throughout the agglomerate (Figure 2.12)
• The heat capacity of coke is neglected, i.e. the heat required to heat the agglomerates is negligible when compared to the heat of reaction.

• Bed temperature is 550 °C

A model developed by House (2007) was used to determine the reaction front temperature, which was found to be 520 °C. Initially, the liquid is distributed uniformly throughout the agglomerate (t=0), and the reaction front is located on the outer surface of the agglomerate. As time progresses, the liquid reacts and the reaction front gradually moves inward resulting in the product vapors diffusing out of the agglomerate. The reaction proceeds until the agglomerate is completely dry (t=∞).

![Diagram of agglomerate behavior](image)

**Figure 2.12: Wet agglomerate behavior in a Fluid Coker (Sanchez, F 2013)**

The time for full conversion of bitumen depends on the size and original liquid content of an agglomerate. Therefore, agglomerates with a higher liquid content are not necessarily worse as can be seen by Figure 2.13. Thus, this method combines the effects of agglomerate size and liquid content on the mass of liquid remaining in agglomerates.
Figure 2.13: Effect of initial liquid content and agglomerate size on the predicted time for full conversion

The time for full conversion can be determined by:

\[ t_c = \frac{R^2 (L/S)_0}{6\Gamma} \] 

(2.25)

where R is the radius of the agglomerate and \( \Gamma \) is a constant that is independent of size and initial liquid content:

\[ \Gamma = \frac{k(T_B - T_R)}{\rho_i \Delta H} \] 

(2.26)
In equation 2.26, \( k \) is the thermal conductivity of coke layers, \( T_B \) is the bed temperature, \( T_R \) is the temperature of the reaction front, \( \rho_s \) is the density of the coke, and \( \Delta H \) is the change in enthalpy associated with the reaction of the liquid oil to vapors, permanent gases and coke.

Once the time for full conversion was determined, the normalized radial position, \( \eta \), of the reaction front was solved for.

\[
(1 - \eta)(1 + \eta - 2\eta^2) = \frac{t}{t_c} \tag{2.27}
\]

The mass of liquid remaining in agglomerates, \( m_{Lr} \), could then be determined using:

\[
\frac{m_{Lr}}{M_L} = \eta^3 \tag{2.28}
\]

where, \( M_L \) is mass of injected liquid.

Then, the mass of liquid remaining in the agglomerates could be plotted vs. time to obtain a profile describing the evolution of liquid from agglomerates in the Fluid Coking Process.

Thus, three methods were utilized to characterize the liquid distribution in a fluidized bed: conductance measurements, a cold simulation method, and a simplified method to estimate the effect of nozzle configuration on the impact of the liquid distribution on Fluid Coker operations.
Chapter 3

3 Effect of Interactions of Horizontal Spray Jets on the Liquid Distribution in a Fluidized Bed

The Fluid Coking Process supplies approximately 20% of Canada’s finished petroleum products (ExxonMobil n.d.); therefore, it would be reasonable to improve the operation of a Fluid Coker. One way to accomplish this would be to improve the liquid distribution in the Fluid Coker, as this reduces heat and mass transfer limitations (House, 2007). In industrial Fluid Cokers, bitumen is sprayed through spray nozzles that are arranged throughout the height of the Fluid Coker. The spray jets do not interact horizontally in existing Fluid Cokers. If interacting horizontal spray jets could improve the liquid distribution in a Fluid Coker, the conversion of bitumen feed to valuable products would be increased, and the Fluid Coker operability would also be improved. Simple modifications to the spray nozzle assembly could be made to incorporate spray jet interactions without jeopardizing the integrity of the body of the Fluid Coker. For example, the nozzle penetration could be increased to promote spray jet interactions. In spite of their potential benefits, interactions of horizontal gas-liquid spray jets have not been studied before.

Interacting spray jets could also reduce the size of agglomerates produced due to more turbulent regimes being formed when spray jets meet. Currently, the size of agglomerates in a Fluid Coker is controlled by the use of attrition jets which incorporate a high velocity gas jet that breaks agglomerates. Thus, if interacting spray jets could reduce the size of agglomerates produced, the excess gas that would have been used by the attrition jets could be used to increase the Gas to Liquid Ratio (GLR), which has been found to greatly improve the liquid distribution in a Fluid Coker (Morales, C 2013; House et al. 2008).
Finally, understanding how spray jet interactions affect the liquid distribution can help better understand the mechanisms through which a single spray nozzle distributes liquid in a fluidized bed. This could result in improvements of the operation of non-interacting nozzles in Fluid Cokers.

3.1 Experimental Setup and Methodology

In the fluidized bed that was used, as shown in Chapter 2 (Figure 2.1), two spray nozzles were installed in ports b and d so that they were co-axial and horizontally opposite as seen in Figure 3.1. One of the spray nozzles was fixed in position, while the other spray nozzle had a variable penetration into the fluidized bed so that the distance between spray nozzles could be varied. First, two 2.2 mm TEB spray nozzles were used to determine the effect of two interacting spray nozzles of the same size on the liquid distribution in a fluidized bed. The 2.2 mm TEB spray nozzle in port b was then changed to a 2.7 mm TEB spray nozzle, while still using a 2.2 mm TEB spray nozzle in port d, so that the effect of varying the spray nozzle size and distance between spray nozzles could be studied.

Figure 3.1: Two horizontally opposite spray jets with one fixed nozzle and one moveable nozzle ($66.4 \leq \Delta x_n \leq 107$ cm, $\Delta z$: 0 cm)
3.2 Results and Discussion

In the present work, the effect of interacting spray jets on the liquid distribution in a fluidized bed was determined. This was accomplished by using two spray nozzles of the same and different sizes. The main objective of this study was to determine the effect of nozzle interactions on the mass of liquid remaining in an industrial Fluid Coker. This was achieved by the use of a simplified model developed by Sanchez et al. 2013 to determine the liquid content in agglomerates produced in an industrial Fluid Coker based on initial values determined in a smaller scale fluidized bed. To obtain the initial liquid content in agglomerates produced in a fluidized bed, a binder solution of gum arabic dissolved in water was injected into a fluidized bed of silica sand particles so actual agglomerates could be produced. This process was time consuming and labour intensive, so a preliminary screening method was applied to quickly select the most promising distances between interacting spray nozzles, by estimating the effectiveness of the liquid distribution. The method that was utilized for preliminary screening was that of Section 2.1.1.1. The most promising results from this conductance method were validated using the gum arabic solution.

3.2.1 Preliminary Selection using the Conductance Measurement Method

From Figure 3.2, the normalized agglomerate breakup time constant was used to describe the liquid distribution in the fluidized bed. The normalized agglomerate breakup time constant is the ratio of the time constant obtained with interacting nozzles to the time constant obtained with non-interacting nozzles. A lower value of the normalized agglomerate breakup constant is desired, as it would indicate that the liquid trapped in agglomerates is released more quickly. In Figure 3.2, negative values in the distance $\Delta x$ between jet tips correspond to overlapping jets.
When the jets are far apart, the performance is that of the non-interacting spray jets. Even when the spray jets are brought closer together until they just merge, the performance is similar to that of the non-interacting case. The performance is greatly improved when the jets overlap by 5 cm, which is due to the unstable merged jet which means that the frequency of the jets expansion-contraction cycle increases and the liquid distribution is improved due to more solids being contacted, as will be discussed below. The merged jets also release a bubble that contains wet solids in its wake, which further improves the liquid distribution in the fluidized bed. When the jets further overlap, no improvement results due to beneficial splashing of liquid in all directions, as a result of the increased degree of merging. The drawback of having highly merging spray jets is that their penetration length becomes similar to that of their minimum penetration lengths, and thus a single merged jet could form which does not allow for wetted solids to be replenished by dry solids. Thus, the splashing of liquid negates the poor liquid distribution that would result from highly merging spray jets.
To understand the effect of separation between spray jets on the expansion time, a model similar to that developed by Mohagheghi (2014) for single spray jets was adapted to merging spray jets. In this model, bed bubbles carrying dry solids are carried to the jet where they are captured and result in an unstable jet cavity. Due to the instability, a bubble is released from the jet and the jet contracts to a minimum value. The jet expands once more, and the cycle continues (Figure 3.3). A faster jet expansion results in a bubble carrying fewer wet solids, and thus the liquid is distributed more uniformly in the fluidized bed.
To estimate the released bubble diameter, equation 3.1 was used by Mohagheghi (2014).

\[ d_\mu = \frac{L_{jet}}{2.88} \]  

(3.1)

where \( L_{jet} \) is the average spray jet penetration length as predicted by Arriyapadi et al. (2003).

Since the length of the spray jet varies due to fluctuations in the fluidized bed, a maximum and minimum penetration length can be defined. The maximum spray jet penetration length was estimated by Mohagheghi (2014), using a correlation from Xuereb et al. (1991):
\[ L_{\text{max}} = 1.23 L_{\text{jet}} \] \hspace{1cm} (3.2)

If the section of the jet cavity between \( L_{\text{min}} \) and \( L_{\text{max}} \) is assumed to be cylindrical (Figure 3.3), the average jet length \( L_{\text{jet}} \) is the arithmetic average of \( L_{\text{min}} \) and \( L_{\text{max}} \) and

\[ L_{\text{min}} = \frac{L_{\text{jet}}}{1.3} \] \hspace{1cm} (3.3)

Therefore, if one assumes that the jet cavity is cylindrical between \( L_{\text{min}} \) and \( L_{\text{max}} \), the volume of the released bubble is:

\[ v_B = \frac{\pi}{6} d_B^3 = \frac{\pi}{4} D^2 (L_{\text{max}} - L_{\text{min}}) \] \hspace{1cm} (3.4)

where the diameter of the jet is given by:

\[ D = d_N + 2(\tan \theta) L_{\text{min}} \] \hspace{1cm} (3.5)

where \( d_N \) is the spray nozzle diameter and \( \theta \) is the half angle of the spray jet.

Mohagheghi (2014) obtained:

\[ v_B = \frac{\pi}{4} D^2 (L_{\text{max}} - L_{\text{min}}) = Q_g t_{\text{exp}} \] \hspace{1cm} (3.6)

by assuming that \( Q_g \), the volumetric flowrate of gas entering the spray jet cavity was constant over the whole expansion cycle time, \( t_{\text{exp}} \) of the spray jet.
However, the volumetric flowrate of gas entering the spray jet cavity is not constant and increases as the jet expands. The gas entering the jet cavity is comprised of atomization gas, \( Q_a \), which is constant, and bubbles from the bed, \( Q_{cb} \), which changes as the jet expands and the jet length, \( L \), increases, as shown by the equation below.

\[
Q_{cb} = D(L - L_{bcs})(v_g - u_{mf})
\]  
\[ (3.7) \]

Following Mohagheghi (2014), this assumes that only the part of the jet cavity that extends from \( L_{bcs} \) to the jet tip captures gas bubbles. \( L_{bcs} \) is the distance from the nozzle tip to the start of bubble capture zone and \( L \) is jet length, which varies with time. This equation assumes that diameter of the jet section between \( L_{bcs} \) and \( L_{min} \) is \( D \): this is not an unreasonable approximation as \( L_{bcs} \) is not much smaller than \( L_{min} \), and, in this region, a bubble can be captured even if its center is not exactly within the jet cavity. Mohagheghi (2014) demonstrated through experimental results that \( L_{bcs} \) was approximately half of the predicted jet penetration length, in this case approximately 20 cm.

\[
\frac{\pi}{4} D^2 dL = \left[ D(L - L_{bcs})(v_g - u_{mf}) + Q_a \right] dt
\]  
\[ (3.8) \]

To simplify the integral a variable, \( \lambda \), was introduced to combine constant terms.

\[
\lambda = L - L_{bcs} + \frac{Q_a}{D(v_g - u_{mf})}
\]  
\[ (3.9) \]

Thus, the expansion cycle time was calculated as

\[
t_{exp} = \frac{\pi D}{4(v_g - u_{mf})} \ln \left( \frac{\lambda_{max}}{\lambda_{min}} \right)
\]  
\[ (3.10) \]
For non-interacting spray jets, the predicted expansion cycle time was approximately triple (0.62 s) that of slightly merging spray jets (0.20 s), as shown in Figure 3.5 because the bubble diameter for merging and non-merging jets is assumed to be the same. Thus, merged jets contract less and require less time to expand. Nonetheless, slightly merging jets result in an improved liquid distribution as can be seen by the τ estimates measurements in Figure 3.2. The liquid content of the agglomerates can be determined by knowing the mass of liquid and solids that are wetted with each expansion

\[ c = \frac{L}{S} \quad (3.11) \]

where S is the mass of solids wetted in one expansion and L is the mass of liquid wetting solids in one expansion.

The mass of bubbles trapped in the wake of the released bubble is

\[ S_w = f_w \rho_{mf} \frac{\pi}{6} d_B^3 \quad (3.12) \]

where \( \rho_{mf} \) is the bulk density of the solids at minimum fluidization and \( f_w \) is the wake fraction: the amount of solids in the bubble.

The mass of solids wetted in one jet expansion can be assumed to be equal to the sum of the mass of bed solids displaced by one jet expansion and the total mass of solids in the wakes of the captured bubbles.

\[ S = \rho_{mf} \frac{\pi}{4} D^2 (L_{\text{max}} - L_{\text{min}}) + \rho_{mf} f_w (v_B - Q_{\text{exp}}) \quad (3.13) \]

Given the liquid flowrate, \( F_L \), \( L \) can be determined.
\[ L = F_l / T_{\text{exp}} \]  

(3.14)

For multiple jet expansions, the liquid content is calculated based on the previous liquid content.

\[ c_n = c_{n-1} \left(1 - \frac{S_R}{S}\right) + c_i \]  

(3.15)

This model is incorporated in Figure 3.4. At steady state, a decreased liquid to solid ratio is observed irrespective of time for slightly merging jets compared to a single jet. Therefore, a lower expansion cycle time means the liquid would be better distributed in the fluidized bed because less liquid is released in the bubble per expansion; hence the liquid is distributed to other parts of the bed in an improved manner.

![Figure 3.4: Liquid to Solid Ratio of a single jet and a merging jet](image-url)
As the degree of merging of the spray jets increased, the predicted expansion cycle time gradually increased, which resulted in a deteriorated performance; however, the expansion time did not increase to the non-interacting case as predicted by the conductance measurements. This is due to the fact that the expansion model has limitations which do not fully predict the interactions between spray jets with a greater degree of merging. The model predicts that the sprayed liquid is uniformly mixed in the entrained and released solids, which is not the case in an actual fluidized bed. As well, the bubble that is released is assumed to be of the same size irrespective of the degree of merging of spray jets, which may be the case when the spray jets are slightly merging. When the spray jets merge at a greater degree, less time is allowed for wetted solids to be replenished by dry solids from the surrounding bed, and the solids become rewetted. At the same time, the volume of wetted solids would be less for highly merging spray jets because of a single jet forming and thus fewer solids are entrained into the jets.

![Figure 3.5: Effect of horizontal distance between spray jets on the expansion time of two merging spray jets produced by 2.2 mm TEB spray nozzles](image-url)
According to the conductance measurements, an increased degree of merging resulted in a performance similar to that of the non-interacting case due to less time being available for wetted solids to be replenished by dry solids. This was canceled out by liquid splashing in different directions due to the increased degree of merging. At the same time, a stable merged jet forms which would not allow for contraction and expansion, which results in poor liquid distribution as wetted solids are not replenished by dry solids. This is mainly due to the fact that the spray jets are merging at the minimum jet penetration length, leading to a mostly stagnant spray jet. Therefore, an improved performance results when two spray jets merge near their tips; however, no improvement results when the degree of merging is increased due to a poorer liquid distribution.

The next spray nozzle configuration that was tested involved the use of two spray nozzles of different sizes, with the conductance measurements summarized in Figure 3.6. The dashed line signifies the fact that the agglomerate breakup time constant remains constant for larger distances between non-merging jets as was found with the use of two spray nozzles of the same size earlier. To ensure the results for two nozzles of the same and different sizes could be easily compared, $\tau_{\text{non-interacting}}$ of the two nozzles of the same size was used for both cases. With varying distances between spray jets, no significant improvement results: the performance is either the same or is significantly decreased. This suggests that the use of interacting spray jets of different size greatly reduces the effectiveness of distributing liquid in a fluidized bed.
Figure 3.6: Agglomerate breakup time constant for two horizontally opposing synchronized spray nozzles of different sizes: 2.2 mm and 2.7 mm TEB spray nozzles

One possible explanation for this phenomenon is due to the fact that the frequency of expansion of the two jets is different since they are of different sizes (0.62 seconds for a single 2.2 mm nozzle, and 0.72 seconds for a single 2.7 mm nozzle).

Since the two jets expand at a different rate, this could be why the performance is completely different than when two nozzles of the same size were studied. To determine whether synchronization between the interacting jets is important, a time lag was introduced between the jets of two 2.2 mm spray nozzles. The expansion cycle time of a single 2.2 mm spray jet was found to be 0.62 seconds, so the initial injection of one of the spray jets was delayed by one second so that the spray jets would be unsynchronized. With a delay of one second, between one and two cycles could be completed, thereby limiting any error resulting from the initial expansion of the spray jet. This was
completed for a distance between spray jets of 3.6 cm, which corresponded to the best case as observed in Figure 3.7.

![Graph](image)

**Figure 3.7: Agglomerate breakup time constant for two horizontally opposing 2.2 mm TEB spray nozzles**

For the same position as the best case for two spray nozzles of the same size with a delayed start time in injection of one of the spray jets, the performance was significantly worse due to the jets being unsynchronized. The value of the normalized agglomerate breakup time constant when the jets were unsynchronized was similar to that of the case where two spray nozzles of different sizes were used at other distances. Thus, the synchronization of spray jets is important to achieve improved performance in a Fluid Coker.
In an industrial Fluid Coker, ensuring spray jets are synchronized could prove to be challenging due to the substantial variations that occur in a fluidized bed of that scale, and the fact that the reactor must run continuously over the course of three to five years. Thus, feedback controllers could be implemented to trim the GLR to synchronize the expansion of the spray jets.

3.2.2 Agglomerate Analysis using Gum Arabic Solution

To better understand the effect of spray jet interactions on the liquid distribution in a fluidized bed, the gum arabic solution described in Section 2.1.2 was utilized. First, experiments were performed with the base case of non-interacting jets and with the best case, according to the conductance measurements, for merged jets.

The size distribution of the agglomerates produced for two synchronized spray nozzles of the same size is shown in Figure 3.8. Due to an improved liquid distribution by the merged spray jets, weaker agglomerates were formed which could easily be broken in the fluidized bed when compared to the non-interacting base case. The initial Liquid to Solid (L/S) ratio in Figure 3.9 describes a similar outcome: the agglomerates produced by the best case are much drier due to the spreading of liquid by the released bubbles and the faster expansion times of the spray jets. When the spray jet expands faster, the L/S ratio is decreased, as found in Figure 3.4, due to the released bubbles spreading liquid throughout the bed at a faster rate. These results also confirm those that were achieved using the conductance measurements, as an overlap between spray jets of 3.6 cm greatly improved the liquid distribution in the fluidized bed.
Figure 3.8: Effect of horizontal position on the mass of agglomerates produced for two horizontally opposing 2.2 mm TEB spray nozzles.
Figure 3.9: Effect of horizontal nozzle position on the initial liquid to solid ratio for two opposing 2.2 mm TEB spray nozzles

The agglomerate size distributions for the case where two spray nozzles of different sizes were used are shown in Figure 3.10. The quantity of agglomerates produced by two spray nozzles of different sizes that were not interacting was essentially equivalent to that of the best case for interacting spray jets also produced by spray nozzles of different sizes. The initial L/S ratio also indicates that the best case for interaction produced agglomerates with a similar wetness as those that produced by non-interacting spray jets. This also confirms the results that were obtained using the conductance measurements (Figure 3.6) that showed that, for nozzles of different sizes, the best case did not give much better results than the base case of non-interacting jets.
Figure 3.10: Effect of horizontal nozzle position on the mass of agglomerates produced for two horizontally opposing TEB spray nozzles of size 2.2 and 2.7 mm
The gum arabic solution was also used for the delayed injection with two unsynchronized spray nozzles of the same size. As seen in Figure 3.12, fewer agglomerates were actually created with unsynchronized spray jets. At first, this may seem counterintuitive since the conductance measurements suggest that more agglomerates should be produced with unsynchronized nozzles; however, these results suggest that, when the jets are not synchronized, the wet solids formed by a spray jet can be broken up by the expanding jet from the other nozzle (Figure 3.14). This interpretation is confirmed by the initial L/S results in Figure 3.13, the agglomerates produced by the unsynchronized spray jets had a higher initial liquid content. Since the spray jets were unsynchronized, the solids were rewetted by the expanding spray jets due to the solids being unable to escape the region between the jets. In the case where synchronized spray jets were used, the solids that were in the region between the two jets could be circulated to other parts of the fluidized bed during the time it took for the spray jets to expand. Thus, unsynchronized jets
produce smaller, wetter agglomerates than those produced by synchronized jets (Figure 3.14).

Figure 3.12: Effect of synchronization of spray jets on the mass of agglomerates produced
Figure 3.13: Effect of synchronization of spray jets on the initial liquid to solid ratio of agglomerates produced

Figure 3.14: Rewetting and breakage of agglomerates due to non-synchronized jet expansion
3.2.3 Simplified Fluid Coking Model

As described in Section 2.1.3, Sanchez (2013) developed a model that incorporated the effect of the agglomerates size and liquid content to determine the quantity of liquid that would remain in a Fluid Coker. Figure 3.15 illustrates the effect of synchronization on the fraction of liquid injected that would remain in agglomerates in an industrial Fluid Coker. For any given time, when compared to non-interacting jets and unsynchronized interacting jets, the best case with synchronized, interacting spray jets of the same size resulted in a smaller amount of liquid that remains trapped in agglomerates. When comparing non-interacting jets to unsynchronized interacting jets, the fraction of injected liquid remaining in agglomerates was approximately the same. Thus, synchronizing spray jets in a Fluid Coker is imperative to achieve an improved performance in the liquid distribution.

Figure 3.15: Effect of synchronization on the fraction of liquid remaining in coke agglomerates in the Fluid Coking Process using two 2.2 mm TEB spray nozzles.
Focusing on the time axis of Figure 3.15, the time required for the bitumen to be converted and evaporated in a Fluid Coker would be approximately 4.5 seconds, which seems unreasonably low. This is due to the fact that the model assumes the same scaled-down 2.2 mm spray nozzles would be used in a Fluid Coker at the operating temperature of the Fluid Coker. Therefore, to obtain truly accurate values, the agglomerates produced in a smaller-scale fluidized bed would need to be scaled-up to match those produced with commercial-scale nozzles in a Fluid Coker.

As concluded from the previous sections, varying the distance between spray jets either has no effect or is detrimental to the operation of a Fluid Coker with two spray nozzles of different sizes. Using the simplified Fluid Coking model, the fraction of injected liquid remaining in agglomerates would be approximately equal between the non-interacting case and the best case (Figure 3.16).

![Figure 3.16: Effect of nozzle interaction on the fraction of liquid remaining in coke agglomerates in the Fluid Coking Process with two nozzles of different size](image-url)
3.2.4 Correlation between $\tau$ and Initial Mass of Liquid in Agglomerates

To better understand the significance of the agglomerate breakup time constant, $\tau$, a correlation was developed between the initial fraction of injected liquid trapped in macroagglomerates and the normalized agglomerate breakup time constant (Figure 3.17). Therefore, a strong relationship exists between the two parameters. This was also confirmed by studies conducted by Weber (2009) and Parveen (2011), which showed that the stability of wet agglomerates in a fluidized bed is greatly affected by their liquid content.

![Figure 3.17: Correlation between $\tau$ and the initial mass of liquid in macroagglomerates](image)

$$y = 0.2683\ln(x) + 0.6218$$

$R^2 = 0.9766$
Chapter 4

4 Effect of Vertical Separation Between Spray Jets on the Liquid Distribution in a Fluidized Bed

The Fluid Coking Process supplies approximately 20% of Canada’s finished petroleum products (ExxonMobil n.d.). Therefore, it would be reasonable to improve the operation of Fluid Cokers. One way to accomplish this would be to improve the liquid distribution in Fluid Cokers, as this reduces heat and mass transfer limitations by forming smaller agglomerates (House, P. 2007).

In industrial Fluid Cokers, spray nozzles are distributed vertically in nozzle banks and can be staggered due to the shape of a Fluid Coker (Figure 4.1). Therefore, if incorporating different spray nozzle penetrations could improve the liquid distribution in a Fluid Coker, the performance would be improved, and consequently the yield of bitumen feed to valuable products would be increased. Thus, the effect of horizontal and vertical distances between non-merging spray jets should be understood to improve the liquid distribution in a Fluid Coker.

Although Fluid Cokers have been used for over 50 years (Pfeiffer et al. 1959), many aspects of their operation are still not well known. For example, the use of vertically separated gas-liquid spray jets have not been studied before, so the present work seeks to determine whether a positive or negative effect results from varying the horizontal and vertical distances between spray jets.
Figure 4.1: Vertical distribution of spray nozzles in a Fluid Coker (Morales, C 2013)

4.1 Experimental Setup and Methodology

A fluidized bed of silica sand was used to test the effect of spray nozzle penetration on the liquid distribution in a fluidized bed (Section 2). Spray nozzles were used to introduce liquid into the fluidized bed in the form of atomized droplets using the TEB spray nozzle design in Section 2. To compare the obtained results, a base case that was developed in Section 3 was used. In this spray nozzle configuration, two parallel, horizontally opposite TEB spray nozzles produced non-interacting spray jets, as can be seen in Figure 4.2. The first nozzle configuration that was tested incorporated two vertically separated 2.2 mm TEB spray nozzles with a movable top spray nozzle (Figure 4.3). In this way, the effect of the penetration of the top nozzle on the liquid distribution in the fluidized bed could be investigated, since the bottom spray nozzle was fixed in position.
The second configuration that was investigated also integrated two vertically separated 2.2 mm spray nozzles (Figure 4.4). The objective of the second spray nozzle configuration was to determine the effect of the penetration of the bottom spray jet on the liquid distribution, as the top spray nozzle had a fixed position.

From the first and second configurations, the effect of the penetration of either the top or bottom spray nozzle on the liquid distribution was determined; however, the spray nozzles in both configurations were installed in ports a and b of the fluidized bed shown in Section 2. Thus, a third configuration was investigated to determine the effect of the vertical distance between spray nozzles on the liquid distribution (Figure 4.5). This configuration was similar to that of the first configuration, where a movable top spray nozzle and fixed bottom spray nozzle were incorporated, with the difference being a greater vertical separation between the spray nozzles was used. In this case, the two spray nozzles were installed in ports a and c.

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**Figure 4.2: Two horizontally opposite, non-interacting spray jets used as the base case to compare the effectiveness of interactions (Δxn: 107 cm, Δz: 0 cm)**

**Figure 4.3: Two spray jets with an adjustable upper nozzle penetration length (0≤Δxn≤45 cm) with a fixed vertical distance (Δz: 20 cm)**
4.2 Results and Discussion

In the present work, the effects of interacting spray jets on the liquid distribution in a fluidized bed were determined. This was accomplished by studying the effect of spray nozzle penetration on the liquid content of agglomerates produced. The main objective of this study was to determine the effect of staggering spray nozzles on the amount of liquid remaining in agglomerates in an industrial Fluid Coker. This was achieved by the use of a simplified model developed by Sanchez (2013) to determine the liquid content in agglomerates produced in an industrial Fluid Coker, based on initial values determined in a smaller scale fluidized bed. To obtain the initial liquid content in agglomerates produced in a fluidized bed, a binder solution of gum arabic dissolved in water was injected into a fluidized bed of silica sand particles so that actual agglomerates could be produced. This process was time consuming and labour intensive, so a preliminary screening method was applied to quickly select the most promising distance between vertically separated spray nozzles, by estimating the effectiveness of the liquid distribution. The conductance method that was utilized for preliminary screening was that of Section 2.1.1.1. The most promising results from this conductance method were then validated using the gum arabic solution.
4.2.1 Effect of Top Spray Nozzle Penetration on the Agglomerate Breakup Time Constant

From Figure 4.6, the normalized agglomerate breakup time constant, obtained by taking the ratio of the measured time constant of agglomerate breakage to the time constant of agglomerate breakage for non-interacting jets, was used to describe the liquid distribution in the fluidized bed. A lower value of the normalized agglomerate breakup time constant is desired, since $\tau$ is the time required for 62.5% of the liquid trapped in agglomerates to be released.

For the case where the top spray nozzle penetration was varied (Figure 4.6), the normalized agglomerate breakup constant varied dramatically from about 0.5 to 2.3. For a horizontal distance between spray jets of 0 cm, the performance was more than twice as worse as the base case. This was due to the fact that the bottom spray jet captured fluidized bed bubbles that could otherwise have been captured by the top jet. Since the capture of fluidized bed bubbles by a spray jet decreases the expansion cycle time of the spray jet, an improved liquid distribution results, as explained in Section 3.2.1. The expansion time could also be decreased if the top spray jet captured bubbles released by the bottom spray jet, but since bubbles rise at an angle to the vertical, this was not the case (Figure 4.7). Therefore, wetter and stronger agglomerates were formed, which resulted in a decreased performance.

When the top spray jet penetrated 5 cm further into the fluidized bed, Figure 4.6 shows that the performance was still worse than that of the base case, but was better than when the spray jets were 0 cm apart. Therefore, the bottom spray jet still partially blocked the top spray jet from capturing fluidized bed bubbles, but the bubbles released by the bottom spray jet could be captured at the tip of the top spray jet. Mohagheghi (2014) found that bubbles were captured near the tips of spray jets, which is why a slight improvement is observed (Figure 4.8).

For horizontal distances of between 10 and 30 cm, a significantly improved performance was observed when compared to the base case. This was due to the fact that bubbles
released by the bottom spray jet were captured by the top spray jet, and a minimal blockage of fluidized bed bubbles occurred (Figure 4.9). The expansion cycle time of the top jet was, thus, greatly reduced.

For horizontal distances between spray jet tips of more than 35 cm, the performance became similar to that of the base case since the spray jets were no longer interacting (Figure 4.10).

Figure 4.6: Normalized agglomerate breakup time constant for two parallel 2.2 mm TEB spray nozzles, with a moveable top nozzle (configuration from Figure 4.3)
Figure 4.7: Diagram of the spray jet positions in the fluidized bed
(Δx: 0 cm, Δz: 20 cm)

Figure 4.8: Diagram of the spray jet positions in the fluidized bed
(Δx: 5 cm, Δz: 20 cm)
As previously discussed, the performance of the configuration where the top spray nozzle penetration was varied greatly depended on the bottom spray jet’s relative position to the top spray jet. Since it was proposed that the bottom spray jet starved the top spray jet of gas bubbles, a baffle was created that incorporated atomization gas to simulate the bottom spray jet (Figure 4.11). The shape of the baffle was a rectangular prism with an extruded centre to ensure bubbles were trapped in the baffle and could only rise from the tip of the baffle, which is similar to a spray jet. An atomization gas line was also attached to the baffle to attempt to simulate a spray jet.
Figure 4.11: Dimensions of the baffle used to mimic the bubble capture and lateral transfer of the lower spray jet

Conductance measurements were performed using a spray nozzle in port b and the developed baffle in port a to test the theory that the bottom spray jet starved the top spray jet of gas bubbles, and that bubbles were released from the bottom spray jet which led to an improved performance. From the normalized agglomerate breakup time constant profile in Figure 4.12, the values achieved using the baffle and moveable top spray nozzle were almost identical to those attained when two spray jets were used, with slight differences due to the fact that the baffle could not perfectly mimic a spray jet. Therefore, the bottom spray jet starved the top spray jet of gas bubbles and released bubbles that could be captured by the top spray jet, if the correct range of vertical and horizontal distances between spray jets were implemented.
Figure 4.12: Comparison of the normalized agglomerate breakup time constant of two parallel 2.2 mm TEB spray nozzles and one moveable 2.2 mm TEB spray nozzle with a baffle replacing the lower jet

The aforementioned analysis of the variations in the normalized agglomerate breakup constant could be summarized through three mechanisms:

1. The bottom spray jet captures gas bubbles that would normally have been captured by the top spray jet, increasing the expansion cycle time of the top jet, which has a negative impact on its liquid distribution performance.
2. The bottom jet periodically releases a bubble from its tip that can be captured by the top spray jet, if the top spray nozzle penetrates farther than the bottom spray nozzle. This large bubble thus greatly reduces the expansion cycle time of the top jet, which has a beneficial impact on its liquid distribution performance.
3. The large bubble released from the tip of the bottom jet can provide additional turbulence that helps break up agglomerates released from the top jet. This will be discussed further when the results of the gum arabic technique are analysed.

The combination of these three impacts determines whether the bottom spray jet has a beneficial or detrimental impact on the liquid distribution.

4.2.2 Effect of Bottom Spray Nozzle Penetration on the Agglomerate Breakup Time Constant

As previously discussed, the second configuration that was investigated incorporated a bottom spray nozzle with an adjustable penetration, and a top spray nozzle that was fixed in position. At a horizontal separation between spray jet tips of 0 cm, the bottom spray jet shielded the top spray jet of bubbles and thus the performance was worse, which illustrates a dominant Mechanism 1. As the horizontal separation between spray jets tips increased to more than 30 cm, the performance was approximately the same as that of the base case. The increased separation between the spray jets eliminated the shielding effect as bubbles could migrate behind the bottom spray jet without being captured (Figure 4.14). Thus the performance was either worse than the base case or approximately the same, so the gum arabic method was not applied to this configuration as no improved performance was observed.
Figure 4.13: Normalized agglomerate breakup time constant for two parallel 2.2 mm TEB spray nozzles, with a moveable lower nozzle (configuration from Figure 4.4)

Figure 4.14: Diagram of the spray jet positions in the fluidized bed (Δx: 30 cm, Δz: 20 cm)
4.2.3 Effect of Top Spray Nozzle Penetration and Vertical Distance Between Spray Jets on the Agglomerate Breakup Time Constant

The third and final configuration that was investigated involved a top spray nozzle with an adjustable nozzle penetration, and a vertical distance between spray nozzles of 37 cm instead of 20 cm as in the previous two configurations. At a horizontal separation between jet tips of 0 cm, the performance was similar to that of the base case. The first configuration, where the vertical distance between spray nozzles was 20 cm, resulted in a dramatic decrease in performance due to the bottom spray jet starving the top spray jet of gas bubbles. This was not the case with the configuration with an increased vertical distance, as observed by the conductance measurements, since the gas bubbles that flow between the bottom spray jet and the walls of the fluidized bed can move back to the central region between the walls and be captured by the top spray jet.

At a vertical distance between spray nozzles of 37 cm and a horizontal separation of 20 cm, an improved performance resulted, due to the bottom spray jet releasing bubbles that could be captured by the top spray jet, as observed in Figure 4.16. This is similar to what was observed in the first configuration and results in the top spray jet expanding at a faster rate and improving the liquid distribution in the fluidized bed due to a dominant Mechanism 2.

As the top spray nozzle penetration is increased, the performance is similar to that of the base case, for both the 20 and 37 cm cases, since the spray jets are no longer interacting.
Figure 4.15: Effect of increasing the vertical distance between two parallel 2.2 mm TEB spray nozzles on the agglomerate breakup time constant, with a moveable top nozzle (configurations from Figure 4.3 and Figure 4.5)

Figure 4.16: Diagram of the spray jet positions in the fluidized bed

($\Delta x$: 20 cm, $\Delta z$: 37 cm)
4.2.4  Effect of Top Spray Nozzle Penetration and Vertical Distance Between Spray Nozzles on the Agglomerates Produced in a Fluidized Bed

After identifying ranges for which different mechanisms apply using conductance measurements, the gum arabic method described in Section 2.1.2 was applied to fully characterize the produced agglomerates. Figure 4.17 shows that significantly fewer agglomerates were produced when the vertical distance between spray nozzles was 37 cm when compared to both the base case and when the vertical separation between spray nozzles was 20 cm. When the initial L/S ratio was considered in Figure 4.18, the performance of the 37 cm case was similar to that of the base case, except for drier larger agglomerates, whereas the case with a 20 cm vertical separation had a deteriorated performance when compared to the base case.

The higher liquid content of the agglomerates produced by a 20 cm separation was due to the bottom spray jet starving the top spray jet of bubbles, as previously discussed; however, the smaller agglomerates could be explained by bubbles released by the bottom spray jet creating a more turbulent region near the tip of the top jet, which broke up some agglomerates. For the case where the spray nozzle separation was 37 cm, the bubble released by the bottom spray jet was bigger when it reached the region near the tip of the top spray jet due to more available time for coalescence to occur. Thus, even more turbulence was caused near the tip of the top spray jet and fewer agglomerates were produced. Since shielding of gas bubbles was not found to occur for a vertical separation of 37 cm, a slight improvement in the initial liquid content of the larger agglomerates was observed.
Figure 4.17: Effect of vertical distance between spray jets on the mass of agglomerates produced for two parallel 2.2 mm TEB spray nozzles at the same horizontal position (configurations from Figure 4.2, Figure 4.3, and Figure 4.5)
Figure 4.18: Effect of vertical distance between jets on the initial liquid to solid ratio with two parallel 2.2 mm TEB spray nozzles at the same horizontal position (configurations from Figure 4.2, Figure 4.3, and Figure 4.5)

To estimate the effect of penetration and vertical separation of spray nozzles in an industrial Fluid Coker, the simplified method developed by Sanchez (2013) described in Section 2.1.3 was used to combine the effects of the mass of agglomerates produced and their initial liquid content. For a horizontal separation of 0 cm, Figure 4.19 shows that, at a vertical separation of 20 cm, the performance was approximately the same as that of the base case. Although the agglomerates have a higher initial liquid content, the effect of fewer agglomerates being produced counteracts this and the performance is similar to that of the base case.

When a greater vertical separation of 37 cm is used, still with a horizontal separation of 0 cm, the agglomerates in an industrial Fluid Coker will always be significantly drier, due to the effect of fewer agglomerates being produced and the fact that the larger
agglomerates have a lower initial liquid content as observed when looking at the initial L/S ratio.

![Graph showing the effect of vertical distance between jets on the fraction of liquid remaining in coke agglomerates in the Fluid Coking Process for jets at the same horizontal distance.](image)

**Figure 4.19**: Effect of vertical distance between jets on the fraction of liquid remaining in coke agglomerates in the Fluid Coking Process for jets at the same horizontal distance (configurations from Figure 4.2, Figure 4.3, and Figure 4.5)

For vertical separations of 20 and 37 cm, a significant improvement was observed for a horizontal distance between spray jet tips of 20 cm. From the agglomerate size distribution in Figure 4.20, both vertical separations produced fewer agglomerates than the base case due to bubbles released by the bottom spray jet and being captured by the top spray jet, thereby increasing the expansion rate of the top spray jet. Fewer agglomerates were also produced when the vertical separation was 37 cm, and the horizontal distance between spray jets was 20 cm due to the bubbles that were released from the bottom jet growing as they rise (Darton et al. 1977). Thus, a larger bubble captured by the top jet results in the top spray jet expanding more rapidly, which leads to an improved liquid distribution, as shown in Section 3.2.1. The initial liquid contents of
both vertical separations were similar and significantly less than that of the base case due to the top spray jet capturing bubbles released by the bottom spray jet. These trends confirm the results obtained using conductance measurements.

Figure 4.20: Effect of vertical distance between jets on the mass of agglomerates produced for two parallel 2.2 mm TEB spray nozzles positioned at the best cases (configurations from Figure 4.2, Figure 4.3, and Figure 4.5)
Figure 4.21: Effect of vertical distance between jets on the initial liquid to solid ratio with two parallel 2.2 mm TEB spray nozzles positioned at the best cases (configurations from Figure 4.2, Figure 4.3, and Figure 4.5)

Using the simplified model developed by Sanchez (2013) to estimate the fraction of injected liquid remaining in agglomerates resulted in both vertically separated configurations having approximately the same liquid content irrespective of time in an industrial Fluid Coker. Another significant trend is that at any given time, the liquid content of the agglomerates in a Fluid Coker would be approximately half of that of the base case due to the combined effects of the mass of agglomerates produced and their liquid initial liquid contents. This is very promising as significantly less fouling would result in the scrubber, as discussed in Section 2.1.3, which results in improved heat distribution and reduces the need for shutdowns to clean the Fluid Coker. Additionally, less time is required to crack the bitumen feed, so less feed is wasted and the throughput to the Fluid Coker could theoretically be increased to process more bitumen. This also depends on other hydrodynamic factors which include bogging of the Fluid Coker due to a higher liquid content.
Figure 4.22: Effect of vertical distance between jets on the fraction of liquid remaining in coke agglomerates in the Fluid Coking Process for jets positioned at the best cases (configurations from Figure 4.2, Figure 4.3, and Figure 4.5)

From the conductance measurements, it was found that the performance of the cases where the distance between spray jet tips was 45 cm was similar to that of the base case, irrespective of vertical separation. Through the agglomerate size distribution presented in Figure 4.23, the mass of agglomerates produced by both vertically separated cases was approximately equal with little variation, which confirms the results obtained from the conductance measurements. The major difference in the trend of the agglomerate size distribution when compared to the results obtained using the conductance measurements is that fewer agglomerates are produced by both of the vertically separated cases compared to the base case. Previously, the explanation for this phenomenon was that bubbles released by the bottom spray jet created turbulence near the tip of the top spray jet; however, the spray jets do not interact at horizontal distances between spray jet tips of 45 cm. Since solids in a fluidized bed circulate, the bubbles that are released by the
bottom spray jet could still break agglomerates that recirculate as they rise in the fluidized bed.

The difference in the agglomerate size distribution between the base case and the vertically separated cases is not as great at a horizontal distance between spray jet tips of 45 cm compared to when the bubbles reach the tip of the top spray jet. Therefore, the bubbles released by the bottom spray jet still contribute to the destruction of agglomerates even though they do not reach the tip of the top spray jet.

The initial L/S ratio in Figure 4.24 provides a similar conclusion as the conductance measurements for a distance between spray jet tips of 45 cm. The initial liquid content was found to be approximately the same as that of the base case, which was due to the fact that the spray jets were out of range of each other and no longer interacting.

![Figure 4.23: Effect of vertical distance between jets on the mass of agglomerates produced for two parallel 2.2 mm TEB spray nozzles at maximum penetration (configurations from Figure 4.2, Figure 4.3, and Figure 4.5)](image_url)
Figure 4.24: Effect of vertical distance between jets on the initial liquid to solid ratio with two parallel 2.2 mm TEB spray nozzles at maximum penetration (configurations from Figure 4.2, Figure 4.3, and Figure 4.5)

The simplified model developed by Sanchez (2013) was employed to estimate the effect of the horizontal and vertical distances between jet tips on the fraction of liquid injected remaining in agglomerates in a Fluid Coker. From Figure 4.25, the use of non-interacting, vertically separated spray jets resulted in a significant improvement when compared to the base case as agglomerates would always be drier. Although the initial L/S ratio showed that all of the cases were approximately the same, the effect of fewer agglomerates produced by the vertically separated spray jets offsets the initial liquid content, as the effect of mass and heat transfer are reduced.
Figure 4.25: Effect of vertical distance between jets on the fraction of liquid remaining in coke agglomerates in the Fluid Coking Process for jets at maximum penetration (configurations from Figure 4.2, Figure 4.3, and Figure 4.5)

To confirm the theory that the bottom spray jet releases bubbles that could be captured by the top spray jet, if at the correct position, the fraction of both blue and yellow dyes in the agglomerates was determined. Since one solution containing blue dye and one containing yellow dye were injected into the fluidized bed through separate spray nozzles, mixing of the dyes would mean agglomerates formed by one spray jet migrated to the other. To test this, the normalized fraction of yellow in agglomerates in the 4000-9500 μm size cut was determined for 20 random samples of single agglomerates in this size cut. This was done by determining the initial mass of dye in the agglomerate by using the gum arabic technique developed by Pardo (2015) to determine the initial L/S ratio.

The normalized fraction of yellow dye in the agglomerate, $M^*$, was determined by using the following fraction
\[ M^* = \frac{C_{\text{yellow}}^* - C_{\text{blue}}^*}{C_{\text{yellow}}^* + C_{\text{blue}}^*} \]  

where \( C_{\text{yellow}}^* \) is the mass of yellow dye in the agglomerate divided by the mass of yellow dye in the original yellow dye container, and \( C_{\text{blue}}^* \) is the mass of blue dye in the agglomerate divided by the mass of blue dye in the original blue dye container. With this fraction, a value of 1 represents an agglomerate that only contains yellow dye, whereas a value of -1 corresponds to an agglomerate with only blue dye.

**Figure 4.26: Effect of horizontal distance between spray jet tips on the degree of mixing between spray jets**

Therefore, some mixing occurs between the spray jets for a horizontal distance between jet tips of 20 cm as \( M^* \) is not always one. This is due to the bubbles carrying solids in their wakes as they rise (Miyahara et al. 1988). Thus, agglomerates formed by the bottom jet are carried by the bubble released by the bottom spray jet and captured by the top spray jet.
For the case where the separation between jet tips is 45 cm, $M^*$ is always approximately either +1 or -1, meaning the agglomerates are either completely yellow or blue. This confirms the fact that the spray jets do not interact at a horizontal separation between spray jet tips of 45 cm and that at a horizontal separation of 20 cm, the top spray jet captures bubbles that are released by the bottom spray jet and thus an improved performance is observed. Yellow dye was used in the bottom jet and blue in the top jet. Therefore, there should be more mixing of yellow agglomerates, since they are carried by bubbles to the top jet, which is what was observed.

To determine the effectiveness of the conductance method at preliminarily screening the results for improved cases, the initial fraction of injected liquid trapped in agglomerates was correlated with the normalized agglomerate breakup constant (Figure 4.27). A strong correlation still exists between the parameters; however the effect of the size distribution of agglomerates produced is not considered, which was found to have greatly impacted the results. This was also confirmed by studies conducted by Weber (2009) and Parveen (2011), which showed that the stability of wet agglomerates in a fluidized bed is greatly affected by their liquid content.
Figure 4.27: Correlation between the initial fraction of injected liquid trapped in agglomerates and the agglomerate breakup time constant for vertically separated jets

\[ y = 0.3414 \ln(x) + 0.5332 \]

\[ R^2 = 0.892 \]
5 Effect of Interactions of Inclined Spray Nozzles on the Liquid Distribution in a Fluidized Bed

The Fluid Coking Process supplies approximately 20% of Canada’s finished petroleum products (ExxonMobil n.d.). Therefore, it would be reasonable to improve the operation of a Fluid Coker. One way to accomplish this would be to improve the liquid distribution in a Fluid Coker, as this reduces heat and mass transfer limitations by forming smaller agglomerates (House, P 2007). In industrial Fluid Cokers, spray nozzles are distributed vertically in nozzle banks. The spray jets do not interact at an angle in existing Fluid Cokers. If interacting angled spray jets could improve the liquid distribution in a Fluid Coker, the performance would be improved and thus the yield of bitumen feed to valuable products would be increased. Simple modifications to the spray nozzle assembly could be made to incorporate spray jet interactions without jeopardizing the integrity of the body of the Fluid Coker (Figure 5.1 and Figure 5.2). In spite of their potential benefits, interactions of inclined gas-liquid spray jets have not been studied before.
Figure 5.1: Simplified cross section of a Fluid Coker with an angled lower spray nozzle

Figure 5.2: Simplified top view of a Fluid Coker with an angled spray nozzle
Interacting spray jets could also reduce the size of agglomerates produced due to more turbulent regimes being formed when spray jets meet. Currently, the size of agglomerates in a Fluid Coker is controlled by the use of attrition jets which incorporate a high velocity gas jet to break agglomerates. Thus, if interacting spray jets could reduce the size of agglomerates produced, the excess gas that would have been used by the attrition jets could be used to increase the Gas to Liquid Ratio (GLR), which has been found to greatly improve the liquid distribution in a Fluid Coker (Morales, C 2013; House, P 2007).

5.1 Experimental Setup and Methodology

Two different spray nozzle setups that used two synchronized 2.2 mm TEB spray nozzles were tested in the fluidized bed that was described in Section 2. In the first configuration, spray nozzles were installed in ports a and b respectively with the spray nozzle in port b having a fixed position, and the spray nozzle in port a having an adjustable angle as seen in Figure 5.4. In the second case, spray nozzles were installed in ports a and d and the spray nozzles’ positions were fixed except for the inclination of the spray nozzle in port a (Figure 5.5). In this way, the effect of the inclination of the lower spray jet on the liquid distribution in the fluidized bed was determined. A base case from Section 3 was established to compare the effects of interaction to the simple horizontal non-interacting case (Figure 5.3).

Figure 5.3: Two horizontally opposite spray jets with one fixed nozzle and one moveable nozzle used as the base case to compare the effectiveness of interactions

($\Delta x_n$: 107 cm, $\Delta z$: 0 cm)
5.2 Results and Discussion

In the present work, the effect of interacting spray jets on the liquid distribution in a fluidized bed was determined. This was accomplished by studying the effect of angled interacting spray jets on the liquid content of agglomerates produced. The main objective of this study was to determine the effect of spray nozzle interactions on the mass of liquid remaining in an industrial Fluid Coker. This was achieved by the use of a simplified model developed by Sanchez (2013), which was described in Section 2.1.3, to determine the liquid content in agglomerates produced in an industrial Fluid Coker, based on initial values determined in a smaller scale fluidized bed. To obtain the initial liquid content in agglomerates produced in a fluidized bed, a binder solution of gum arabic dissolved in water was injected into a fluidized bed of silica sand particles so that actual agglomerates could be produced. This process was time consuming and labour intensive, so a preliminary screening method was applied to quickly select the most promising distance between interacting spray nozzles, by estimating the effectiveness of the liquid distribution. The conductance method that was utilized for preliminary screening was that of Section 2.1.1.1. The most promising results from this conductance method were validated using the gum arabic solution.
5.2.1 Effect of Inclination of Spray Jets Spraying in the Same Direction on the Liquid Distribution

From Figure 5.6, the normalized agglomerate breakup time constant is used to describe the liquid distribution in the fluidized bed. A lower value of the normalized agglomerate breakup time constant is desired, since $\tau$ is the time required for 62.5% of the liquid trapped in agglomerates to be released. For parallel spray nozzles, when the angle is 90°, the performance of the spray nozzle configuration was dramatically reduced to more than twice as worse as two horizontal, non-interacting spray nozzles. Decreasing the angle from the vertical did not result in a significant change in the normalized agglomerate breakup time constant, which suggests that the angle of inclination has a negligible effect on the liquid distribution in a fluidized bed, and always results in severe consequences.

The contraction of spray jets is due to the spray jet capturing bubbles from the fluidized bed and releasing them, which results in better liquid distribution in the fluidized bed. Since the spray jets were contiguous, the bottom spray jet captured bubbles from the fluidized bed; however, the top spray jet did not have access to as many bubbles from the fluidized bed since most of them were captured by the bottom spray jet. Therefore, the bottom spray jet starved the top spray jet of fluidized bed bubbles and thus, the top jet did not have an increased expansion cycle time. This is similar to what was found in Section 4.2.1, where Mechanism 1 was the starvation of bubbles, Mechanism 2 was the capture of bubbles released from the bottom jet by the top jet, and Mechanism 3 was the reduction of the mass of agglomerates produced by the released bubble from the bottom jet. In this case, Mechanism 1 dominated.
Figure 5.6: Agglomerate breakup time constant from conductance measurements for two 2.2 mm TEB spray nozzles with one horizontal and one upwardly inclined nozzle (configuration of Figure 5.4)
Figure 5.7: Effect of lower nozzle inclination on the mass of agglomerates produced for two 2.2 mm TEB spray nozzles (configurations of Figure 5.3 and Figure 5.4)

From Figure 5.7, the mass of agglomerates produced for the 70 and 90° cases were approximately equal, and were less than the mass of agglomerates produced by the base case. When the lower spray nozzle was directed at the top nozzle at an angle of 70°, the meeting spray jets destroyed agglomerates. Similar to Section 4.2.1, fewer agglomerates were produced due to the lower spray jet releasing bubbles that created a more turbulent region near the tip of the top spray jet. Arriyapadi et al. 2003 found that agglomerates are produced near the jet tip, so bubbles that disrupt this region ensure fewer agglomerates are produced.
Figure 5.8: Effect of lower nozzle inclination on the initial liquid to solid ratio with two 2.2 mm TEB spray nozzles (configurations of Figure 5.3 and Figure 5.4)

When comparing the initial Liquid to Solid (L/S) ratios in Figure 5.8, the cases with parallel spray nozzles on the same side and the lower spray nozzle having an angle of 70° from the vertical resulted in agglomerates with a significantly higher liquid content. These results confirm those that were obtained by the preliminary conductance measurements and are due to the fact that the lower spray jet starves the top spray jet of gas bubbles from the fluidized bed.

Compared to the base case, fewer agglomerates were produced by parallel spray jets and interacting spray jets with a lower inclined spray jet; however, they also had a higher liquid content due to the lower spray jet starving the top spray jet of bubbles. Therefore, a method that incorporates the size of agglomerates and their liquid content is used to determine the overall effect of these two parameters.
Figure 5.9: Effect of lower nozzle inclination on the fraction of liquid remaining in coke agglomerates in the Fluid Coking Process (configurations of Figure 5.3 and Figure 5.4)

From Figure 5.9, agglomerates produced by the configuration with an inclined lower spray nozzle would always contain more liquid than the base case and the case with parallel spray nozzles on the same side. The fraction of injected liquid remaining in agglomerates would be approximately the same for the base case and the case where parallel spray nozzles on the same side are used. Thus, the effect of the agglomerates having a higher initial liquid content is cancelled out by the fact that the agglomerates are destroyed by the spray jets during injection. Although this seems encouraging, no real benefit results from having parallel spray jets on the same side. Hence, the performance of parallel spray nozzles in nozzle banks in a Fluid Coker perform similarly to that of two horizontal, non-interacting spray jets.
5.2.2 Effect of Inclination of Near Spray Jets on the Liquid Distribution

Conductance measurements were used to preliminarily characterize which spray nozzle configurations resulted in an improved performance. A first inspection of Figure 5.10 shows a substantial difference in the normalized agglomerate breakup time constant. This is a strange outcome since the spray jets were not merging, so variations in the inclination of the spray nozzle should have had no effect on the performance of the non-interacting nozzle configurations. Thus, there must have been indirect interactions between the spray jets. One possible explanation for this occurrence is that the maximum penetration length was actually more than the predicted maximum penetration length. From Section 3.2.1 however, the normalized agglomerate time constant did not change when jets did not merge. Therefore, it was suggested that the spray nozzles penetrations into the fluidized bed were large enough that it allowed for indirect interactions between the spray jets as is discussed below.
To test this theory, the spray nozzles penetrations were reduced to a minimum value, so that the spray jets would not interact with the wall of the fluidized bed. The normalized agglomerate breakup time constant summary in Figure 5.11 illustrates that when the spray nozzles were furthest apart, the normalized agglomerate breakup time constant was approximately equal to one, which implies the performance was the same as that of two horizontally opposite, non-interacting spray jets. This was the case for the 70 and 90° configurations which indicated that the performance was irrespective of the spray jet inclination when the spray jets were truly non-interacting. Therefore, the bottom spray jet captures bubbles that could have been captured by the top spray jet since bubbles coalesce and rise at an angle in a fluidized bed (Darton et al. 1977). Another observation from Figure 5.11 is that agglomerate breakup is almost five times faster when the lower spray jet is angled at 70° compared to when the spray jets are parallel, due to the inclined
lower spray jet releasing a bubble that could be captured by the top spray jet, which increases the frequency of expansion of the top spray jet.

Figure 5.11: Effect of increasing the distance between near jets on the agglomerate breakup time constant for two opposite 2.2 mm TEB spray nozzles with one horizontal and one inclined nozzle (Δx: 30 cm for case where nozzles are furthest apart) (configuration of Figure 5.5)
Figure 5.12: Diagram of the spray jet positions in the fluidized bed

\((\Delta x_n: 80 \text{ cm}, \Delta z: 20 \text{ cm}, \alpha: 90^\circ)\)

From Figure 5.12, it can be seen that the lower spray jet starves the top spray jet of bubbles that would have been captured if the spray jets were not interacting. The bubbles would rise at an angle, due to the shape of the predicted bubble profile in Figure 5.16, determined by Darton et al. 1977.

Figure 5.13: Diagram of the spray jet positions in the fluidized bed

\((\Delta x_n: 108 \text{ cm}, \Delta z: 20 \text{ cm}, \alpha: 90^\circ)\)

When the spray jets were parallel and non-interacting, the lower spray jet did not starve the top spray jet of bubbles, thus the performance was similar to that of the horizontally opposite, non-interacting case.
Figure 5.14: Diagram of the spray jet positions in the fluidized bed  
\((\Delta x_n: 80 \text{ cm}, \Delta z: 20 \text{ cm}, \alpha: 80^\circ)\)

For the case when the lower spray jet was at an angle of 80° from the vertical, the performance was found to be improved. In this configuration, the spray jets were in close proximity to one another and thus the bubbles produced by each spray jet created a turbulent region between the spray jets. The bubbles that were released by the spray jets would also interact with the opposite spray jet and result in a faster expansion time, which is beneficial as an improved liquid distribution results.

Figure 5.15: Diagram of the spray jet positions in the fluidized bed  
\((\Delta x_n: 80 \text{ cm}, \Delta z: 20 \text{ cm}, \alpha: 70^\circ)\)

Similarly to the 80° case, the 70° configuration resulted in spray jets which were in close proximity to each other. Thus, the bubbles that were released created a turbulent region
between the spray jets. From Figure 5.15, the top spray jet was no longer starved of bubbles, and thus would behave as a single spray jet; however, the lower jet allowed for interactions to occur that improve the performance.

Figure 5.16: Coalescence of bubbles in a fluidized bed (Darton et al. 1977)
Figure 5.17: Effect of nozzle inclination on the mass of agglomerates produced for two opposing 2.2 mm TEB spray nozzles with near jets (configurations of Figure 5.3 and Figure 5.5)

Figure 5.17 and Figure 5.18 show that with the gum arabic analysis, significantly fewer and drier agglomerates were produced by the 70° configuration than those produced by the base case. The decreased mass of agglomerates was due to the turbulent region between the jet tips that was created, thus reducing the quantity of agglomerates produced. As well, the faster expansion time that resulted due to the released bubbles produced fewer agglomerates, as found in Section 4.2.1. The liquid content of the agglomerates was also reduced by the spray jet interactions, due to the released bubbles interacting with the jet tips.
Figure 5.18: Effect of bottom nozzle inclination on the initial liquid to solid ratio with two opposite 2.2 mm TEB nozzles (configurations of Figure 5.3 and Figure 5.5)

In an industrial Fluid Coker, the 70° spray nozzle configuration would result in a dramatically reduced liquid content in agglomerates due to the effect of the mass of agglomerates and the initial L/S ratio. Thus, less time would be required for the liquid to evaporate which leads to a better performing unit with the risk of fouling being reduced, and less feedstock being wasted.
Figure 5.19: Effect of inclination of the lower nozzle on the fraction of liquid remaining in coke agglomerates in the Fluid Coking Process with two opposite jets (configurations of Figure 5.3 and Figure 5.5)
Figure 5.20: Effect of lower nozzle penetration on the mass of agglomerates produced for two parallel 2.2 mm TEB spray nozzles (configurations of Figure 5.3 and Figure 5.5)

From the conductance measurements, it was evident that the nozzle penetration in the fluidized bed had a great impact on the performance of the spray nozzle configuration. The gum arabic analysis indicated that parallel non-interacting opposite spray jets at a different vertical position produced a similar mass of agglomerates compared to the base case, and the liquid content was also approximately the same. For interacting parallel spray jets, significantly fewer agglomerates were produced when compared to the base case. This was due to the fact that the lower spray jet releases bubbles that created more turbulence near the top jet’s tip. Since agglomerates are formed near the tips of spray jets (Ariyapadi et al. 2003), the turbulence that was caused by the released bubble from the bottom spray jet effectively reduced the mass of agglomerates produced. A second mechanism also determines the performance of this spray nozzle configuration: the starvation of the top spray jet of bed bubbles by the bottom spray jet. This was reflected in the initial L/S ratio where the liquid content was much greater than the base case.
Figure 5.21: Effect of bottom nozzle penetration on the initial liquid to solid ratio with two opposite 2.2 mm TEB nozzles (configurations of Figure 5.3 and Figure 5.5)

To determine the full effect of interacting spray jets in an industrial Fluid Coker, a simplified model was used to incorporate both the mass of agglomerates produced, and the initial liquid content. The case where interacting parallel spray jets are used would result in agglomerates with a higher initial liquid content; however, the mass of liquid remaining in agglomerates is approximately the same as the base case thereafter. This is due to the decreased mass of agglomerates produced, which counteracts the higher initial liquid content in the produced agglomerates. When parallel non-interacting spray jets are used, the mass of liquid remaining in agglomerates would be approximately equivalent to that of the base case, which is due to the agglomerates having a similar liquid content and mass as the base case.
Figure 5.22: Effect of penetration of the lower nozzle on the fraction of liquid remaining in coke agglomerates in the Fluid Coking Process with two opposite (configurations of Figure 5.3 and Figure 5.5)

To determine the accuracy of the conductance model at screening the results for improved cases, the initial fraction of injected liquid trapped in agglomerates was correlated once more with the normalized agglomerate breakup time constant (Figure 5.23). Thus, a strong correlation exists between the two parameters and the use of the conductance model was efficient at identifying the most improved cases. This was also confirmed by studies conducted by Weber (2009) and Parveen (2011), which showed that the stability of wet agglomerates in a fluidized bed is greatly affected by their liquid content.
Figure 5.23: Correlation between the initial fraction of injected liquid trapped in agglomerates and the normalized agglomerate breakup constant for inclined jets

\[ y = 0.2675 \ln(x) + 0.5962 \]

\[ R^2 = 0.9361 \]
Chapter 6

6 Conclusions and Recommendations

The following are a summary of the most important conclusions of this thesis and recommendations for future research are also presented.

6.1 Conclusions

- In a fluidized bed, the initial liquid content of agglomerates is greatly affected by the synchronization of horizontally merging spray nozzles. No negative effect would result from using unsynchronized spray nozzles in a Fluid Coker; however, to achieve an improved liquid distribution, the spray nozzles must be synchronized.

- With two synchronized nozzles of the same size, the liquid distribution is greatly improved when the spray jets slightly merge due to the expansion time being significantly reduced. The merged spray jets result in an unstable single jet, which allows for bubbles to be released faster from the spray jet. The released bubbles force the jets to contract and thus dry solids can replenish the wetted solids. Fewer agglomerates are also produced due to the improved liquid distribution. Thus, less attrition gas would be required in a Fluid Coker.

- The gum arabic method and conductance measurements were used experimentally to better understand the liquid distribution in a Fluid Coker. To fully understand the relationship between the two methods, a correlation was established between the initial fraction of injected liquid in agglomerates and the agglomerate breakup constant, $\tau$. A strong correlation exists between the initial fraction of injected liquid in agglomerates and $\tau$.

- Optimum nozzle positions exist for vertically separated spray nozzle configurations depending on the horizontal and vertical distances between jet tips.
The bottom spray jet can affect the liquid distribution performance of the top spray jet though three mechanisms:

1. The bottom jet captures gas bubbles that would normally have been captured by the top jet, increasing the expansion cycle time of the top jet, which has a negative impact on its performance.

2. The bottom jet periodically releases a bubble from its tip that can be captured by the top jet, if the top spray nozzle penetrates farther than the bottom spray nozzle. This bubble thus greatly reduces the expansion cycle time of the top jet, which has a beneficial impact on its performance.

3. The bubble released from the tip of the bottom jet can provide additional turbulence that helps break up agglomerates released from the top jet.

The combination of these three impacts determines whether the bottom jet has a beneficial or detrimental impact on the liquid distribution of the top spray jet.

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Although fewer agglomerates are produced by spray jets at the same horizontal position but different vertical positions, they have a higher liquid content than those produced by two horizontally opposite, non-interacting spray jets. In this case, Mechanism 1 increases the liquid content of the agglomerates produced by the top jet, while Mechanism 3 helps breakup these agglomerates (Mechanism 2 does not apply). When both effects are integrated to estimate the impact on a Fluid Coker, the overall performance is approximately the same as that of the horizontally opposite, non-interacting spray jets. No improved performance results from having a bottom spray jet penetrating further than the top spray jet, because Mechanism 3 is effective only when the bubble released from the bottom jet reaches the agglomerates produced by the top jet before they can dry.

Increasing the vertical distance between jets reduces the detrimental impact of the bottom jet on the top jet, by minimizing Mechanism 1 without reducing its beneficial impact through Mechanisms 2 and 3.
Interaction between inclined nozzles in the same vertical plane and spraying in the same direction always degrades the liquid distribution as a result of Mechanism 1: the bottom spray jet starves the top spray jet of bubbles. With the two nozzles in the same vertical plane and spraying in the same direction, there can be degradation even when the two jet cavities do not meet as the lower nozzle jet starves the top jet of gas bubbles. With the two nozzles entering from opposed directions, there can be degradation when the bottom jet penetrates far enough to starve the top jet of gas bubbles, taking into account that bubbles move up at an angle with the vertical. With two inclined nozzles from opposed directions that have spray jets that are in close proximity of one another, the performance can be improved when the bottom jet is close enough to the top jet so that the released bubble, causes a significant increase in the turbulence surrounding the top spray jet.

### 6.2 Recommendations

Various experiments were performed with inclined and straight spray nozzles. One of the cases that were investigated involved two spray nozzles spraying in the same direction with a straight top nozzle and an inclined lower nozzle. Due to limitations in the fluidized bed’s width that was used, the spray nozzles could not be placed alongside each other and study the effect of inclination of one of the spray nozzles. This case would be more feasible in a Fluid Coker as spray nozzles are distributed along the circumference of the Fluid Coker. In this way, a significant benefit could arise due to the fact that the spray jets could merge, which was found to be beneficial in Section 3.2, and there would be no starvation of bubbles.

The effect of the penetration and vertical separation of spray jets was also studied. To minimize the effect of the bottom spray jet starving the top spray jet of bubbles, the bottom spray jet could be angled horizontally, while keeping the top spray jet straight. In this way, the top spray jet is not starved of bubbles and benefits from the release of bubbles from the bottom spray jet.
The simple conductance method that was used was meant for preliminary characterization of the liquid distribution in a fluidized bed. The accuracy of the model could be increased by either performing more replicates for a given data point or incorporating more variables in the adjusted parameters (see Appendix A). The other models mentioned in Appendix A could provide more accurate results, but would also result in more degrees of freedom which could overspecify the equation.

Although the developed model from Sanchez (2013) was utilized to understand the effect of interactions on the liquid content in a Fluid Coker, the effect of scale-up was not considered. Therefore, similar experiments with gum arabic should be performed in a pilot scale fluidized bed to determine the effect of industrial scale spray nozzles on the agglomerates produced, so that future experiments in smaller scale units could easily be scaled to a Fluid Coker. In this way, unified explanations could be proposed for the different interactions discussed.

Predicting exactly how spray jets interact in a fluidized bed is difficult and thus other procedures were used to try to explain the interactions. To fully explain the interactions, a binder solution with a dissolved radioactive tracer could be injected into the fluidized bed and X-ray imaging analysis could be performed to observe the actual interactions in a fluidized bed.
References


Hamidi, M. (2015). *Development And Study Of Measurement Methods For Jets And*
Bogging In A Fluidized Bed. London, Ont.: Faculty of Graduate Studies, University of Western Ontario.


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Appendices

Appendix A: Discussion of Different Conductance Models that could be used

Various mixing models could be applied to the conductance measurements to obtain the required parameters. Depending on the complexity of the mixing model, some could be simpler to incorporate, especially since the conductance measurements served as a preliminary screening method. Two simple linear mixing models were investigated in this study. The first model assumed the electrode region was segregated from the rest of the well-mixed bed. Therefore,

\[
G_{\text{local}} = G \frac{t}{\Delta t_m} \tag{A.1}
\]

for \( t < \Delta t_m \).

\[
G_{\text{local}} = G \tag{A.2}
\]

for \( t \geq \Delta t_m \).

The second model assumed the region where the liquid is sprayed is segregated from the rest of the bed, including the electrode region, which is well-mixed. The wet region has a small free moisture initially, and since little fluidization gas would come in contact with the negligible free moisture, no evaporation is assumed to occur for \( t < \Delta t_m \).

\[
G_{\text{local}} = \beta \frac{t}{\Delta t_m} \left[ 1 - \left( 1 - \gamma_0 \right) e^{-\frac{t}{\tau}} \right] \tag{A.3}
\]

For \( t \geq \Delta t_m \),

\[
G_{\text{local}} = \beta \left[ 1 - \left( 1 - \gamma_0 \right) e^{-\frac{t - \Delta t_m}{\tau_{ek}}} \right] \tag{A.4}
\]
The parameters that could be adjusted are summarized in Table A.1. In the experimental analysis conducted in this thesis, the linear mixing model summarized by equations A.1 and A.2 were incorporated, along with assuming $\gamma_0$ was 0, and adjusting $\tau$, $\Delta t_m$, $\beta$, and $t_{ek}$. Different parameter adjustments could be incorporated; however, increasing the number of parameters could overspecify the problem as four replicates were required. This was found to be the case when almost all of the problems including the start time of injection were adjusted.

**Table A.1: Summary of variables that could be adjusted in the conductance models**

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# Curriculum Vitae

**Name:** Helal Elkolaly

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**Related Work Experience:**

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