Numerical study of the effect of gas distributors and baffles on the bubble distribution, gas and solid mixing in a fluidized bed

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Abstract

In this study, the multi-phase Eulerian-Eulerian two-fluid method (TFM) coupled with the kinetic theory of granular flow (KTFG) was used to investigate hydrodynamics of particle flows (Geldart Group B) in a lab-scale fluidized bed Geldart Group B particles, operating in bubbling and turbulent regimes. The effect of gas distributors and baffles on the distribution of the gas bubbles, and the mixing of gas and solids were investigated under various superficial gas velocities.

The numerical model is validated by experimental results from two different measurement methods with various gas distributor configurations and a superficial gas velocity ranging from 0.4 m/s to 1.0 m/s; the E-probe method measured the local gas flux while the radiation transmission method provided the local solid hold-up. Simulation with different gases and particles spanned the range from lab to industrial conditions. The gas distributor configuration and angle were found to have a significant impact on the gas bubble distribution.

Baffles are used to modify fluidized bed hydrodynamics in industrial processes. This work simulated the impact of various baffles on fluidized bed hydrodynamics. A ring baffle can redirect gas bubbles and induce strong liquid recirculation currents. Adding a vertical fluxtube to a baffle can significantly modify its impact on the gas flow patterns. With a fluxtube that does not extend past the baffle lip, the gas is more evenly distributed in the fluidized bed. The fluxtube length has a stronger impact than the fluxtube diameter on the fluidized bed hydrodynamics.

New methods were developed to characterize the gas and solids mixing patterns from the simulation results. Gas and solids mixing in both horizontal and vertical directions are affected by the gas distributor configuration and the presence of a ring baffle. The ring baffle separates the bed into two regions and reduces the back mixing of gas and solids between the upper and lower regions.
Keywords

Fluidized bed, numerical study, CFD modelling, gas-solid fluidized bed, bubble distribution, solid mixing, particle disperse, particle tracking.
Summary for Lay Audience

In fluidized beds, gas is injected into a bed of particles to impart a liquid-like behavior to the gas-solids mixture. Fluidized beds are used in many industrial processes ranging from oil refining and biofuels production to pharmaceuticals and food applications. This thesis uses a CFD numerical method to simulate fluidized bed systems with a set of governing equations. Due to the fast development of computer technology, CFD modelling has become an effective and economical tool to investigate fluidized beds. Different numerical methods have been developed in this work. New technology is introduced to track the particle and gas molecule in the fluidized bed using the two-fluid model. The statistical time distributions, dispersion rate, and the mixing rate in the lateral and vertical direction for solid and gas were investigated.

For this thesis, the strategy was to use experimental data obtained in a small laboratory fluidized bed to validate CFD modelling tools. These tools were applied to verify that the lab experimental data were obtained under conditions that would be relevant to industrial processes. They were then used to show that the performance of the lab-scale fluidized bed could be greatly improved by using baffles and/or modifying the initial gas distribution into the bed. In the future, these tools will be applied to the optimization of industrial fluidized beds in which it would be very difficult, unsafe and extremely costly to run experiments. Optimizing industrial beds reduces their cost and minimizes their environmental impact.
Co-Authorship Statement

Chapter-1: Xuelian Xing prepared the original manuscript, including collecting references, writing the manuscript. Professor Cedric Briens reviewed and revised the manuscript. Professor Chao Zhang gave directions at the time of writing the manuscript.

Chapter-2: Xuelian Xing prepared the original manuscript, including collecting references, writing the manuscript. Professor Chao Zhang reviewed and revised the manuscript. Professor Cedric Briens gave directions at the time of writing the manuscript.

Chapter-3: Xuelian Xing prepared the original manuscript, including conducting the simulation work, processing the data, and writing the manuscript. Colleague Yuan Li provided the experimental data. Professor Chao Zhang reviewed and revised this manuscript. Professor Cedric Briens aided the simulation work and design data analysis. Professor Chao Zhang helped with the simulation process, reviewed and revised the manuscript.

Chapter-4: The original manuscript was prepared by Xuelian Xing, including the design of the simulation work, analysis of the result data, and writing the manuscript. Professor Chao Zhang reviewed and revised this manuscript. Professor Cedric Briens and Professor Chao Zhang advise on the simulation model and experiment design.

Chapter-5: Xuelian Xing prepared the original manuscript, including the code for the calculation, analysis of the data, and writing the manuscript. Professor Cedric Briens and Professor Chao Zhang helped with the method chosen and the design of the experiments and the data organize. Professor Cedric Briens reviewed and revised the manuscript.

Chapter-6: The manuscript was prepared by Xuelian Xing, including the writing of the calculating code, analysis of the data and writing the original manuscript. Professor Cedric Briens helped with the chapter reviewed and revised by Cedric Briens.

Chapter-7: The manuscript was prepared by Xuelian Xing, which includes the writing of the calculating code, analysis of the data and writing the original manuscript. Professor Cedric Briens helped with the chapter reviewed and revised by Cedric Briens.
Chapter-8: The manuscript was prepared by Xuelian Xing and reviewed and revised by Cedric Briens and Chao Zhang
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<td>volume fraction of gas phase</td>
</tr>
<tr>
<td>$\alpha_s$</td>
<td>volume fraction of solid phase</td>
</tr>
<tr>
<td>$\rho_g$</td>
<td>gas density, kg m$^{-3}$</td>
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<td>$\rho_s$</td>
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<td>$\vec{\dot{\vartheta}}_g$</td>
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<td>Gas-solid momentum exchange coefficient</td>
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<td>minimum fluidization velocity</td>
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<td>shear viscosity</td>
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<tr>
<td>$\Theta_s$</td>
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</table>
\( t_{10} \) \hspace{3cm} \text{Smallest 10 \% of cumulative time distribution, s}

\( \tau_{10} \) \hspace{3cm} \text{Dimensionless time distribution}
Chapter 1

1 Introduction

Fluidization is widely applied in the industry due to its excellent gas and solids mixing and rapid mass and heat transfer. This dissertation focuses on the application of numerical methods to study the impact of gas distributors and baffles on bubbles flow patterns, solids mixing and gas mixing in 2D fluidized beds.

This chapter introduces the research background, literature review, and the structure of the thesis.

1.1 Background

1.1.1 Fluidization system

Fluidization is an operation in which a bed of solid particles is transformed into a liquid-like state by forcing a fluid through the bed. Fluidized beds have unique properties: rapid mixing of solids leads to a uniform temperature distribution and improve the gas-solid contract efficiency for the reaction, and the solids can be easily circulated between reactors [1,2]. Depending on the fluid media, fluidization can be divided into gas-solid fluidization, liquid-solid fluidization, and gas-liquid-solid fluidization. This thesis focuses on gas-solid fluidized beds.

In a gas-solid fluidized bed, gas is distributed into the reactor at the bottom and flows up through the bed. Figure 1-1 shows the different regimes obtained as the gas velocity is increased, going from fixed bed to bubbling fluidization, turbulent fluidization and fast fluidization [2,3]. While bubbling fluidization was the most popular regime for chemical and physical processes, many industrial applications have migrated to turbulent fluidization, which maximizes throughput through a given column size [4]. Many inherent fluidization characteristics are related to bubble properties, such as bubble size, bubble distribution, bubble velocity, bubble coalescence and splitting [1,2,5,6]. Bubble properties can be studied experimentally, using scaled-down laboratory equipment, or through modelling.
For experimental studies, bubble measurement techniques can be classified as non-intrusive (NMT) and intrusive (IMT) techniques. Intrusive probes can be used to measure the radial and axial solids concentration, solid velocity and distribution, but their presence may affect bubble properties. Non-intrusive methods such as digital image analysis or X-ray computed tomography is more desirable as they do not affect the gas bubbles [7–9].

Modelling the fluidized bed has also become a useful tool to investigate fluidized bed hydrodynamics. A series of mathematical models have been proposed. There are two numerical approaches: one is the Eulerian-Eulerian model in which the gas and solid phases are treated as interpenetrating continua, and each phase has its governing equations for momentum, continuity, and energy. The other is the Eulerian-Lagrangian method; in this method, the trajectory of each particle is tracked by Newton’s second law. Chapter 2 includes a detailed review of the numerical methods.

Figure 1-1 Flow regimes of gas–solid fluidization [9]
1.2 Fluidized bed applications

Gas-solid fluidization has been widely applied to different industrial processes. This introduction reviews the fluidized bed coal gasification and combustion, Fluid coking and Flexi-coking, and drying processes.

1.2.1 Combustion and gasification

Fluidized bed combustors and gasifiers are widely used in many chemical and power industries to supply process heat, steam generation, and electrical power. Compared to a traditional combustor, a fluidized bed combustor has many advantages, such as high combustion efficiency and low pollutant emissions [10,11]. Fluidized bed combustion (FBC) is a useful technology that is mainly used for coals and biomass fuels [12]. Both bubbling fluidized beds and circulating beds have achieved commercial status for combustion. In a fluidized bed combustor, carbonaceous fuel is feed into a fluidized bed with Sulphur-sorbent particles or other particles that may include a small portion of char. The bed is fluidized by air, and the combustion happens in a relatively well-mixed and uniform temperature environment.

Three different types of fluidized beds can be applied gasification process: bubbling fluidized beds, circulating fluidized beds and twin fluidized beds, as shown in Figure 1-1. The basic principles of fluidized bed combustors and gasifier are similar. They both operate at relatively low temperatures and achieve a uniform temperature throughout the bed. They also need to overcome the same design and operation difficulties, such as gas bypassing, nonuniform fuel feeding and mixing, agglomerate formation and hot spots. The only difference is that the gasification is an endothermal conversion technology where solid fuel is converted into a combustible gas [11]. Therefore, mixing property plays an essential role in the gasification and combustion processes.
Figure 1-2 Schematic diagram of different type of fluidized bed gasifier (a) bubbling fluidized bed gasifier (b) circulating fluidized bed gasifier (c) Twin fluidized bed gasifier [13]
1.2.2 Fluid Coking™

Fluid Coking™ is a refining process that is mainly used for thermal cracking of heavy oils [14–16]. In Canada, it is applied to the conversion of bitumen from the oil sands into synthetic crude oil [17,18]. The process uses two vessels: a fluidized bed reactor and a fluidized bed burner, as shown in Figure 1-3. The fluidized bed reactor can be divided into a stripper, reaction, and scrubber sections. The heavy feedstock is pre-heated to 350 °C and injected through steam atomization spray nozzles into the fluidized bed of coke particles, and thermal cracking occurs on the surface of the hot coke particles at 500 to 550 °C. The products of the oil cracking are hydrocarbon vapor, permanent gas and coke solids. The gas and vapor products flow up through the reactor while the coke particles go down to the stripper section.

In the stripper, trapped vapors are displaced by stripping steam from the down flowing coke particles. The stripped cold coke particles are then conveyed to the burner, where a fraction burns to provide heat for the process. Reheated coke is conveyed back to the reactor to provide the heat for the thermal cracking reactions [15,18].

The local bed hydrodynamics have a significant impact on the initial liquid distribution of the sprayed liquid on the fluidized particles [17]. Li et al. [17] found out that by increasing the superficial gas velocity, the number of wet agglomerates decreased and that, in particular, increasing the gas velocity at the end of the jet improves the liquid distribution. Li et al. [19] also found out that increasing the gas velocity reduces agglomerates, and that adding asymmetrical baffle also reduces agglomeration decreased.
1.2.3 Drying process

The fluidized bed is also widely applied for the drying process in many applications like the chemical, food, ceramic, pharmaceutical, agriculture, polymer, and waste management industries [20]. Compared to traditional dryers, fluidized bed dryer has many advantages, the high drying rate due to the excellent gas-particle contact. Easy to control and low maintain costs compared to the rotary dryers. Uniform moisture reduction with less drying time and high drying rate [21]. With a fluidized bed dryer (Figure 1-4), it provides a more uniform temperature throughout the drying period and increases the gas-particle contract time. However, it also has its limitations, like the hotspot formation will lead to an uneven moisture content distribution.
A typical drying process has three different periods. In the first period is the pre-heat period with heat transfer into the moisture feedstock. Furthermore, after that is the constant rate period, in which the free moisture on the surface of the particle begins to evaporate, and this process continues until all surface moisture evaporated to the material critical moisture constant. After that, a longer time will need for the moisture from the internal interstices of the particle to be diffused. Many factors can influence the drying process in the fluidized bed dryer, the bed height, the gas velocity and the particle size [20].

Figure 1-4 Schematic diagram of the typical fluidized dryer [21]
1.3 Literature review

1.3.1 Importance of bubble distribution in the fluidized bed

In the fluidized bed, as superficial gas velocity beyond the minimum fluidization velocity, bubbles form at the distributor, rise through the bed, and grow by coalescence with other bubbles. Many characteristics of fluidized beds, such as mass and heat transfer, are dominated by bubble behavior [1,2,22]. The performance of a fluidized bed highly depends on the spatial distribution of the bubbles and their physical properties such as bubble velocity, bubble diameter and bubble frequency. Therefore, a proper understanding of the hydrodynamic behavior of the fluidized bed system, especially the bubble dynamics, is essential for the design and scale-up of fluidized beds.

Bubbling fluidized beds can be divided into two regions: the gas-rich bubble phase constituted of the gas bubbles, and the emulsion phase, which is similar to the bed at minimum fluidization conditions [2]. In the two-phase theory, the gas flow rate through the emulsion phase remains the same as at minimum fluidization, and the rest flows in the form of bubbles.

In a bubbling fluidized bed, particle mixing is entirely induced by the passage of gas bubbles [23,24]. The bubbles carry solids in their wake from which particles are exchanged with the emulsion phase as the bubble rises; to balance the upward movement of the bed materials, the solids recirculate downward in the emulsion. The solids exchange between bubbles and emulsion intensifies mass transfer. Besides, when the bubble eruption at the bed surface, the convection effects also increase the gas-solid mixing [1,2]. However, as bubbles coalesce and become more significant, they rise faster, which reduces the time available for gas-solid contract and worsens performance [25]. Apart from the vertical movement of bubbles, bubbles also tend to move towards the center as they rise. As a result, most of the bubbles concentrate far from the wall [1,26,27]. However, this will limit convection and selectivity, which may not be beneficial for the chemical reaction.

When the gas velocity continues to increase, the bubble phase becomes more prevalent, and the bed turns into a turbulent bed. Compared to bubbling fluidization, turbulent fluidization has a large renewal frequency of the bubbles and more significant freeboard activity.
Ideally, the excellent performance of the fluidized bed should be the bed with a large number of bubbles of small diameter uniformly distributed throughout the bed. In this thesis, we will focus on how the spatial distribution of bubbles can be controlled to optimize it for each process.

1.3.2 Modified methods for bubble distribution

How bubbles are distributed over the fluidized bed cross-section has a significant impact on solid mixing and gas-solid heat and mass transfer. In processes where the liquid is injected into the bed, such as fluid cokers, fluidized bed coaters, or polyolefin reactors, the modification of the radial distribution of bubble flux can also reduce the formation of unwanted wet agglomerates [16,17,28]. Modifications in gas distributor design and the addition of internals to the bed are recognized methods to change the flow patterns of gas bubbles and particles [29–35].

1.3.2.1 Effect of distributor

There are different methods applied to improve fluidization performance. Even though the gas distributor only represents a small part of the fluidized bed, its primary function is to introduce uniform and stable gas bubbles over the entire bed cross-section, prevent uneven fluidization, minimize bed materials erosion and decrease the leakage of solids into the plenum under the grid. In practice, there are a variety of different forms of distributors. Despite physical forms, the distributor can be classified by the direction of gas entry: normal, lateral or inclined [36]. The choice of a proper gas distributor depends on operating conditions, mechanical feasibility and cost.

The perforated plate (normal direction) is the most common distributor type, which is cheap and easy to modify; however, it needs a high-pressure drop to prevent solids leakage. Bubble caps and nozzles (lateral direction) can successfully prevent solids backflow and decrease the required pressure drop; however, they cost more and can be challenging to clean and modify. Spargers (lateral or downward direction) and conical grids (downward direction) are also widely used. Therefore, it is essential to predict the impact of the distributor on the performance of fluidized beds.
Generally, the distributor design greatly influences the gas bubble generation, growing and coalesce, which further has a great impact on the gas-solid hydrodynamics in terms of solid mixing and heat transfer. Most of the study focuses on a more even and uniform gas distribution in the entire fluidized bed reactor, avoid dead zone or de-fluidized zone. Some studies focus on modifying the geometry of the distributor, such as the opening area, the number of orifices and the orifice size. Sánchez-Delgado et al. [37] compared the four perforated plates with the same opening area, the different number of holes and the results showed that as the number of holes increased, the bubble generated tends to spread more evenly. Different gas jets arrangements have been tested to provide a uniform radial distribution of the gas holdup: (1) center-sparse side-dense air jets arrangement; (2) center-dense side-sparse air jets. The center-sparse side-dense air jets arrangement can improve the uniform of the radial flow. Furthermore, it can significantly flat core-annulus structure [38]. Feng et al. [39] used two different orifice diameters in the same distributor and results in a non-uniform gas bubble flow with bubbles tend to move to above of the small orifice. Some researchers investigated the effect of the slope of the distributor on the gas-solid hydrodynamics inside the fluidized bed reactor. Cai et al. [30] studied the effect of the gas distributor angle on fluidized bed hydrodynamics. An inclined distributor can intensify the transverse flow heterogeneity of the bottom zone of a fluidized bed. Each type of gas distributor has its advantages and disadvantages; a comprehensive comparison is needed for choosing a suitable distributor based on the operation process. Sobrino et al. [40] compared the flow structure near the bottom region with bubble-cap and perforated distributors. The bubble-cap distributor provided a more homogeneous radial voidage and a higher bed density in the bottom region. Rahimpour et al. [41] compared the performance of three common distributors: perforated plate, bubble cap, and porous plate with various superficial gas velocities and particle sizes. Based on the time-frequency, the initial bubble size is greater with the cap distributor. The standard deviation of pressure fluctuations is greatest with the porous plate and then perforated plate and cap distributor. Akbari et.al.[42] investigated the effect of perforated distributor characteristics on the fluidization behavior by a two-dimensionally Eulerian–Eulerian multiphase flow model coupled with a population balance modelling (CFD–PBM). It was found that bubbles tend to follow in the center of the bed, reducing the particle concentration in the central region and increasing it near the wall.
Even gas distribution is also a critical factor for fluidized bed dryers, to achieve uniform bed temperature and high heat and mass transfer rate [43]. Delgado et al. [37] used different measurement methods such as Digital Image Analysis and Wavelet Analysis to study the relationship between gas distributor design and bubble generation, growth and interactions in bubbling fluidized beds. The comparison was based on four different perforated plates with the same opening area. Moreover, they found out that the distributor does not affect the bubble generation characteristic frequencies and the long-term dynamics of the bed. However, the comparison is limited to the perforated plate, which is the reason for no effect on the bed dynamic.

The difficulty of distributing gas evenly into the fluidized bed has encouraged innovation in distributor design.

Afrooz et al. [29] studied the performance of a new swirl distributor plate to enhance the gas-solid mixing in the vertical and radial direction numerically and experimentally. Instantaneous solid holdup contours were used to validate the numerical model, and the vector diagram of particle velocity in the swirl distributor and conventional distributor were compared to show the improvement in evenly the radial distribution of solids. By using the swirl distributor, the solid particles are swirling upward from one side and swirling down from the other side of the bed. However, it may not suit equipment with a large diameter.

Brink et al. [44] investigated the influence of the novel multi-vortex distributor on the interphase mass transfer, gas axial dispersion and bubble size, and chemical reaction is included in the whole simulation. By using the new multi-vortex distributor, the conversion efficiency has been dramatically improved, and a giant bubble detected in the novel multi-vortex distributor. Like Yudin et al. [45], an early transition to turbulent fluidization is observed compared to a perforated distributor. That is all because of the new gas distributor increase the gas circulation and radial momentum.

Since fluidized bed has many applications, not all the process needs an evenly gas distribution. The design of the distributor needs to satisfy the requirement of the process. Formation of agglomerates in Fluid Coking™ can cause operation problems, such as the fouling [16,19,28,46].
The local gas-solid hydrodynamics in the fluidized bed has a significant impact on the injected liquid distribution [17,19]. Li et al. [17] investigated the effect of local hydrodynamics on the liquid distribution in a fluidized bed. They found out that increasing the superficial gas velocity near the liquid injection stage can reduce the number of agglomerates, and during the drying, intensify the shear forces that break up wet agglomerates. Li et al. [19] found out that modifying the gas lateral or radial distribution at the spray level can also significantly improve the liquid distribution. Both changing inlet gas distributor configuration and adding baffles under injection level can achieve the modification results.

In the fluidized bed reactor, bubbles play an essential role in the solid mixing. The rising bubbles cause the motion of the particles, which intense the solid mixing. The mixing quality is vital for the mass and heat transfer in the fluidized bed, such as combustion and drying processes. The gas-slit mixing pattern in a fluidized bed is significantly impacted by the gas distributor [29,47,48]. To promote solid mixing in a fluidized bed, Yudin et al. [45] applied a novel inclined slotted swirling distributor to fluidized beds with different bed aspect ratios. With 45 inclined slots, it triggered the early transition from fixed bed to fluidized bed and enhanced the solids circulation rate. Norouzi et al. [49] studied the effect of distributor type on the solid flow mixing pattern and found out that by applying the injection, the distributor can increase the time lag between the successive bubbles, reduce the tendency of bubble breakup and coalescence. Sriniketh et al. [50] found out the effect of different distributor configurations on the solids circulation rate. Three different gas perforated plates: flat, convex, and concave were compared in the study, and the convex one can increase solid recirculation rate, which is an option for the process that good solid mixing is required.

1.3.2.2 Internals and baffles

Internals are generally introduced into the fluidized bed to modify the complex gas-solid flow structures to form a more uniform gas distribution and improve the heat and mass transfer to improve the overall performance of the fluidized bed reactor. Therefore, it is vital to understand the mechanism of the effect of the baffles and select suitable baffles for different operation conditions and fluidized bed scale.

Baffles can be classified in many different categories based on physical structures and performance characteristics. In general, the internals can be divided into several different types:
baffles, tubes, packings, inserted bodies, and other novelty configurations. Baffles are one of the most commonly used internals, like the wire mesh, perforated plate, ring baffles and single or multi-turn plates.

In a bubbling fluidized bed, most of the gas goes through the bed as bubbles, which makes the bubble movement and properties play an essential role in the behavior of the bubbling fluidization.

Large bubble size is not welcome in gas-solid fluidization. First is because the large bubble usually moves fast and limits the time for gas to transfer with the emulsion phase. The second reason is the lower interfacial area between the bubbles and the emulsion phase. However, in a baffle-free fluidized bed, the bubble movement is unrestricted, the bubble size getting larger and tends to move to the center as it rises, which significantly limits the conversation. Therefore, the primary purpose of adding baffles to the bubbling fluidized bed is either to break up bubbles or to modify the bubble move direction to evenly the bubble distribution, thus enhancing the interchange between bubbles and emulsion phase to improve the mass and heat transfer, increase the reaction rate. As superficial gas velocity increases, the fluidized bed begins to transition from the bubbling regime to a turbulent regime. Different from the bubbling regime, the turbulent fluidized bed is a feature as its unstable. Like the bubble eruption rate increases, the size of the bubble becomes smaller. As the bubble renew rates is already high enough in a turbulent fluidized bed, the function of internals or baffles also different from them in a bubbling bed. The baffle's primary emission in the turbulent regime is to increase the radial movement of gas and particles [51–54].

Jahanmiri et al. [55] found that ring baffles with fluxtube change the distribution of bubbles over the cross-section and can be used to improve the liquid distribution significantly with lower agglomerates formed. Louver baffles consist of a bundle of inclined surfaces, in which the inclined surfaces can create the horizontal velocity component that can improve the efficiency of breaking bubbles and gas-solids contracting [51,54]. Zhang et al. [56] investigated the effects of louver baffles on the hydrodynamics and gas-solid mixing characteristics of a fluidized bed reactor, which operated in both bubbling and turbulent regimes. The ability of louver baffles to break bubbles for superficial gas velocities $< \sim 0.7$ m/s was proven using pressure fluctuations and steady gas tracer experiments. The breaking ability
was also validated by numerical work from Yang et al. [57]. Besides, Yang et al. [57] also found that the bubbles regenerated after baffles tend to have a similar diameter, which is equal to the distance between two adjacent vanes in the louver baffle. Jin et al. [58] used a photographic method to study the effect of louver baffle on the behavior of bubbles under different velocities. When the gas velocities are relatively low \( U_g = 0.395 \frac{m}{s} \), the effects of the baffle are apparent, it easy to observe the bubbles breakup, regeneration and uniform distribution. However, a thick “air cushion” appears below baffles which is detrimental effect on the heat transfer at higher velocities. Jin et al. [58] also investigated the effect of the vertical tube on the bubble behavior, and find out that with vertical tube inserting in the bed can even up the spatial distribution of the bubbles and suppress the tendency of concentrated in the center of the bed. The presence of horizontal tubes tends to cause the bubbles split, which can limit the bubble size and even up the bubble distribution. Yang et al. [59] found out with the presence of the tube bundle inside the fluidized bed can limit the bubble diameter and results in a homogeneous distribution of bubbles, and the tube also decreased the vertical velocity.

Baffle can also redistribute gas and solids flow in the fluidized bed reactor. Yuan et al. [19] successfully applied asymmetrical baffle to redirect the gas bubble as the spray region and further optimize the liquid distribution. Jahanmiri et al. [16] investigated the impact of ring baffle and ring baffle with fluxtube on the bubble flow patterns and liquid distribution in a fluidized bed. The objective of their work is to reduce the formation of wet agglomerates in the Fluid Coking™ process. By compared between no baffle, baffle and baffle with fluxtube, they found out that both the baffle and baffle with fluxtube can improve the liquid distribution. The improvement with fluxtube is small compared to baffle. The core-annulus concentration profile is characteristic of a circulating fluidized bed, which will lead to a solid concentrated near the wall. Baffles have also been applied to fluidized bed risers, where they have been shown to increase the gasoline yield by forcing catalyst particles from the wall region to the central region of the riser [13,60–64]. With ring baffle inserted in the fluidized bed, it even the core-annulus solid radial distribution and improves gas-solid contact [65].

As superficial gas velocity increases, the fluidized bed begins to transmit from the bubbling regime to a turbulent regime. Different from the bubbling regime, the turbulent fluidized bed is a feature as its unstable. Like the bubble eruption rate increases, the size of the bubble becomes smaller. As the bubble renew rates is already high enough in a turbulent fluidized
bed, the function of internals or baffles also different from them in a bubbling bed. The baffle's primary emission in the turbulent regime is to increase the radial movement of gas and particles [51–54]. Geng et al. [66] found a higher solid hold up and more uniform solid hold up distribution in the novel fast-turbulent fluidized bed, especially with the vortex ring-feeder, which indicates an improved interaction between gas and solid phases.

In a fluidized bed reactor, the gas-solid mixing quality is essential for many processes, such as combustion and drying process. Baffles can provide staging for solids mixing [67], reducing solids bypassing in circulating fluidized beds [46,68]. They reduce gas back mixing [56,67]. Zhang et al. [53] use the steady-state tracer technique to study the effect of louver baffle on the gas back-mixing in the fluidized bed with FCC particles. The results show that the louver baffle significantly suppressed the gas-solid back mixing and highly improve the gas-solid contract efficiency. Sanchez used radioactive particle tracking to study the effects of internal baffles in the stripping section of the Fluid Coking™ and discovered that a baffle increases the time that wet agglomerates spend above the baffle and reduces fouling on the sheds of the stripper section [15]. The efficiency of ring baffle on improving the gas flow uniformity and enhance the gas-solid contract and weaken the solid back mixing in the circulating fluidized bed has been studied by many researchers. However, not many researchers have focused on the effect of ring baffle configuration parameters. Samruamphianskun et al. [70] did a comprehensive study of the effect of ring baffle configuration on the solid distribution in a circulating fluidized bed experimentally and numerically. Baffle opening area, space between baffle, baffle thickness, and the number of baffles was all compared in their study. The standard deviation of radial solid volume fraction and the average of solid volume fraction along the height of the CFBR riser were chosen as response variables of the system mixing. They found out that the interaction between the baffle opening area and the space between the baffle effect the mixing quality the most. Wang et al. [71] also investigated the configurations of ring baffle on the circulating fluidized bed performance. With the number of ring baffles and the space between the baffle increased, the solids circulation rate and solid inventory height increased and decreased the solid cycle time.

Zhang et al. [53] use the steady-state tracer technique to study the effect of louver baffle on the gas back-mixing in the fluidized bed with FCC particles. The results show that the louver baffle significantly suppressed the gas-solid back mixing and highly improve the gas-solid contract.
efficiency. Zhang et al. [52] proposed a new multilayer baffle to intensify the FCC regeneration process. Moreover, they found out that compared to the baffle-free fluidized bed, the new baffle can significantly reduce the internal gas circulation flux by 89-96%. Moreover, the established baffled regenerator model shows a positive effect on the FCC generator performance. However, the new baffle is modified based on the louver baffle, and it would be better to compare between louver baffle and the new baffle to identify the improvement of the new baffle. Yang et al. [30] considered the baffle as a distributor, and the baffled fluidized bed was separated into several sections, which were treated as a series of free fluidized beds, and fewer whirlpools were formed in the baffled one, which indicated restrained of the solid back-mixing with the baffle.

Other kinds of internals were also investigated to improve fluidization quality. Heat exchanger tubes are generally installed into the fluidized bed to supply or remove heat to maintain the excellent heat transfer inside the fluidized bed. However, most of the studies on the heat transfer tubes focus on the heat transfer efficiency, seldom noticing the effect of the tube on the hydrodynamic of the gas-solid fluidized bed. Heat transfer tubes are arranged either vertically or horizontally in the reactor. Jin et al. [58] also investigated the effect of the vertical tube on the bubble behavior and found out that with vertical tube inserting in the bed can even up the spatial distribution of the bubbles and suppress the tendency of concentrated in the center of the bed. The presence of horizontal tubes tends to cause the bubbles split, which can limit the bubble size and even up the bubble distribution. Yang et al. [59] found out with the presence of the tube bundle inside the fluidized bed can limit the bubble diameter and results in a homogeneous distribution of bubbles, and the tube also decreased the vertical velocity.

Zhang et al. [52] proposed a new multilayer baffle to intensify the FCC regeneration process. Moreover, they found out that compared to the baffle-free fluidized bed, the new baffle can significantly reduce the internal gas circulation flux by 89-96%. Moreover, the established baffled regenerator model shows a positive effect on the FCC generator performance. However, the new baffle is modified based on the louver baffle, and it would be better to compare between louver baffle and the new baffle to identify the improvement of the new baffle.
Therefore, it is impossible to specify one kind of internals that is perfect for all the fluidized bed conditions or the requirements of industrial applications. One kind of internals is a benefit for one process that may hinder the other one. In conclusion, it is essential to investigated different baffles based on the conditions that suit best.

1.3.2.3 The numerical model of fluidized bed hydrodynamics

The industrial application of fluidized bed usually is complex in geometry, which makes it difficult to measure the gas-solid hydrodynamics and bubble behavior. Apart from experimental methods, modelling can be useful for understanding the details of bubble distribution and properties. Reliable numerical modelling can facilitate the study of the fluidization process and the effect of the gas distributor. It can also be used to scale-up the results of experimental studies to the conditions and scale of industrial processes.

However, modelling also has its limitations [72,73]. A numerical model is “validated” when its predictions agree with the results of experiments with different parameters. When CFD is applied to simulate the hydrodynamics of a fluidized bed, different independent measurement methods and more operating conditions should be considered to validate the numerical model to ensure its accuracy and reliability over a wide range of conditions.

There are two approaches for simulating the gas-solid two-phase flow, one is the Eulerian-Lagrangian (E.L.) approach [74–79], the other is Eulerian-Eulerian (E.E.) approach. The details of each method will be discussed in Chapter-2.

1.4 Motivation of this research

Different distributor configurations and baffles are applied in fluidized bed to modify the bubble dynamic, gas mixing and solid mixing in experiments which have already been discussed. However, hydrodynamic information obtained by modern measurement instruments is not enough to understand the hydrodynamic of the gas-solid flow structure affected by them. Accurate simulation is required to scale up to industrial units. Besides, there are several numerical studies of baffle with and without fluxtube on the bubble dynamic.

Besides, a new method is applied in this work to track particles and gas molecules in a two-fluid model (TFM). It is particularly useful for studying the solid and gas mixing which can
open a new way to study the particle individual properties by combing the TFM and tracking approach. If the new tracking method is successful, it can not only overcome the drawbacks of TFM model on lack of dispersed phase, but also saving time on using Eulerian- Lagrangian method to calculate.

1.5 Objectives of this research

The overall objective is to comprehensively study hydrodynamics and the underlying flow mechanisms of a gas-solid fluidized bed (bubbling regime and turbulent regime) under various conditions via a computational fluid dynamic approach.

- Investigate the gas-solid flow structure in bubbling fluidized bed and turbulent fluidized bed via a validated numerical model.
- To detect the transition phenomenon in the fluidized bed from bubbling to turbulent.
- To study the effect of gas and particle properties on the gas bubble distribution in the fluidized bed.
- To study the effect of the different gas inlet distributor slope on the gas bubble distribution in the fluidized bed.
- To study the effect of different inlet gas distributor configurations on the gas bubble distribution in the fluidized bed.
- To study the effect of baffles on the gas-solid flow structure in the fluidized bed:
  - To study the effect of fluxtube on the gas bubble distribution in the fluidized bed reactor.
  - Modify fluxtube geometry.
- To study the combined effect of the inlet gas distributor configuration and baffle on the gas-solid flow patterns in the fluidized bed.
- The effect of gas inlet conditions and baffle on the particle mixing and gas back mixing.
1.6 Thesis structure

Chapter-1 gives a simple introduction of this research work and a detailed review study of the experimental and numerical study of the gas distributor and baffle in bubbling and turbulent fluidized beds.

Chapter-2 gives the details of the numerical methods applied in this research work: equations, correlations and parameters chosen for this work.

Chapter-3 gives the fundamental study of the numerical model of the bubbling and turbulent model which includes the numerical model validation, the effect of different gas and particle properties as well as the superficial gas velocity.

Chapter-4 compares the effects of different gas distributor configurations and different gas distributor angles on bubble distribution.

Chapter-5 discusses the effects of different baffles with and without baffle on the bubble distribution on the injection level, baffle area, and the whole fluidized bed.

Chapter-6 tracks the particle in fluidized bed based on the Two Fluid Model and compares the effect of gas distributor, and baffle on the particle time distribution, particle dispersion in vertical and horizontal direction.

Chapter-7 tracks gas molecule in fluidized bed based on the Two Fluid Model and studies methods to prevent the gas back mixing.

Chapter-8 presents the conclusions of this study and recommendations for future work.

1.7 References


[55] M. Jahanmiri, Use of a baffle to enhance the distribution of a liquid sprayed into a gas-solid fluidized bed, Western univeristy, 2017.


Chapter 2

2 Numerical methods for simulations of gas-solid fluidized beds

2.1 Introduction

Numerical modelling of multiphase flows in a fluidization process has snowballed in recent years with the rapid development of computer technology. The numerical simulation can provide a better understanding of the flow phenomenon. Which means it can be used as a useful tool for the design and scale-up of fluidized bed [1].

In this Chapter, the numerical approaches for the multiphase flow, the correlations and equations applied in the numerical model. The simple introduction of experimental set-up and method and the simulation procedures will be discussed.

2.2 Numerical methods

The fluidization phenomenon involves gas and solid movement, as well as the interaction between them. Two different methods have been used mainly for simulating gas-solid two-phase flows in fluidized beds: the Eulerian-Lagrangian (EL) approach in which the particle trajectory model is applied, and Eulerian-Eulerian (EE) approach, which is based on the continuum mechanics for both phases [2].

2.2.1 Eulerian-Lagrangian (EL) method

In the EL approach, the gas phase is treated as a continuum, and the solid phase is not treated as a continuum. Solid particles are tracked individually with similar physical and chemical properties [3]. The so-called discrete element method (DEM) is applied to track each particle based on the Lagrangian force balance equation [4]. The equation based on Newton’s law for individual particle is solved with detailed particle-particle and particle-wall collisions. Particle-particle collisions are modelled with the hard-sphere model or the soft sphere model. For the soft sphere model, the particles are allowed to overlap and exert both normal and tangential forces on each other [4]. In the hard-sphere model, the collision is considered at a time and instantaneous, making it more useful for rapid granular flow
Therefore, no additional equations are needed for the solid phase. The method has been applied in many studies of fluidized beds where the solid phase is not dense, and the bed is not large scale [6–9]. Because every particle is tracked in the system, it makes the method high computational cost and time-consuming. Less detailed solution methods with various approximating models have to be utilized when simulating the behavior of the fluid solid system.

2.2.2 Eulerian-Eulerian (EE) method

The Eulerian-Eulerian (EE) approach, in which, both gas and particle phases are treated as interpenetrating continua, and each phase has its governing equations for momentum, continuity, and energy [10]. As a typical Eulerian-Eulerian method, the two-fluid model (TFM) has been widely applied to study the gas-solid dense fluidized bed [11–14]. Because of the particle phase is treated as a continuum, the particle properties, particle-particle interaction and particle-gas interaction need to be defined explicitly. Currently the kinetic theory of granular flow (KTGF) is widely applied to close the governing equations for the solid phase [15,16]. In this theory, the fluctuation energy of particles was described by introducing the concept of granular temperature [17]. The method provides the closures for the solid phase, in which the closures in equation are related to the different stresses, viscosity terms and solids pressure. The two-fluid model has been used to simulate the bubble dynamic in the bubbling fluidized bed systems [11,12,14,18,19]. Compared to Eulerian-Lagrangian method, the Eulerian-Eulerian approach costs less computational time and resource, which is the main reason makes it more favorable for the simulation of large-scale fluidized bed.

Li et.al [20] simulate the cold model of FLUID COKING™ unit by multi Eulerian-Eulerian method. Both FCC and coke particles were used to study the hydrodynamics. The reactor section was designed to be both geometrically and dynamically similar to the commercial FLUID COKING™ unit. The radial voidage profiles were compared with experimental data. Besides, single gas jet and multi gas jets in a bubbling fluidized bed were also simulated by Eulerian-Eulerian method in a three-dimensional fluidized bed. And the jet penetrations as well as the interactions between the jet and the surrounding gas, solids, bubbles, and other jets were investigated [21,22]. The chemical stripping process in fluid
catalytic cracking stripper were also investigated by Eulerian–Eulerian two-fluid model coupled with modified drag model [23]. The gas and solid mixing quality in the stripper section with and without internals were also studied. The residence time distribution model and axial dispersion model were also utilized to obtain the parameters indicating the back-mixing degree [24].

In this work, the 2D TFM KTGF is employed to investigate the hydrodynamic of bubbling and turbulent fluidized bed over a wide range of fluidization velocities, different gas distributor configurations and different baffles inserting in the fluidized bed.

## 2.3 Experiments

### 2.3.1 Configuration of the fluidized bed

The experimental data from in a lab-scale fluidized bed reactor with a rectangular cross-section by Li et al [25] were used to validate the numerical model. The experimental setup shown in Figure 2-1 consists of two sections. The bed thickness is 0.1 m. The height of the fluidized bed unit is 2.18 m, with an expansion in the upper section. The bed width expands from 0.5 m to 1.0 m from the lower section to the expansion section. Initially, the bed was filled with about 100 kg of silica sand with a Sauter-mean diameter of 190 μm. Air at ambient conditions is used as the fluidizing agent. The gas distributor consists of two rows of 10 tuyeres distributed on an angled slope as shown in Figure 2-1, and each tuyere is supplied by a dedicated sonic orifice to maintain the required gas flow rate, which, in this study, was the same for each active tuyere. The minimum fluidization velocity is 0.033 m/s, and the minimum turbulent velocity is 0.6 m/s at 30 °C. The liquid was injected at the height of 1.165 m from the bottom of the bed, as shown in Figure 2-1.

To investigate the effect of the initial gas distribution on bubble distributions in the fluidized bed, three different initial gas distribution cases were considered as shown in Figure 2-2. The base case is the even distributor in which the ten active gas tuyeres are evenly distributed. The second gas distribution is the western distributor with ten active tuyeres near the western side of the column, and the third gas distribution is the eastern distributor with ten active tuyeres near the eastern side of the column.
2.3.2 Measurement method

The triboelectric method, which is a novel measurement technique used to detect the bubble distribution over the cross-section [26], was used by Li et al.[25]. The lateral bubble profile was measured with nine triboelectric probes, which were inserted into the bed by 0.05 m at the same height, as shown in Figure 2-1. The local volumetric flux of bubble gas can be obtained from and the triboelectric signal generated by the impact of the gas bubbles on each probe [27]. The lateral profile of the bubble gas flux was reported with the ratio of the local bubble flux to the average cross-sectional volumetric flux. The equation for this ratio is shown below:

\[
\frac{q_{b,i}}{q_{b,c}} \approx \frac{\epsilon_i - \vartheta_i}{\frac{1}{x_w} \int_0^{x_w} (\epsilon_i - \vartheta_i) \, dx} \quad (1)
\]

Where

\(\epsilon_i\) represents the gas volume fraction

\(\vartheta_i\) represents the local gas velocity

\(x_w\) represents the width of the fluidized bed reactor

Radiation transmission were also applied to measure the gas voidage at different heights in the experiments [28]. Instead of several detectors applied in the literature, only one detector was used in the experiments.
Figure 2-1 Schematic diagram of the lab-scale bubbling fluidized bed [20]
2.4 Simulations

All the simulations were carried out in a two-dimensional domain of the lab-scale fluidized bed set-up shown in using standard TFM with the set of closure laws that most widely applied in the literature.

2.4.1 Inlet condition

Rather than using a uniform flat inlet boundary condition, which was implemented widely in other fluidized bed simulations, a specific gas inlet geometry was used based on the nozzle opening used in the experiments. In the experimental column, two rows of 10 tuyeres were used to supply gas over the whole column depth. For the purpose of this 2D
simulation, a single row of the gas inlet was used to represent each couple of side-by-side tuyeres in the third direction. In the even case, all ten inlets are open to inject gas. For the western case, only five inlets near the western side of the fluidized bed reactor are open. In the eastern case, five inlets near the eastern side are open. To obtain the same superficial velocity in the freeboard, the gas flow rate through each inlet in the eastern and western cases was double the gas flowrate through each inlet in the even case, i.e. the total flow rates for all three cases are the same. Simulations were also performed for two additional distributor configurations for which no experimental data were available, to determine whether such distributors are more effective. The computational domains of fluidized bed reactor (BFB) with and without baffles are shown in Figure 2-3.

2.4.2 CFD model descriptions

A set of basic governing equations consisting of the mass and momentum conservation equations of both gas phase and dispersed phase are used to solve the gas-solid flows in the fluidized bed.

The Euler-Euler approach is applied to simulate the flow in a gas-solid fluidized bed. Gas and solid phases could be present at the same time in the same computational volume by introducing the volume fraction for each phase. The governing equations for the two phases are summarized in In the TFM approach, the interaction between the gas and solid phases is accounted for by the drag force between the two phases. The dynamic balance of particles within the fluidized bed depends on the drag, gravity and buoyancy forces. It is vital to have a drag model that is suitable for the gas-solid fluidization processes under different operation conditions. Several drag force correlations are available. Gidaspow model [10], Syamlal-O’s Briens model [29], Huilin-Gidaspow model [30] and EMMS model [31] are the commonly used drag models. Gidaspow model combined Ergun [32] and Wen-Yu [4] correlation to obtain the drag coefficient, which suits the dense fluidization regime ($\varepsilon < 0.8$) and dilute fluidization regime ($\varepsilon > 0.8$) separately. EMMS is based on the energy minimization multi-scale method. The Syamlal-O’s Briens model obtained the drag model for a multi-particle system form a single particle drag correlation by non-dimensional analysis. Moreover, the Syamlal-O’s Briens model can be adjusted by matching the
predicted minimum fluidization velocity with the experimental data, which is called adjusted the Syamlal-O’s Briens model [2,33], which is also applied in this study [34]. The equations for the drag coefficient are shown in Table 2-3. The Syamlal and O’Brien drag model is based on the single-particle terminal velocity and adjusted based on the fluid properties and the expected minimum fluidization velocity.

Table 2-1 Other constitutive equations for the two-phase flows based on the kinetic theory of granular flow are listed in Table 2-2. The phase-coupled SIMPLE algorithm was applied for the pressure-velocity coupling to solve the mass and momentum conservation equations. The quadratic upwind interpolation for convection kinematics (QUICK) scheme was used to discretize the convection terms in the momentum equations. The modified type of the high-resolution interface capturing (HRIC) scheme was used to estimate the volume fraction of the gas or solid phase. A commercial CFD package (ANSYS Fluent 18.2) was used for the simulations.

The physical properties of gas and particles are specified in Table 2-4. It was assumed that particles are of uniform size, and their diameter is equal to the Sauter mean diameter. The minimum fluidization velocity is 0.033 m/s. The superficial gas velocity consider in this study is from 0.4 m/s to 1.0 m/s. A time step of 0.001s with 100 iterations per time step was chosen in this work. A convergence criterion of 5×10^-4 for each scaled residual component was specified. The simulations were run for 30 s, and the time-averaged values were obtained using the data from the last 20 s since a steady condition was achieved after 10 s.
In the TFM approach, the interaction between the gas and solid phases is accounted for by the drag force between the two phases. The dynamic balance of particles within the fluidized bed depends on the drag, gravity and buoyancy forces. It is vital to have a drag model that is suitable for the gas-solid fluidization processes under different operation conditions. Several drag force correlations are available. Gidaspow model [10], Syamlal-O’s Briens model [29], Huilin-Gidaspow model [30] and EMMS model [31] are the commonly used drag models. Gidaspow model combined Ergun [32] and Wen-Yu [4] correlation to obtain the drag coefficient, which suits the dense fluidization regime ($\varepsilon < 0.8$)

![Schematic of the simulated 2-D baffled fluidized bed](image)

(a) No baffle, (b) Asymmetrical baffle, and (c) Symmetrical baffles

2.4.3 Drag force model
and dilute fluidization regime ($\varepsilon > 0.8$) separately. EMMS is based on the energy minimization multi-scale method. The Syamlal-O’s Briens model obtained the drag model for a multi-particle system from a single particle drag correlation by non-dimensional analysis. Moreover, the Syamlal-O’s Briens model can be adjusted by matching the predicted minimum fluidization velocity with the experimental data, which is called adjusted the Syamlal-O’s Briens model [2,33], which is also applied in this study [34]. The equations for the drag coefficient are shown in Table 2-3. The Syamlal and O’Brien drag model is based on the single-particle terminal velocity and adjusted based on the fluid properties and the expected minimum fluidization velocity.

Table 2-1 Governing equations

| Gas phase |
|--------------------------|------------------------|
| Continuity $\frac{\partial}{\partial t} (\alpha_g \rho_g) + \nabla(\alpha_g \rho_g \vec{v}_g) = 0,$ (2) |
| Momentum $\frac{\partial}{\partial t} (\alpha_g \rho_g \vec{v}_g) + \nabla(\alpha_g \rho_g \vec{v}_g \cdot \vec{v}_g) = -\alpha_g \nabla p + \nabla \cdot \vec{f}_g + \alpha_g \rho_g g + K_{gs}(\vec{v}_g - \vec{v}_g),$ (3) |
| Volume $\alpha_s + \alpha_g = 1,$ |

| Solid phase |
|--------------------------|------------------------|
| Continuity $\frac{\partial}{\partial t} (\alpha_s \rho_s) + \nabla(\alpha_s \rho_s \vec{v}_s) = 0,$ (4) |
| Momentum $\frac{\partial}{\partial t} (\alpha_s \rho_s \vec{v}_s) + \nabla(\alpha_s \rho_s \vec{v}_s \cdot \vec{v}_s) = -\alpha_s \nabla p + \nabla \cdot \vec{f}_s + \alpha_s \rho_s g + K_{gs}(\vec{v}_s - \vec{v}_s),$ (5) |
Table 2-2 Constitutive equations

The granular temperature transport equation:

$$\frac{3}{2} \left[ \frac{\partial}{\partial t} \left( \rho_s \alpha_s \Theta_s \right) + \nabla \cdot \left( \rho_s \alpha_s \vec{v}_s \Theta_s \right) \right] = \left( -p_s \vec{I} + \vec{t}_s \right) \cdot \nabla \Theta_s + \nabla \cdot \left( k_{\Theta_s} \nabla \Theta_s \right) - \gamma_{\Theta_s} + \varphi_{ls}$$ (6)

Where

$$\left( -p_s \vec{I} + \vec{t}_s \right) \cdot \nabla \Theta_s$$ is the generation of energy by the solid stress tensor;

$$k_{\Theta_s} \nabla \Theta_s$$ is the diffusion of energy;

$$\gamma_{\Theta_s}$$ is the collisional dissipation of energy;

$$\varphi_{ls}$$ is the energy exchange between the $l$th solid phase and the $s$th solid phase;

The stress tensor for gas and solid phase are:

$$\bar{t}_g = \alpha_g \mu_g \left( \nabla \vec{v}_g + \nabla \vec{v}_g^T \right) - \frac{2}{3} \mu_g \nabla \cdot \vec{v}_g \vec{I}, \quad (7)$$

$$\bar{t}_s = \alpha_s \mu_s \left( \nabla \vec{v}_s + \nabla \vec{v}_s^T \right) + \alpha_s \left( \lambda_s - \frac{2}{3} \mu_s \right) \nabla \cdot \vec{v}_s \vec{I}, \quad (8)$$

Solid shear viscosity:

$$\mu_s = \mu_{s, \text{col}} + \mu_{s, \text{kin}} + \mu_{s, \text{fr}}, \quad (9)$$

Collisional viscosity:

$$\mu_{s, \text{col}} = \frac{4}{5} \alpha_s \rho_s d_s g_{0,ss} (1 + e_{ss}) \left( \frac{\Theta_s}{\pi} \right)^{1/2} \alpha_s, \quad (10)$$

Kinetic viscosity:

$$\mu_{s, \text{kin}} = \frac{\alpha_s d_s \rho_s \sqrt{\Theta_s \pi}}{6(3 - e_{ss})} \left[ 1 + \frac{2}{5} (1 + e_{ss}) (3e_{ss} - 1) \alpha_s g_{0,ss} \right], \quad (11)$$
Frictional viscosity:

\[
\mu_{s,fr} = \frac{p_{\text{friction}} \sin \varphi}{2 \sqrt{l_2 D}},
\]  

(12)

Solid bulk viscosity:

\[
\lambda_s = \frac{4}{3} \alpha_s^2 \rho_s d_s \sigma_{0,ss}(1 + e_{ss}) \left(\frac{\Theta_s}{\pi}\right)^{\frac{1}{2}},
\]  

(13)

Solid pressure:

\[
p_s = \alpha_s \rho_s \Theta_s + 2 \rho_s (1 + e_{ss}) \alpha_s^2 \sigma_{0,ss} \Theta_s,
\]  

(14)

Radial distribution function:

\[
g_{0,s} = \left[1 - \left(\frac{\alpha_s}{\alpha_{s,\text{max}}}ight)^{\frac{1}{3}}\right]^{-1},
\]  

(15)

Diffusion coefficient of granular temperature (Syamlal-O’Brien):

\[
k_{\Theta_s} = \frac{15 d_s \rho_s \alpha_s \sqrt{\Theta_s \pi}}{4(41 - 33 \eta)} \left[1 + \frac{12}{5} \eta^2 (4 \eta - 3) \alpha_s \sigma_{0,ss} + \frac{16}{15 \pi} (41 - 33 \eta) \eta \alpha_s \sigma_{0,ss}\right]
\]  

(16)

\[
\eta = \frac{1}{2} (1 + e_{ss}),
\]  

(17)
Table 2-3 Momentum exchange coefficient

Syamlal–O’Brien drag function [29]

\[ K_{gs} = \frac{3\alpha_s \alpha_g p_g}{4v_f^2 d_s} C_D \left( \frac{Re_s}{\nu_{r,s}} \right) |\vec{v}_s - \vec{v}_g|, \]  

\[ \nu_{r,s} = 0.5 \left( A - 0.06 Re_s + \sqrt{(0.06 Re_s)^2 + 0.12 Re_s (2B - A) + A^2} \right), \]

\[ A = \alpha_g^{4.14}, \]

\[ B = c_1 \cdot \alpha_g^{1.28} \text{ for } \alpha_g \leq 0.85, \]

\[ B = \alpha_g^{d_1} \text{ for } \alpha_g > 0.85, \]

Where \( d_1 = 1.28 + \frac{\log_{10} c_1}{\log_{10} 0.85} \)

\[ C_D = \left( 0.63 + \frac{4.8}{\sqrt{\left( \frac{Re_s}{\nu_{r,s}} \right)}} \right), \]

\[ Re_s = \frac{\rho_g d_s |\vec{v}_s - \vec{v}_g|}{\mu_g}, \]

Where \( C_D \) is the drag coefficient and \( Re_s \) is the Reynolds number.
Table 2-4 Summary of physical properties of the reactor, particles and gas

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<td>$U_{mf}$</td>
<td>Minimum fluidization velocity [m/s]</td>
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<tr>
<td>$P$</td>
<td>Operation pressure [atm]</td>
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</tr>
<tr>
<td>$T$</td>
<td>Operation temperature [℃]</td>
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</tr>
<tr>
<td>$\varepsilon_{s0}$</td>
<td>Initial solids packing</td>
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<table>
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<tr>
<td>$\mu_g$</td>
<td>Shear viscosity [kg/ms]</td>
<td>1.85×10⁻⁵</td>
</tr>
<tr>
<td>$U_g$</td>
<td>Superficial velocity [m/s]</td>
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</tbody>
</table>

<table>
<thead>
<tr>
<th>Particles</th>
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<tbody>
<tr>
<td>$\rho_s$</td>
<td>Solid density [kg/m³]</td>
<td>2650</td>
</tr>
<tr>
<td>$d_s$</td>
<td>Sauter mean diameter [μm]</td>
<td>190</td>
</tr>
<tr>
<td></td>
<td>Particle-particle restitution coefficient</td>
<td>0.95</td>
</tr>
<tr>
<td></td>
<td>Specularity coefficient</td>
<td>0.0001</td>
</tr>
</tbody>
</table>

42
2.4.4 Boundary and initial conditions

In the inlet at each tuyere, the uniform gas velocity inlet condition is used. The inlet gas velocity was specified based on the superficial gas velocity used in the experiment. The atmosphere pressure was selected at the outlet boundary condition for the reactor. No-slip boundary conditions for the gas phase, and Johnson and Jackson [35] slip boundary conditions for the solid phase were used.

\[
\overline{U}_{SW} = -\frac{\mu_s \alpha_{s_{max}}}{\sqrt{3} \pi \rho_s \phi \alpha_s g_{0,ss} \sqrt{\alpha_s}} \frac{\partial \overline{U}_{SW}}{\partial n},
\]

(20)

The specularity coefficient \( \phi \) is an empirical parameter that represents the particle-wall collision. The value of the specularity coefficient depends on the wall roughness \( \phi=0 \) means a perfect specular collision, and \( \phi=1 \) means perfectly diffusion collusion. The value of 0.0001 is chosen based on an earlier study [31]. The details of the boundary conditions for the gas and solid phases are listed in Table 2-5.

In the 2D simulation work, the initial conditions specify the concentration of the solid bed, and the settled bed was 1.60 m deep, and the initial solids volume fraction was defined as 0.60. The upper section of the reactor was considered to be occupied by gas only at t=0.

<table>
<thead>
<tr>
<th>Table 2-5 Boundary conditions</th>
</tr>
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<tbody>
<tr>
<td><strong>Inlet of gas phase</strong></td>
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<tr>
<td>Superficial gas velocity</td>
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<table>
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<tr>
<th><strong>Wall</strong></th>
</tr>
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<tbody>
<tr>
<td>Gas-phase</td>
</tr>
<tr>
<td>No-slip velocity</td>
</tr>
<tr>
<td>Solid-phase</td>
</tr>
<tr>
<td>Partial-slip</td>
</tr>
<tr>
<td>Specularity coefficient: 0.0001</td>
</tr>
<tr>
<td>Particle-wall restitution coefficient: 0.9</td>
</tr>
</tbody>
</table>
Outlet

<table>
<thead>
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<th>Condition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>Pressure-outlet</td>
</tr>
<tr>
<td>Solids</td>
<td>Pressure-outlet</td>
</tr>
</tbody>
</table>

Baffle

<table>
<thead>
<tr>
<th>Phase</th>
<th>Condition</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gas</td>
<td>No-slip velocity</td>
</tr>
<tr>
<td>Solid</td>
<td>Partial-slip</td>
</tr>
</tbody>
</table>

Specularity coefficient: 0.0001

Particle-wall restitution coefficient: 0.9

2.4.5 Local gas bubble profile

In experiments, the triboelectric probes were applied to measure the local gas bubble flux in a dry bed as mentioned in the measurement method. To validate the numerical model, similar parameters were obtained in numerical calculation, and compared to experimental data. The total gas flux is equal to the local gas voidage multiply the local gas velocity. The total gas flux is the sum of the emulsion phase flux and the bubble phase flux. The solids are Geldrat’s group B particles and their minimum fluidization velocity is 0.033 m/s. Over the range of superficial gas velocities explored in this thesis, the bubble phase flux is at least 91.75 % of the total gas flux. The local gas bubble flux was, therefore, assumed to be the same as the total local gas flux.

The profile of the gas bubble flux is calculated by:

\[
\frac{q_{bl}}{\overline{q}_b} \approx \varepsilon_i \cdot \vartheta_i \tag{21}
\]

\[
q_{bl} \approx \varepsilon_i \cdot \vartheta_i \tag{22}
\]

\[
\overline{q}_b = \frac{1}{x_w} \cdot \int_0^{x_w} q_{bl} \cdot dx \tag{23}
\]
Where

\( \varepsilon_l \) local gas voidage

\( \theta_l \) local gas velocity

### 2.5 References


Chapter 3

3 A fundamental study of the numerical model for the simulation of a gas-solid bubbling fluidized bed

3.1 Introduction

Gas-solid fluidized bed reactors have been widely used in chemical, pharmaceutical, energy, petroleum, and other industries owing to their inherent benefits, such as high rates of heat and mass transfer and quick mixing of solids [1]. However, these benefits highly depend on the quality of the fluidization process. A proper design of the gas distributor plays a crucial role in improving the uniformity of gas-solid distributions in a fluidized bed since the flow hydrodynamics in the bottom zone of the fluidized bed has a significant influence on the bed performance [2–5]. In the Fluid Coking™ process, undesired agglomerates will decrease the heat and mass transfer rates, thereby causing operating problems [6–8]. The bubble flux distribution in the cross-section of the fluidized bed plays a vital role in the agglomeration [9–11]. The most considerable risk in biomass fuel gasification is the defluidization and plugging of nozzles and drains caused by particle agglomeration, which means the hydrodynamics of the gas-solid mixture inside the fluidized bed has a significant impact on the gasification process [12].

In commercial units, like thermal cracking or fluid coking process, the fluidized bed is filled with hot coke particles, and the hydrocarbon is steam-atomized and injected horizontally into the bed through multiple nozzles. However, during the experimental study, it is always an issue to whether to use the more controllable and measurable particles or use those in actual high-temperature full-scale process. Song et al. [13] investigated the hydrodynamics of a fluidized bed reactor, which is scaled down commercial units. Both FCC particles and fluid coke particles were operated under the same conditions, and similar voidage distribution and solids momentum flux distribution is observed. However, in their study, the difference of particle properties between FCC ($\rho_p = 1700 \text{ kg/m}^3$, $\bar{d}_p = 99 \mu m$) and fluid coke ($\rho_p = 1600 \text{ kg/m}^3$, $\bar{d}_p = 133 \mu m$) are not obvious, which could be one reason for the similar results. Qi et al [14] used sand particles and FCC as fluidized material to study the combined effect of particle properties and nozzle gas distributor
design in two risers. The researchers found that in the fully developed section, the radial
distribution is more uniform with sand particles and FCC particles are more uniform in the
axial distribution. However, there is no further study or explanation for this phenomenon.
Moreover, in the experimental method, it is not easy to control variables to determine the
main reason for the difference.

There are several measurement methods applied in the experiments to determine the
transition from bubbling to turbulent fluidization. Pressure based measurements like
pressure fluctuation are one of the most widely used parameters to detect the regime
transition velocities [15–20]. Like the optical fiber probe [21], electrostatic probe and novel
dual-tip probes [22]. There is also some non-intrusive technique applied to measure
fluidized bed behaviour [23–26]. Tebianian et al. [23] visually detect the turbulent
fluidization regime by the X-ray system. Azizpour et al. [24] found out that the vibration
signals can reflect the bubble movement in the fluidized bed and the maximum value of
the Hurst exponent means the onset of the turbulent fluidization. Nedeltchev et al. [25]
using the radioactive particle tracking (RPT) technique to determine the minimum
turbulent velocity. Zhou et al. [26] investigated the non-instructive acoustic emission
technique with the standard deviation and multi-scale analyses to identify the regime
transitions in the gas-solid two-phase fluidization. Statistical parameters like the standard
deviation of acoustic signals can effectively reflect the transition velocities. CFD as a
useful tool to study the fluidization properties. Not many researchers use it to detect
different fluidization regimes. In this chapter, it may be interesting to find out some
representative parameters to reflect the regime transition as superficial gas velocity
increases.

3.2 The objective of this work

1. Develop an accurate numerical model to simulate the gas-solid two-phase flows in
the fluidized beds.

2. Check the possibility of using different particles and gas to get similar
hydrodynamic behaviours for the operation condition study.
3. Validate the proposed numerical model by comparing the numerical results with the experimental data.

4. Investigate the effect of the superficial gas velocity on the gas-solid behaviour and the transition from the bubbling regime to a turbulent regime.
Figure 3-1 Schematic diagram of the lab-scale bubbling fluidized bed [10]
Results and discussion

The configuration of the bubbling fluidized bed used in this study is shown in Figure 3-1. There are ten air jets at the inlet configuration part. In experimental work, the author compared three different inlet gas distributor configurations, which is shown in Figure 3-2. In order to validate the numerical model with experimental results, these three inlet gas distributor configurations will all be considered in the validation part.

3.3.1 Grid size independency

In the CFD simulation, it is necessary to ensure that the grid size is appropriate. A grid size sensitivity test was performed using three grid resolutions. The mesh intervals spacing were...
5, 4, and 2 mm, respectively. All the simulations for the grid-independent tests were carried out at the same superficial gas velocity 0.4 m/s for the Even gas distributor case. Table 3-1 shows the predictions of the pressure drop across the bed using three different mesh sizes, in which the pressure difference between mesh-2 and mesh-1 is smaller than mesh-2 and mesh-3. The time-averaged gas bubble flux profile in the injection level is shown for different grid sizes in Figure 3-3. Based on the comparison, it can be seen that the difference in the results between mesh-1 and mesh-2 is small enough. Therefore, the medium size mesh (102,221 cells) was used for the rest of the study.

**Table 3-1 Grid independent test**

<table>
<thead>
<tr>
<th>Name</th>
<th>Mesh interval spacing (mm)</th>
<th>Grid Nodes</th>
<th>Grid Cells</th>
<th>Bed Pressure drop (Pa)</th>
<th>Error (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mesh-1</td>
<td>2×2</td>
<td>226854</td>
<td>225458</td>
<td>7920</td>
<td>0.31</td>
</tr>
<tr>
<td>Mesh-2</td>
<td>5×2</td>
<td>102983</td>
<td>102221</td>
<td>7945</td>
<td>0.9976</td>
</tr>
<tr>
<td>Mesh-3</td>
<td>5×4</td>
<td>54808</td>
<td>54171</td>
<td>8025</td>
<td></td>
</tr>
</tbody>
</table>
For the dense gas-solid fluidization, the sensitivity of numerical results to the selection of turbulence and laminar model has been studied by many researchers. Liang et.al [27] compared the large eddy simulation (LES) turbulence model with the baseline model and found out that the results on solid hold up and velocity are nearly same which means the effect of gas turbulence is insignificant in TFM simulations. Similar conclusions were also achieved by other researchers [28–31]. Adnan et.al [32] found out that the laminar models showed more consistent results over turbulent models based on 2D simulations. Shi et.al [33] not limited to the sensitive of turbulent model on the numerical results, but also studied the turbulent viscosity ratio inside the fluidized bed in both two-dimensional and three-dimensional numerical model. They conclude that conclude that the laminar model can be

Figure 3-3 Lateral gas bubble distribution profiles at the injection level with different grid sizes under the superficial gas velocity of 0.4 m/s

3.3.2 Laminar model vs. turbulent model

For the dense gas-solid fluidization, the sensitivity of numerical results to the selection of turbulence and laminar model has been studied by many researchers. Liang et.al [27] compared the large eddy simulation (LES) turbulence model with the baseline model and found out that the results on solid hold up and velocity are nearly same which means the effect of gas turbulence is insignificant in TFM simulations. Similar conclusions were also achieved by other researchers [28–31]. Adnan et.al [32] found out that the laminar models showed more consistent results over turbulent models based on 2D simulations. Shi et.al [33] not limited to the sensitive of turbulent model on the numerical results, but also studied the turbulent viscosity ratio inside the fluidized bed in both two-dimensional and three-dimensional numerical model. They conclude that conclude that the laminar model can be
used when the isothermal fluidized bed is considered. Similar methods are applied in this chapter to select the suitable model.

The effect of turbulence on computational fluid dynamics modelling of fluidized bed reactors is analyzed by comparing the results obtained from laminar and turbulent cases. The turbulent viscosity ratio contour at 20 s (Figure 3-4) using an unsteady RANS $k - \varepsilon$ model showed that the turbulent viscosity is low in most of the bed. At the same time, more substantial turbulence develops above the bed surface near the outlet. Therefore, it is necessary to assess whether the laminar or turbulent model should be used.

The particle volume fraction distributions predicted by the laminar and turbulent models are compared with the experimental data, as shown in Figure 3-5. It is easy to notice that the particle distribution from the laminar model is more even, and the expanded bed height is higher than that from the turbulent model. To illustrate qualitative differences between the results from the laminar model and turbulent model, Figure 3-6 shows the comparisons between the results from the two models and the experimental results for the gas bubble profile on the injection level. The gas bubble volume fractions from both the experimental and laminar results are concentrated on the western side of the bed, and the peak point locations are close, and the gas bubble distribution from the turbulent model is flatter compared to others and concentrated on the other side of the bed. Because the result from the laminar model is closer to the experimental data due to the low-speed flow, the laminar model is applied in the rest of this work.
Figure 3-4 Contour of the turbulent viscosity ratio under the superficial gas velocity of 0.4 m/s
Figure 3-5 Comparison of the time-averaged gas volume fraction contours between the laminar model and turbulent model in the fluidized bed reactor under the superficial gas velocity of 0.4 m/s
Figure 3-6 Comparison of lateral gas bubble flux distribution profiles between the numerical and experimental results under the superficial gas velocity of 0.4 m/s

3.3.3 Model validation
Western

Even

\( \frac{q_{bi}}{q_b} \)

Lateral location, cm

Experimental data

Numerical data

Lateral location, cm

Experimental data

Numerical data
(b)
In order to validate the numerical model, the numerical results are compared to the experimental data obtained using two different measurement methods, which are E-probe and radiation transmission methods [27] Three inlet distributor configurations are used in this study, the Eastern, Western and Even configurations as shown in Figure 3-1 and Figure 3-2. The base case is the Even distributor in which the ten active gas tuyeres are evenly distributed. The second gas distribution is the Western distributor with ten active tuyeres near the Western side of the column, and the third gas distribution is the Eastern distributor with ten active tuyeres near the eastern side of the column. Figure 3-7 shows the numerical results of the gas flux profile on the injection level compared with experimental results for three different inlet gas distributor configurations under different superficial gas velocities. For the Even configuration, both the predicted and measured profiles indicate a moderate lateral variation in the gas bubble flux, but the flux is lower on the right-hand side (the eastern side), which is likely caused by the sloped distributor. The predicted results are in

Figure 3-7 Comparison of the numerical and experimental results for the radial gas bubble distributions at the injection level of the bubbling fluidized bed under superficial gas velocities of (a) 0.40 m/s, (b) 0.60 m/s, (c) 0.80 m/s (d) 1.0 m/s

In order to validate the numerical model, the numerical results are compared to the experimental data obtained using two different measurement methods, which are E-probe and radiation transmission methods [27] Three inlet distributor configurations are used in this study, the Eastern, Western and Even configurations as shown in Figure 3-1 and Figure 3-2. The base case is the Even distributor in which the ten active gas tuyeres are evenly distributed. The second gas distribution is the Western distributor with ten active tuyeres near the Western side of the column, and the third gas distribution is the Eastern distributor with ten active tuyeres near the eastern side of the column. Figure 3-7 shows the numerical results of the gas flux profile on the injection level compared with experimental results for three different inlet gas distributor configurations under different superficial gas velocities. For the Even configuration, both the predicted and measured profiles indicate a moderate lateral variation in the gas bubble flux, but the flux is lower on the right-hand side (the eastern side), which is likely caused by the sloped distributor. The predicted results are in
good agreement with the experimental data at higher superficial gas velocities (0.8 m/s and
1.0 m/s). In the Eastern configuration, the predicted and measured profiles show that the
gas bubbles are concentrated on the eastern side, and both the predicted and measured
bubble fluxes peak at around 35 cm in the lateral direction. In the Western gas inlet
distributor configuration, the predicted and measured profiles show that the gas bubbles
are concentrated on the western side, and both predicted and measured bubble fluxes peak
at around 10 cm in the lateral direction. Quantitative discrepancies are noticed between the
predicted and measured data for the two points near the western side of the wall, which
might be due to either the influence of boundary conditions used in numerical simulation
or the effect of the wall on the triboelectric probe during experiments. In all cases, the
predicted bubble flux variation is slightly smaller than that from the experimental data.
However, the general gas bubble distribution tendency is consistent with the experimental
observation.

Apart from comparing the bubble profile on the injection level with E-probe, the results at
the same height with different lateral locations are also compared with those from the
radiation transmission method. In Figure 3-8, the gas volumetric flux value at 10 cm, 25
cm, and 40 cm were compared with the results from E-probe and gas voidage results from
the radiation transmission method. In the Even configuration, the numerical values are very
close to the experimental ones. The same peak point location is observed in both the Even
and Western configurations. In the Eastern configuration, the numerical value at 40 cm is
lower compared to the experimental data, which makes the peak point in the numerical
model moves to 25cm. The values near the wall tend to be small in the numerical results.
The value in the center part of the bed agrees well with experimental results.
Effect of gas and particle properties

In commercial units, like thermal cracking or fluid coking process, the fluidized bed is filled with hot coke particles, and the hydrocarbon is steam-atomized and injected horizontally into the bed through multiple nozzles. However, in the experimental study, there is always a debate about whether to use the more controllable and measurable particles or use those used in the actual high-temperature full-scale process. In this work, fluidization experiments were carried out in the lab-scale set-up with dry compressed air and silica sand. The objective of this study is to find out whether the hydrodynamic behaviors predicted by using different particles and gases between the lab condition and commercial conditions are similar.

The gas and particle properties under these two different conditions are listed in Table 3-2. The results under these two conditions are shown in Figure 3-9. One is under the lab condition (dry compressed air and silica sand), which is shown in the solid line. The other
is under the commercial condition (vaporized hydrocarbons and steam and coke particles) shown in the dashed line. The comparison is based on the gas bubble profiles at different heights along the bed under the same superficial gas velocity of 0.60 m/s. For the two different conditions, similar profiles are found at three different heights. Figure 3-9 shows that the difference in the lateral gas bubble distributions between the two conditions is tiny at $H = 0.9$ m, and it increases slightly at $H = 0.95$ and 1.0 m. At $H = 1.0$ m, the peak value is at about $x=15$ cm in the lateral direction under the lab condition, and the peak point is at around $x = 25$ cm under the commercial condition, as shown in Figure 3-9 (a). The locations of the peak points under these two conditions are getting closer at a lower height.

At $H=0.9$ m, the gas bubble flux profiles between the two conditions are almost identical, and the locations of the peak values from those two conditions are almost the same, as shown in Figure 3-9 (c). Furthermore, the peak value under the commercial condition is slightly higher than the one under the lab condition at a higher height, and they are almost the same at $H=0.9$ m.

The gas hold up profiles and gas velocity profiles between the commercial condition and lab condition at different heights are compared and given in Figure 3-10 and Figure 3-11. The gas volume fraction under the commercial condition is always higher than that under the lab condition, as shown in Figure 3-10. Moreover, the difference is getting more significant at higher heights. It can be seen from Figure 3-11 that the difference in the gas velocity profiles between the two conditions is not as noticeable as the voidage profiles at different heights.

In addition to the lab condition and commercial condition, two more conditions, dry compressed air + coke particles and vaporized hydrocarbons + silica sand, are also used to determine the possible reasons for the different conditions. Comparing the lateral gas bubble distributions under the four different conditions is present in Figure 3-12. In general, the four gas bubble flux profiles under the four different conditions at $H=1.0$ m are quite similar. However, the difference between the cases with the same solid particles is larger than that between the cases with the same gas, i.e. the gas bubble profiles for the cases of air with sand and coke particles have a similar flux profile, the same for the cases of the
vaporized hydrocarbons and steam with sand and coke particles. Therefore, the gas properties influence more on the gas bubble distributions than the particle properties. In general, the gas bubble flux distribution profiles are quite similar for both the lab condition and commercial condition.

<table>
<thead>
<tr>
<th>Table 3-2 Summary of gas and particle properties</th>
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<tbody>
<tr>
<td><strong>Gas</strong></td>
</tr>
<tr>
<td>Gas density, $\rho_g$, kg/m$^3$</td>
</tr>
<tr>
<td>Gas viscosity, $\mu$, Pa.s</td>
</tr>
<tr>
<td><strong>Particle</strong></td>
</tr>
<tr>
<td>Particle density, $\rho_p$, kg/m$^3$</td>
</tr>
<tr>
<td>Geldart powder group</td>
</tr>
<tr>
<td>Minimum gas velocity, m/s</td>
</tr>
</tbody>
</table>
Table 3-3 Operation conditions for gas and particle properties

<table>
<thead>
<tr>
<th>Condition</th>
<th>Gas</th>
<th>Particle</th>
</tr>
</thead>
<tbody>
<tr>
<td>Lab condition</td>
<td>Dry compressed air</td>
<td>Silica sand</td>
</tr>
<tr>
<td>Commercial condition</td>
<td>Vaporized hydrocarbons and steam</td>
<td>Coke particles</td>
</tr>
<tr>
<td>Compared to condition 1</td>
<td>Dry compressed air</td>
<td>Coke particles</td>
</tr>
<tr>
<td>Compared to condition 2</td>
<td>Vaporized hydrocarbons and steam</td>
<td>Silica sand</td>
</tr>
</tbody>
</table>
(a) Lab condition
Commerical condition

(b) Lab condition
Commerical condition
Figure 3-9 Comparison of lateral gas bubble distributions at different heights of air+sand and hydro+coke cases under the superficial gas velocity of 0.60 m/s (a) $H_1 = 1.00$ m; (b) $H_2 = 0.95$ m; (c) $H_3 = 0.90$ m
Figure 3-10 Comparison of lateral gas volume fractions at different heights between air+sand and hydro+coke cases under the superficial gas velocity of 0.60 m/s: (a) H₁ = 1.00 m; (b) H₂ = 0.95 m; (c) H₃ = 0.90 m
(a)

(b)
Figure 3-11 Comparison of lateral gas velocities at different heights between air+sand and hydro+coke cases under the superficial gas velocity of 0.60 m/s (a) $H_1 = 1.00$ m; (b) $H_2 = 0.95$ m; (c) $H_3 = 0.90$ m
Effect of the superficial gas velocity

When the gas velocity continuously increases, the fluidized bed also goes through different fluidization regimes, from the fixed bed, particulate fluidization, bubbling fluidization, slugging fluidization to turbulent fluidization. Moreover, there are also different transition points. The transition point from a fixed bed to a bubbling bed is when the superficial gas velocity is equal to the minimum fluidization velocity \( U_{mf} \) for group B and Group D particles. During bubbling fluidization, with the gas velocity continuously increases, the bubbles grow more prominent, and then the bed can transfer into a slugging bed if the bed diameter is small and particle size is big or into a turbulent bed if the bed diameter is large. The velocity, denoted as \( U_k \), is the transition point when large bubbles start to break up into small bubbles. However, in a specific regime, bubble profiles also change when the superficial gas velocity changes.

Figure 3-12 Comparison of lateral gas bubble distributions of air+sand and hydro+coke, air+coke and hydro+coke cases under the superficial gas velocity of 0.60 m/s at \( H_1 = 1.00 \) m

3.3.5 Effect of the superficial gas velocity

When the gas velocity continuously increases, the fluidized bed also goes through different fluidization regimes, from the fixed bed, particulate fluidization, bubbling fluidization, slugging fluidization to turbulent fluidization. Moreover, there are also different transition points. The transition point from a fixed bed to a bubbling bed is when the superficial gas velocity is equal to the minimum fluidization velocity \( U_{mf} \) for group B and Group D particles. During bubbling fluidization, with the gas velocity continuously increases, the bubbles grow more prominent, and then the bed can transfer into a slugging bed if the bed diameter is small and particle size is big or into a turbulent bed if the bed diameter is large. The velocity, denoted as \( U_k \), is the transition point when large bubbles start to break up into small bubbles. However, in a specific regime, bubble profiles also change when the superficial gas velocity changes.
Figure 3-13 (a) shows the gas bubble flux distributions under different superficial gas velocities at H=0.85 m. When the gas velocity is below 0.65 m/s, the gas bubble profiles are similar, and the peak points are at around x = 15 cm. With an increase in the gas velocity, the gas bubble profiles are different at different gas velocities and become more irregular. From Figure 3-13 (b), it can be seen that at a higher location in the fluidized bed, the bubble profiles become similar between different superficial gas velocities and becomes more irregular at higher gas velocities. However, 0.65 m/s is still a turning point when the peak point of the gas flux distribution changes from the left to the right side of the bed, shown in Figure 3-13 (b). Moreover, the gas voidage profiles along the lateral direction under different velocities at different heights are shown in Figure 3-14. Generally, the bubble volume fraction increases with the increase in the superficial gas velocity. In a lower height (H<sub>1</sub> = 0.85 m) as shown in Figure 3-14 (a), the shapes of the profiles are similar under different superficial gas velocities, the gas voidage is high in the middle and low near the wall, it increases with the increase in the gas velocity. In higher heights, as shown in Figure 3-14 (b)-(d), the profiles have different shapes at different superficial gas velocities.

Figure 3-15 shows the lateral gas velocity distributions under different superficial gas velocities and different heights. It can be seen that when the superficial gas velocity is below 0.65 m/s, the velocity distribution is similar at different heights. The peak point occurs at around x = 15 cm for H =0.85 to 0.95 m as shown in Figure 3-15 (a-c). As the superficial gas velocity increases, the peak point of the gas velocity moves to the right-hand side of the bed. It can be indicated as one evidence that the minimum turbulent velocity is between 0.65 m/s and 0.7 m/s. As a different gas bubble flux, the gas velocity profile in horizontal direction changed obviously between 0.65 m/s and 0.7 m/s. Figure 3-16 shows the bubble volume fraction profiles along the vertical direction under different gas velocities. Below H = 0.5 m, because of the sloped inlet gas distributor, the bubble volume fraction fluctuates a lot in the vertical direction. The bubble volume fraction distribution is almost uniform along the vertical direction in the fluidized bed reactor between H = 0.5 and 1.65 m. Above H =1.65 m, the bed expands, and the gas hold up also begins to increase. The bubble volume fraction increases with the increase in
the superficial gas velocity from 0.4 m/s to 0.6 m/s. Between the superficial gas velocity of 0.65 m/s and 0.8 m/s, the bubble volume fraction does not change with the superficial gas velocity. However, if the gas velocity is increased to 1.0 m/s, the bubble volume fraction suddenly increases to around 0.8 at H=0.5 to 1.65.

Figure 3-17 shows the contours of the bubble volume fraction in the fluidized bed under different superficial gas velocities. It can be seen from the contours that the gas volume fraction distributions when the superficial gas velocity is lower than 0.65 m/s is very different from those when the superficial gas velocity is higher than 0.65 m/s. The bubble volume fraction is around 0.8 in almost the entire bed when the superficial gas velocity is higher than 0.65, as shown in Figure 3-17 (d-f), which indicates that the bed is full of bubbles. However, when the superficial gas velocity is lower than 0.65, the bubble volume fraction is much lower.

(a) \( H_1 = 0.85 \) m
(b) $H_2 = 0.90 \text{ m}$

(c) $H_3 = 0.95 \text{ m}$
Figure 3-13 Comparison of lateral gas bubble distributions at different heights under different superficial gas velocities
(b) Gas volume fraction vs. lateral location, cm

(c) Gas volume fraction vs. lateral location, cm
Figure 3-14 Comparison of lateral gas volume fraction distributions at different heights under different superficial gas velocities (a) $H_1 = 0.85$ m; (b) $H_2 = 0.90$ m; (c) $H_3 = 0.95$ m (d) $H_4 = 1.00$ m
Figure 3-15 Comparison of lateral gas velocity distributions at different heights under different superficial gas velocities (a) $H_1 = 0.85$ m; (b) $H_2 = 0.90$ m; (c) $H_3 = 0.95$ m (d) $H_4 = 1.00$ m
Figure 3-16 Comparison of the vertical gas hold up distributions under different superficial gas velocities
Figure 3-17 Time-averaged bubble volume contours under different superficial gas velocities (a) $U_1 = 0.3$ m/s; (b) $U_2 = 0.4$ m/s; (c) $U_3 = 0.5$ m/s (d) $U_4 = 0.6$ m/s (e) $U_5 = 0.8$ m/s (f) $U_6 = 1.0$ m/s
3.4 Conclusion

A multi-phase Eulerian-Eulerian two-fluid method (TFM) coupled with the kinetic theory of granular flow (KTGF) can successfully predict the impact of the inlet gas distribution on the lateral profile of the gas bubble flux, at a level well above the gas distributor level.

The gas bubble distribution on the injection level was similar enough that the gas and particle in lab condition can be used for studying the fluid coking process.

The superficial gas velocity influences the gas bubble lateral distribution and vertical distribution. Increasing gas velocity will increase the gas volume fraction inside the fluidized bed and increase the intensity of the gas bubble movement. Moreover, the potential minimum turbulent fluidization velocity $U_{mt}$ is between 0.65 m/s and 0.7 m/s.

3.5 References


Chapter 4

4 Effect of the inlet gas distributor on the bubble distribution in a fluidized bed

4.1 Introduction

Fluidized bed systems are widely used in many industrial applications due to their high mixing and uniform temperature distribution. Most industrial fluidized bed reactors operate in the bubbling or turbulent fluidization regimes[1]. The behavior of bubbles, also called “voids” for turbulent beds dominates the fluidized bed performance. On one side, particles mixing in the fluidized bed is entirely dominated by gas bubbles, which carry solids in their wake. On the other side, bubbles are also one of the reasons for solids entrainment and back mixing of solids and gases [2,3]. A properly designed gas distributor can let to a uniform and stable fluidization across the entire bed and avoid nonfluidized regions on the grid [2,4–7]. The flow hydrodynamics in the bottom zone of the fluidized bed has a significant influence on the bed performance [8–11]. In the Fluid Coking™ process, undesired agglomerates will decrease heat and mass transfer rates to the heavy oil feedstock, thereby delaying its residence time and causing operating problems [12–14]. The bubble flux distribution in the cross-section of the fluidized bed plays a vital role in agglomerate formation [15–17]. In biomass fuel gasification, the most considerable risk is defluidization and the plugging of nozzles and drains that results from agglomerate formation. Therefore, the fluidized bed hydrodynamics have a significant impact on the gasification process [18].

The gas distributor can affect the bed hydrodynamics. Chen et al. [8] used CFD to study the effect of the distributor shape on the flow behavior by comparing plane and triangle distributors. Although there was not much difference in the overall fluidization performance using both distributors, solid particles were fluidized more efficiently with the triangle distributor. The voidage profile could be flattened using a bubble cap distributor instead of a perforated plate distributor, over a wide range of gas velocities, in both bubbling and turbulent regimes [9]. Peng et al. [10] found that in the 2D riser, the gas distributor geometry affects the fully developed lateral profiles of the solids holdup and
velocity. Inclined gas distributors can be used to facilitate the discharge of the bed particles, and Cai et al. [3] found that increasing the angle of such a distributor can enhance the heterogeneity of the transverse particle and bubble movements in the bottom zone of a fluidized bed.

The objective of this chapter is to use the validated numerical model from chapter 3 to study the effects of the gas distributor geometry on the gas flow patterns in a bubbling fluidized bed.

4.2 Equipment description

Numerical simulations are performed for a lab-scale bubbling fluidized bed reactor, which as a rectangular cross-section shown in Figure 4-1. It consists of two sections. The bed thickness is 0.1 m. The height of the fluidized bed unit is 2.18 m, with an expansion in the upper section. The bed width expands from 0.5 m to 1.0 m from the lower section to the expansion section. The bed is filled with 100 kg of silica sand particles with a Sauter-mean diameter of 190 microns. Air at ambient conditions is used as the fluidizing agent. Figure 4-1 also shows the “injection level” where the liquid would be injected in a corresponding experimental setup; However, the liquid injection is neglected in the numerical simulations in this study. But, knowing the lateral gas profile at that level would be useful to optimize a spray nozzle location [19].

All the simulations are carried out in a two-dimensional domain of a lab-scale fluidized bed unit shown in Figure 4-2. In all configurations, the gas flowrate is evenly distributed between all inlets. The base configuration, which is the Even configuration has ten gas inlets uniformly distributed along the sloped distributor. The Western configuration has five inlets near the western side of the fluidized bed reactor, and the Eastern configuration has five inlets near the eastern side. To obtain the same superficial velocity in the freeboard, the gas flow rate through each inlet in the Eastern and Western configurations is doubled the gas flowrate through each inlet in the Even configuration. Simulations are also performed for two additional distributor configurations. Figure 4-2 shows the five different gas distributors. Finally, different distributor inclined angles, 45°, 30°, 0° as shown in Figure 4-3, are also investigated in this study.
Figure 4-1 Schematic diagram of the lab-scale bubbling fluidized bed [16]
Figure 4-2. Different gas inlet distributors used in the two-dimensional simulation model
Figure 4-3 Different gas inlet distributors angles used in the simulations

4.3 Results and discussion
4.3.1 Effect of the gas distributor configuration

\[ \frac{q_{bi}}{q_{b\text{h}}} \]

(a) \( u_g = 0.4 \text{ m/s} \)

Western Even Eastern

(b) \( u_g = 0.6 \text{ m/s} \)

Western Even Eastern

Lateral location, cm
Lateral location, cm

$u_g = 0.8 \text{ m/s}$

- Western
- Even
- Eastern

$q_{hy}/q_h$
Figure 4-4 Time-averaged lateral gas bubble distribution at the “injection level” under different gas inlet configuration

(a) $u_g = 0.4 \text{ m/s}$ (b) $u_g = 0.6 \text{ m/s}$ (c) $u_g = 0.8 \text{ m/s}$ (d) $u_g = 1.0 \text{ m/s}$

Figure 4-4 shows that even at the injection level, well above the distributor, the gas inlet condition due to the configuration of the inlet gas distributor strongly affect the lateral gas bubble distribution. This is observed at gas velocities ranging from 0.4 m/s to 1.0 m/s. To compare the lateral profiles of the bubble flux for those three gas distributors under different superficial gas velocities, it is easy to observe that both the Eastern and Western gas distributors give a higher peak bubble flux than that from the Even gas distributor. The peak bubble fluxes from all three distributors are off-centre. However, it is desirable to have more gas bubbles flowing to the first half of the jet cavity (18 cm - 29.5 cm), which can reduce the agglomerate according to the experimental results [16]. Therefore, two additional gas inlet distributor configurations are proposed, which are the Center-inlet-1 and Center-inlet-2, as shown in Figure 4-2. Figure 4-5 shows that by using the two new
inlet distributor configurations, the peak in gas bubble flux at the injection level can be moved from the left to the right of the column and any position in-between. Therefore, by modifying the inlet gas distributor configuration, the gas bubble flux peak at the injection level could be moved to a desired lateral position.

Figure 4-6 confirms that the influence of the gas distributor configurations on the bubble distribution persisted over the entire bed height based on the time-averaged gas volume fraction contours from the five different gas distributor configurations. Bubbles are concentrated on the west side for the case using the Western distributor (Figure 4-6 (a)), are relatively evenly distributed for the case using the Even distributor (Figure 4-6 (b)), and are concentrated in the central part for the case of the Eastern distributor (Figure 4-6 (c)). The difference between the two center inlet distributors (Figure 4-6 (d), (e)) is small in the lower part of the fluidized bed reactor. There are more bubbles located on the eastern side around the injection level (0.6 m - 0.8 m) for the Center-inlet-1 case than the Center-inlet-2 case, which has a more uniform gas bubble profile on the injection level, as shown in Figure 4-6 (d), (e).

Figure 4-7 shows the time-average velocity contours using the five different inlet gas distributors at a superficial gas velocity of 0.4 m/s. Compared to the case of the Even inlet gas distributor (Figure 4-7b), the uneven inlet gas distributors promote the formation of zones of high gas velocities above the distributor. Generally, these high-velocity zones become attenuated at higher locations, but with some inlet distributor, they propagate to relatively higher locations (Figure 4-7 e & f).

Figure 4-8 shows the averaged gas volume fraction contours using the Western inlet gas distributor configuration under different gas superficial velocities (0.4 m/s -1.0 m/s). Clearly, as the superficial gas velocity increases, more gas flows into through the fluidized bed, and the freeboard also increases. Moreover, the low gas voidage region exists for all four different superficial gas velocities, but gets smaller as the superficial gas velocity increases.
Figure 4-5 Gas bubble profiles at the injection level with different gas distributor configurations under the superficial gas velocity of 0.4 m/s
Figure 4-6 Time-averaged bubble volume fraction contour inside the fluidized bed reactor for different gas inlet configurations under the superficial gas velocity of 0.4 m/s
Figure 4-7 Time-averaged gas velocity contour inside the fluidized bed reactor for different gas inlet configurations under the superficial gas velocity of 0.4 m/s
Effect of the distributor slope

Figure 4-3 shows the three different gas distributors with an angle with a horizontal plane of 45°, 30°, or 0°. Figure 4-9 presents the influence of the inclined angle of the air distributor on the gas bubble flux profile on the injection level, with Even gas inlet distribution. With increasing the inclined angle, the peak value location moves from the center (25 cm) to the western side (around 14 cm). A possible reason for that is as inclined angle increases, increase tangential airflow through the distributor, which as a result, increases the bubbles transversal mixing and weakens the concentration of gas bubbles in the center of the reactor.

When combining the effects of the inlet gas distributor configuraton and the inclined distributor angle, Figure 4-9 and Figure 4-10 show that for the flat inlet and 30° inclined distributors, the effect of inlet gas distributor configuraton is much weaker when compared
to the 45° inclined distributor (Figure 4-4 (a)). If the objective is to adjust the gas bubble distribution at the injection level by modifying the inlet gas distributor configuration, a distributor with a strong incline is, therefore, preferable.

Figure 4-12 shows the time-averaged gas volume contours using three different inlet gas distributors with a 30-degree slope. It can be seen that the inlet gas distributor configuration affects the bubble distribution in the entire bed, but more on the bottom region of the bed. More gas bubbles are on the western side when using the Western gas distributor, and more gas bubbles are concentrated close to the eastern side when using the Eastern inlet gas distributor. In the case of the Even gas distributor, bubbles are evenly distributed in the entire fluidized bed. The contours for the 30° distributor are similar to the contours of the 45° distributor as shown in Figure 4-8 (a)-(c).

Figure 4-13 shows the time-averaged bubble volume fractions along the axial direction using three different inlet gas distributor configurations at 45° and 30° inclined angles. The bubble volume fraction in the Western gas distributor with 45° inclined angle is lower than other conditions. The lower gas volume fraction zone is in the bottom region, as shown in Figure 4-8. However, in the case of the 30° slope inlet, the lower gas volume fraction zone in the bottom area is not apparent, and the averaged gas hold up is similar from all three different inlet gas distributor configurations. The bubble volume fraction fluctuation in the area below 0.5 m is because of the multi-inlet distributor.
Figure 4-9 Comparison of the lateral gas bubble distributions at the injection level with different inlet gas distributor slope angles using the Even inlet gas distributor configuration under the superficial gas velocity of 0.4 m/s
Figure 4-10 Comparison of the lateral gas bubble distributions at the injection level using three different inlet gas distributor configurations under the superficial velocity of 0.4 m/s and the distributor angle of 30°
Figure 4-11 Comparison of the lateral gas bubble distributions at the injection level using three different inlet distributor configurations under the superficial velocity of 0.4 m/s and the flat inlet gas distributor.
Figure 4-12 Time-averaged bubble volume fraction contours using three different inlet gas distributor configurations with 30° distributor inclined angle under the superficial gas velocity of 0.4 m/s
Gas bubble distributions in a bubbling fluidized bed can be modified by changing the configuration of the gas distributor. The inlet gas distributor configuration can significantly affect the gas bubble lateral distribution well above the distributor level.

Figure 4-13 Time averaged bubble volume fractions using three different inlet gas distributor configurations at 45° and 30°inclined angles

4.4 Conclusion

Gas bubble distributions in a bubbling fluidized bed can be modified by changing the configuration of the gas distributor. The inlet gas distributor configuration can significantly affect the gas bubble lateral distribution well above the distributor level.

Gas bubble distributions in a bubbling fluidized bed can also be modified by changing the gas distributor angle. The effect of the inlet gas distributor is much more substantial for distributors with a large inclined angle. If the objective is to adjust the gas bubble
distribution at the injection level by modifying the inlet gas distributor configuration, a distributor with a large inclined angle is preferable.

4.5 References


Chapter 5

5 Effect of baffle and fluxtube on the hydrodynamics of a fluidized bed

5.1 Introduction

Fluidization is a process that allows solids to be handled like a fluid [1]. Gas-solid fluidized beds are widely applied in chemical and process industries for the synthesis of fuels and chemicals, and combustion and gasification of coal or biomass [2–4]. Fluidized beds are also used for drying, coating and granulation in pharmaceutical and food processes [2,5]. A fluidized bed is characterized by vigorous mixing of fluid and solids, uniform temperature, controllable handling of solids, and excellent mass and heat transfer. Nevertheless, the complexity of the gas-solid flow structure has been challenging to the researchers and led to numerous studies of its gas-solid hydrodynamics.

Baffles can be used to modify fluidized bed hydrodynamics. Baffles can provide staging for solids mixing [6] reduce solids bypassing [7,8] and the gas back mixing [6,9] in circulating fluidized beds. They can also improve the distribution of injected liquid on fluidized particles by redirecting gas bubbles to the spray jet cavity [10]. They significantly improve the performance of strippers [11–13]. They can reduce back mixing and improve the lateral mixing of particles in risers [14–18]. In risers, baffles can also help distribute injected feedstock and reduce gas bypassing [19,20]. They enhance gas-solid mass transfer by breaking gas bubbles. They can also be used to promote particle segregation [9,21–23].

A typical industrial application of baffles is in the Fluid Coking™ process, Fluid Coking™, and its variant, Flexicoking™ upgrade heavy crude oil or bitumen to lighter products. In the Fluid Coking™ process, bitumen is sprayed with atomization steam into a bubbling or turbulent fluidized bed with hot coke particles. Hot coke conveyed from a burner vessel provides the heat required for the thermal cracking and evaporation of hydrocarbons in the reactor vessel [24–26]. Imperfect dispersion of the sprayed liquid on the fluidized particles leads to the formation of wet agglomerates that decrease heat and mass transfer rates, thereby causing operating problems, especially in the stripper section where hydrocarbons vapors are removed from the coke particles just before they exit the reactor vessel [27–29].
A significant proportion of the injected liquid can be trapped in agglomerates that rapidly bypass the reaction section and contact the stripper sheds, causing their accelerated fouling [27–29]. This bypassing can be significantly reduced with using baffles [8,30]. Some studies also suggest that baffles may be used to modify the bubble distribution in the spray region to enhance liquid distribution on the particles, which will reduce agglomerate formation [10,31].

A proper design of the gas distributor can provide a uniform flow of small bubbles into the fluidized bed since the flow dynamics in the bottom zone of the fluidized bed have a significant influence on the bed performance, which we already discussed in Chapter-4 [32–36]. However, in the industrial systems, adding baffles into the fluidized bed would be more convenient and relatively low-cost, while providing the additional benefit of reducing solids bypassing [37].

Baffles can be used to redirect or break gas bubbles. Jahanmiri et al. [31] found that ring baffles with fluxtube change the distribution of bubbles over the cross-section and can be used to improve the liquid distribution, which can lower the formation of agglomerates significantly. Zhang et al. [9] investigated the effects of louver baffles on the hydrodynamics and gas-solid mixing characteristics of a fluidized bed reactor, which operated in both bubbling and turbulent regimes, using pressure fluctuations and steady gas tracer. In their study, a 2-D column structure with transparent plexiglass walls for visual observation was used. The study showed that louver baffles can break bubbles for superficial gas velocities < ∼0.7 m/s. A “gas cushion” phenomenon with a more dilute region appearing below the louver baffles and a denser region above the baffle was observed. The height of ‘gas cushion’ increased with the superficial gas velocity and reduction of solid backmixing was also observed. The conclusions were also validated by the numerical work from Yang et al. [38].

Baffles can be used to reduce solids bypassing and back mixing. Sanchez et al. [39] used radioactive particle tracking to study the effects of internal baffles in the stripping section of the Fluid Coking™, and discovered that using baffle increases the time that wet agglomerates spend above the baffle and reduces fouling on the sheds of the stripper
section [39]. Samruamphianskun et al. [40] did a comprehensive study of the effect of ring baffle configuration on the solid distribution in a circulating fluidized bed experimentally and numerically. Baffle opening area, space between baffle, baffle thickness, and the number of baffles were all compared in their study. The standard deviations of the solid volume fraction along the radial direction and the average of solid volume fraction along the height of the CFBR riser were chosen as response variables of the system mixing. They found out that the interaction between the baffle opening area and the space between the baffle affect the mixing quality the most. Baffles have also been applied to fluidized bed risers, where they have been shown to increase the gasoline yield by forcing catalyst particles from the wall region to the central region of the riser [13,14,41–44], and CFD has been successfully applied in the studies of the effect of baffles on the hydrodynamics of the bed.

5.2 Problem statement

The objective of this study is to study the effect of a baffle on the bubble flow patterns in a 2D fluidized bed. It will focus on the following aspects:

1. Effect of the baffle, with and without a flux tube, on the gas bubble flow patterns.
2. Effect of the bubble flow patterns below the baffle on the redirection of the bubbles achieved with the baffle, using different inlet gas conditions.
3. Effect of baffle geometry and location on bubble distributions.

5.3 Configuration of the Fluidized Bed and Numerical model

The numerical simulations of a lab-scale bubbling fluidized bed reactor with a rectangular cross-section, which were provided by Li et al. [10], were carryout and the experimental data from Li et al. [10] were used to validate the numerical model.

The experimental setup is shown in Figure 5-1, which consisted of a fluidized bed section and an expansion section. The bed thickness was 0.1 m in both sections. The total height of the unit was 2.18 m, and the unit width expanded from 0.5 m in the lower bed section to 1.0 m in the expansion section. Initially, the bed was filled with about 100 kg of silica sand.
with a Sauter-mean diameter of 190 μm. Air at ambient conditions with relative humidity below 12% was used as the fluidizing agent. The gas distributor consisted of two rows of 10 tuyeres, for a total of 20 tuyeres, distributed on an angled slope, as shown in Figure 5-1. Each tuyere was supplied by a dedicated sonic orifice to maintain the required gas mass flow rate, which, in this study, was the same for each active tuyere. The minimum fluidization velocity was 0.033 m/s, and the minimum turbulent velocity is about 0.4 m/s at 30°C [10].

Three different gas distributor configurations are shown in Figure 5-2. The base case was the Even distributor in which the ten active gas tuyeres were evenly distributed. The second gas distribution was the Western distributor with ten active tuyeres near the western side of the column, and the third gas distribution was the Eastern distributor with ten active tuyeres near the eastern side of the column.

The two-dimensional numerical model for the simulation of the lab-scale fluidized bed reactor with different gas inlet condition is shown in Figure 5-3. The Euler-Euler approach was applied with the Reynolds-averaged Navier-Stokes formulations to simulate the flow in a gas-solid fluidized bed. Gas and solid phases could be present simultaneously in the same computational volume by introducing the volume fraction for each phase. The detailed information on the numerical model was given in Chapter 2.

Simulations for twelve different cases were carried out to validate the numerical model. The predicted gas bubble profiles at the injection level with three different gas distributors, i.e. three different inlet boundary conditions for the gas phase (Western case, Even case, and Eastern case) at four different superficial gas velocities (0.4 m/s, 0.6 m/s, 0.8 m/s, 1.0 m/s) were compared with the experimental results to validate the numerical model.

Different types of baffles were used in the bed, as shown in Figure 5-4, Figure 5-5, and Figure 5-6. The lower tip of each baffle was always at the height of 0.915 m from the bottom of the bed. The asymmetrical baffle was an open-ended right triangle shape baffle with an internal angle of 45°, which spanned the entire depth of the bed, from wall to wall. The dimensions of the asymmetrical baffle were 0.18 m × 0.18 m × 0.10 m on the western
side. The symmetrical baffles were at the same height on both sides of the bed with a relatively smaller size, which is 0.125 m × 0.125 m × 0.10 m.

A schematic of the baffled fluidized bed reactor (BFB) is shown in Figure 5-4, Figure 5-5, Figure 5-6, and Figure 5-7. The two-dimensional numerical model of the baffled fluidized bed is shown in Figure 5-8, Figure 5-9 and Figure 5-10.

Figure 5-1 Schematic diagram of the lab-scale bubbling fluidized bed
Figure 5-2 Schematic diagram of different gas distributor configurations
Figure 5-3 Different gas inlet conditions used in the two-dimensional simulations
Figure 5-4 Location and dimensions of the baffles without fluxtubes that were tested in this study
Figure 5-5 Location and dimensions of the baffles with fluxtubes of different lengths that were tested in this study
Figure 5-6 Location and dimensions of the baffles with fluxtubes of different width that were tested in this study
Figure 5-7 Location and dimensions of the symmetrical baffle and fluxtube that were tested in this study
Figure 5-8 Schematic of the simulated 2-D baffled fluidized bed (a) No baffle, (b) Asymmetrical baffle, and (c) Symmetrical baffles
Figure 5-9 Schematic of the simulated 2-D fluidized bed with fluxtube (a) Long fluxtube, (b) Regular fluxtube, and (c) Short fluxtube
Figure 5-10 Schematic of the simulated 2-D fluidized bed with fluxtube (a) Regular fluxtube, (b) Fluxtube baffle-D2, and (c) Fluxtube baffle-D3
5.4 Results and discussion

5.4.1 Effect of the fluxtube on the gas bubble distribution

Figure 5-11 Time-averaged bubble contours of the baffle zone (0.40 m < y < 0.70 m) with and without baffle and regular fluxtube under the superficial gas velocity of 0.4 m/s) using the Even gas inlet distributor configuration
a. Without baffle

b. Asymmetrical baffle
c. Symmetrical baffle
d. Fluxtube baffle

Figure 5-12 Time-averaged gas velocity magnitude contours with velocity vectors at the baffle zone (0.40 m < y < 0.70 m) with and without baffle and regular fluxtube under the superficial gas velocity of 0.4 m/s using the Even inlet gas distributor

This section presents results on the effect of baffles and fluxtube on hydrodynamics. Figure 5-11 shows the time-averaged gas holdup contour of the baffle zone (0.40 m < y < 0.70 m) with and without baffle and regular fluxtube with a superficial gas velocity 0.4 m/s, near the transition from the bubbling to turbulent fluidization regimes. A “gas pocket” with a high gas concentration appears each the baffle: the asymmetrical baffle, the symmetrical baffle and the baffle with regular fluxtube. Simultaneously, a denser, lower voidage region appears on the top of all three baffles. The gas pocket under the baffle and the denser region on top of the baffle are greatly reduced with the flux tube (Figure 5-11 d).
Figure 5-12 shows the gas velocity magnitude contour with the vector direction indicated with arrows. Without a baffle, the bed can be separated into two zones: in the central core region, the gas moves quickly upward, and in the annulus region, near the wall, the gas moves slowly, and upward or downward. With baffles, a clear internal circulation pattern appears above the asymmetrical baffle without and with fluxtube (Figure 5-12b, d). In the fluxtube velocity contour (Figure 5-12d), gas bubbles go up through the tube, but an animation of the simulation results (see Appendix) showed that a small proportion of the bubbles go down through the tube near the western side. With the symmetrical baffle (Figure 5-12c), an internal circulation pattern appears above each baffle half: the circulation patterns are not precisely symmetrical because of the gas flow from the sloped distributor is not symmetrical.

Figure 5-13 shows the time-averaged gas volume contours of the whole fluidized bed with and without baffle. It shows that all the tested baffles, with or without fluxtube, can redistribute the gas flow in the whole bed. Compared to the no baffle fluidized bed, baffles increase the volume fraction occupied by gas bubbles (Figure 5-13) and intensify the turbulence (Figure 5-12), and these changes are observed both below and above the baffle level. The symmetrical baffle evens up the gas bubble distribution across the cross-section of the bed.

Figure 5-14 compares the lateral profiles of the gas bubble flux at the injection level for the cases with and without baffles and fluxtube. The symmetrical baffle modifies the asymmetrical bubble distribution observed without baffle so that the resulting profile is nearly perfectly symmetrical. The asymmetrical baffle, with and without fluxtube, moves the bubble flow peak from the right-hand side to the left-hand side.
Figure 5-13 Time-averaged bubble volume fraction contour inside the fluidized bed reactor with and without baffle and regular fluxtube under the superficial gas velocity of 0.4 m/s with even case inlet configuration

Lateral location, m

a. Without baffle  b. Asymmetrical baffle  c. Symmetrical baffle  d. Regular fluxtube
Figure 5-14 Lateral gas bubble distribution profile at injection level with and without baffle and fluxtube under the superficial gas velocity of 0.4 m/s with even inlet configuration.
5.4.2 Influence of fluxtube length with the Even case

Figure 5-15 Time-averaged bubble volume fraction contour of the baffle zone ($0.40 \, m < y < 0.70 \, m$) without and with fluxtube baffle of different lengths under the superficial gas velocity of 0.4 \, m/s with even gas inlet configuration
Fluxtubes, which Wyatt et al. [30] patented, are used on industrial ring baffles to help reduce fouling of the stripping section of the Fluid Cokers. This section and the next one will explore the fluxtube configuration parameters. This section focuses on the impact of the fluxtube length.

A “gas pocket” appears under the baffle with the three different fluxtube sizes (inside length: 18 cm, 16 cm, 12 cm) and is always much smaller than observed in the absence of a fluxtube, as shown in Figure 5-15. The volume of the “gas pocket” near the baffle tip increases with increasing fluxtube length. In all cases, the fluxtube reduces the size of the low gas holdup and high solid holdup region on the top of the baffle, especially with the shortest fluxtube: this is likely due to the agitation provided by the bubbles rising through

Figure 5-16 Time-averaged bubble volume fraction contour inside the fluidized bed reactor without and with fluxtube baffle of different lengths under the superficial gas velocity of 0.4 m/s with even gas inlet configuration

Fluxtubes, which Wyatt et al. [30] patented, are used on industrial ring baffles to help reduce fouling of the stripping section of the Fluid Cokers. This section and the next one will explore the fluxtube configuration parameters. This section focuses on the impact of the fluxtube length.

A “gas pocket” appears under the baffle with the three different fluxtube sizes (inside length: 18 cm, 16 cm, 12 cm) and is always much smaller than observed in the absence of a fluxtube, as shown in Figure 5-15. The volume of the “gas pocket” near the baffle tip increases with increasing fluxtube length. In all cases, the fluxtube reduces the size of the low gas holdup and high solid holdup region on the top of the baffle, especially with the shortest fluxtube: this is likely due to the agitation provided by the bubbles rising through
the fluxtube. Figure 5-16 show that more gas bubbles are going through the short fluxtube than through the other fluxtubes, as the gas holdup in the tube is larger. Figure 5-16 also shows that the gas is more evenly distributed through the whole bed volume with the shortest flux tube. According to Figure 5-17, the regular flux tube does not greatly change the impact of the baffle on the gas bubble flux. On the other hand, both the long and short flux tubes enhance the shifting of the bubble gas flux towards the fall opposite the baffle.

![Radial gas bubble distribution profile](image)

**Figure 5-17** Radial gas bubble distribution profile at injection level without and with fluxtube baffle of different lengths under the superficial gas velocity of 0.4 m/s with even gas inlet configuration
5.4.3 Influence of fluxtube width

Different fluxtube diameters, with an inside width of 8 cm, 6 cm, or 2 cm, were compared. Figure 5-18 shows that both the “gas pocket” below the baffle and the dense region above the baffle appeared with all fluxtube diameters. Figure 5-18 and Figure 5-19 show that the gas holdup within the tube is more massive for the smallest tube diameter. According to Figure 5-20, the different diameters of fluxtube do not influence the gas bubble profile at the injection level.

Figure 5-18 Time-averaged bubble volume fraction contour of the baffle zone (0.40 m < y < 0.70 m) without and with fluxtube baffle of different widths under the superficial gas velocity of 0.4 m/s with even gas inlet configuration
Figure 5-19 Time-averaged bubble volume fraction contour inside the fluidized bed reactor without and with fluxtube baffle of different lengths under the superficial gas velocity of 0.4 m/s with even gas inlet configuration.
Figure 5-20 Radial gas bubble distribution profile at injection level without and with fluxtube baffle of different widths under the superficial gas velocity of 0.4 m/s with the Even inlet gas distributor configuration.
5.4.4 Effect of symmetrical fluxtube

Figure 5-21 Time-averaged bubble volume fraction contour of the baffle zone (0.40 m < y < 0.70 m) with symmetrical baffle and fluxtube under the superficial gas velocity of 0.4 m/s with even gas inlet configuration

The effects of the symmetrical baffle, with and without fluxtubes, are compared in this section. Figure 5-21 shows that, as with the asymmetrical baffle (Figure 5-15), the fluxtube reduces the size of both the “gas pocket” below the baffle and of the dense region above the baffle. According to Figure 5-22, the gas pocket under the baffle, with and without fluxtube, is larger on the western side than on the eastern side: this is likely caused by the sloped gas inlet distributor that results in more gas bubbles on the western side. The symmetrical baffle, with and without fluxtubes, redistributed the gas bubbles to the central region and increased the gas bubble rise velocity, as shown in Figure 5-12. The relatively high gas velocity in the central region may be the result of a smaller opening area, between the baffle tips, and stronger recirculation liquid flow patterns above the baffle.

Figure 5-23 indicates that, although the results are similar for the symmetrical baffle with and without flux tube, the presence of the fluxtube increases the gas holdup throughout the bed. Figure 5-24 shows that, with the symmetrical baffle, the flux tube does not significantly affect the gas bubble distribution above the baffle.
a. Symmetrical baffle

b. Symmetrical fluxtube

Figure 5-22 Time-averaged gas velocity magnitude contour with velocity vectors with directions of the baffle zone (0.40 m < y < 0.70 m) with symmetrical baffle and fluxtube under the superficial gas velocity of 0.4 m/s with even gas inlet configuration.
Figure 5-23 Time-averaged bubble volume fraction contour inside the fluidized bed reactor with symmetrical baffle and fluxtube under the superficial gas velocity of 0.4 m/s with even gas inlet configuration
Figure 5-24 Radial gas bubble distribution profile at injection level with symmetrical baffle and fluxtube under the superficial gas velocity of 0.4 m/s with even gas inlet configuration
5.4.5 Effect of fluxtube for different inlet gas distributor configurations

Figure 5-25 Time-averaged bubble volume fraction contour of the baffle zone (0.40 m < y < 0.70 m) without and with fluxtube baffle for different gas inlet configurations under the superficial gas velocity of 0.4 m/s
b. Western  
b. Even  
c. Eastern  
d. Western (baffle)  
e. Even (baffle)  
f. Eastern (baffle)
According to Figure 5-25 a, b, c, the effect of three inlet configurations on the gas holdup in the baffle zone is not very significant, although a gradual decrease in gas holdup from the western side to the eastern side appears for the Eastern distributor. Figure 5-25 d, e, f shows that, with the three gas distributors, the baffle concentrates the gas bubbles to the central region, and in all cases, there is a “gas pocket” below the baffle and a dense region above the baffle.

With all gas distributors, the flux tube reduces the size of the “gas pocket” below the baffle and of the dense region above the baffle (Figure 5-25 g, h, i). The fluxtube also reduces the gas concentrate in the baffle zone when compared to the baffle without fluxtube (Figure 5-25).

Figure 5-26 Time-averaged bubble volume fraction contour inside the fluidized bed reactor without and with baffle and fluxtube for different gas inlet configurations under the superficial gas velocity of 0.4 m/s

g. Western (fluxtube) h. Even (fluxtube) i. Eastern
Figure 5-26 a, b, c, study the effect of different inlet gas distributor configurations on gas holdup distribution throughout the fluidized bed: with the western inlet configuration, a region with a low gas holdup is found near the bottom of the bed (Figure 5-26a). Figure 5-26d and g show that this region is reduced by the liquid recirculation currents induced by the baffle and, significantly, by the baffle with flux tube. Figure 5-26 confirms that the gas holdup is greatly increased by the baffle for all distributors, with a smaller increase in gas holdup observed with the flux tube.

5.5 Conclusion

In this chapter, a multi-phase Eulerian-Eulerian two-fluid method (TFM) coupled with the kinetic theory of granular flow (KTGF) was used to study the impact of baffles and flux tube on the hydrodynamic of the gas-solid fluidized bed. The study found that:

Baffles can redirect gas bubbles. Baffles, with and without flux tube, can modify the uneven radial distribution of gas bubbles in the baffle zone. In all configurations, a gas pocket appeared under the baffle, and a denser region appeared above the baffle. Baffles increased the gas holdup throughout the bed, with the increase being moderated by adding a flux tube to the baffle. Baffles also induce strong liquid recirculation currents.

The length of the flux tube has a more considerable impact on the column hydrodynamics than its diameter. The length of the flux tube has more effect on the gas bubble distribution on the injection level. The redirection of the gas bubbles above the baffle can be adjusted by modifying the length of the flux tube.

5.6 References


[31] M. Jahanmiri, Use of a baffle to enhance the distribution of a liquid sprayed into a gas-solid fluidized bed, Western university, 2017.


Chapter 6

6 Particle tracking in a fluidized bed

6.1 Introduction

Fluidized beds have many industrial applications because of their inherent characteristics, such as good solids mixing, easy temperature control, and adaptability to high-pressure and high-temperature operations [1]. Fluidized beds are ideal for chemical processes such as olefin polymerization or acrylonitrile synthesis, where it is essential to maintain a constant and uniform temperature, which is made possible by the good solids mixing [2]. In processes such as combustors, gasifiers or pyrolyzers, the reacting feed particles have a different density than the bed particles, and intense solids back mixing is required to prevent or at least mitigate segregation of the reacting particles [3–6]. Good solids mixing makes the fluidized bed attractive for granulation, where it is essential to move particles rapidly in and out of the region where the liquid binder is sprayed [2]. However, the solid bypassing reduces the gas-solid contract efficiency in the bubbling fluidized bed [5–7].

In polyolefin fluidized bed reactors, the reactor operates close to the polymer melting point, and high-quality solids mixing is critical to avoid the formation of hot spots. However, it is also vital to allow enough time for catalyst or prepolymer particles that are continuously injected into the bed to get a chance to grow before they are withdrawn from the bed.

In the Fluid Coking™ process, heavy oil is injected into a fluidized bed of hot coke particles that provide enough heat for the heavy liquid to crack and vaporize [8,9]. Hot coke particles are conveyed to the top of the Coker, flow down through the Coker, a stripper section and are conveyed from the reactor bottom to a burner. As in granulation processes, intense solids mixing helps evacuate wet solids quickly from spray regions, minimizing the formation of wet agglomerates [10]. In wet agglomerates, the conversion of liquid trapped in the core of the agglomerates is slowed down by low heat transfer from the hot bed through the outer layers of the agglomerate [11–13]. Agglomerates exiting the reactor with unconverted liquid are responsible for the fouling of the stripper sheds and the loss of valuable hydrocarbons to the burner [11,12]. Rapid bypassing of wet agglomerates from the spray regions to the bottom of the reactor should, therefore, be minimized [11,14].
Intense solids bypassing from the spray regions to the bottom of the reactor could even lead to wetted individual particles leaving the reactor [15,16].

In fluidized beds, solids mixing is induced by the motion of the gas bubbles, which carry solids in their wake [17,18]. Solids mixing can, therefore, be altered by modifying the bubbles flow patterns. There are two main approaches to modify the bubble movement: changing the gas distributor design and adding a baffle.

The first modifies the gas distributor. Li et al. [19] show that the solid flow pattern in a fluidized bed can be modified by changing the gas distributor and the superficial gas. The mixing quality increased with the number of symmetrical vortexes in the fluidized bed. The solid vertical dispersion coefficient was 3–5 times larger than the horizontal dispersion coefficient [20].

Secondly, most internals can suppress axial solids back mixing. Inserting a horizontal baffle is a standard and practical method to limit solid back mixing [21]. Baffles can provide staging for solids mixing [22], reducing solids bypassing in circulating fluidized beds [23,24]. Ring baffles can reduce back mixing and improve the lateral mixing of particles in risers [25–29]. Experimental results show that the solid backing mixing has a close relationship with gas velocity, baffle free area, and baffle spacing.

Solid dispersion efficient and mixing index are the common parameters for quantifying the solid mixing rate in the fluidized bed. Moreover, usually, researchers consider the solid mixing separately in the axial and lateral direction. In a fluidized bed, the solid mixing in the vertical direction mainly occurs when bubble drag surrounding solids into their wake regions and the horizontal mixing is due to the bubble motion as bubbles coalesce and when bubbles erupt at the bed surface [17,21].

The ideal method to determine the solid dispersion efficient is to track the movement of the tagged particles inside a fluidized bed reactor, where all the tagged particles are released from the same starting position at the same time [30]. Li et al. [30] applied the Positron emission particle tracking (PEPT) tracking system to track the tracer particle trajectory in the fluidized bed reactor. PEPT is a tracking method using radioactive $^{18}$F to label tracers.
The two-fluid model (TFM) is based on the kinetic theory of granular flow (KTGF) as one of the models that applied to predict the behavior of fluidized bed. In the TFM model, both gas and solid phases are considered as interpenetrating continua, which means volume-averaged Navier-Stokes equations solve both the gas phase and solid phase. The TFM model has been validated and widely applied by researchers to simulate the gas-solid fluidized bed reactor [31–35]. However, because of the lack of particle motion, it is not suitable for tracking individual particles. Several methods were applied in the TFM model to track discrete tracer particles [36–38]. Banaei et al. [38] used a newly defined parameter CFP to measure the solid mixing properties. The tracking procedure is based on the solids flow rate through each of the eight-cell faces and the probability of the tracer particles leaving or staying in the cell. The technique is not perfect. However, it opens the door to particle tracking from simulations based on the TFM approach.

6.2 Problem statement

The objective of this study is to use CFD simulations to predict the impact of changes in gas distributors and baffles on solids mixing and bypassing. In particular, this chapter will explore the following topics:

1. Effect of superficial gas velocity on solids mixing and bypassing;
2. Effect of different inlet gas distributor configuration;
3. Effect of baffle geometry and location.

6.3 Methodology and Experimental Setup

In this section, the gas-solid two-phase model introduced in Chapter 2 is used to simulate the fluidization. Furthermore, solid mixing is characterized by a solid dispersion coefficient at different locations [31].

6.3.1 Tracking of particles

In this chapter, we consider two aspects of particle mixing properties. The first aspect is how quickly particles can reach the bottom of the reactor. The second aspect is how quickly particles. Flowchart for particle tracking is shown in
For \( p = 1: N \) (starting location point) \( N \) is the number of location \((x_p, y_p)\)

**Read the location value** \( x, y \)

- For \( i = 1: \) number (data files)
  
  \[
  \text{point}_a = \text{results}_\text{data}\{i+3\}; \text{read data from result (time-}i, x, y, u_x, u_y) \]

- Finding the first point in \( \text{point}_a \) by searching for the closest point \( x_1, y_1 \)

  \[
  \begin{align*}
  xx &= x_1 + u_x \times \Delta t \\
  yy &= y_1 + u_y \times \Delta t 
  \end{align*}
  \]

  if \( xx \leq 0 \), the point will hit the wall and stop, \( \Delta t = 0.1s \);  
  if \( yy \leq 0 \), the point already goes to the stripper zone, \( \Delta t = 0.1s \);

- if \( (xx > 0 \) and \( yy > 0) \) continue to the next time file to looking for the data

- For \( j = i+1: \) number
  
  \[
  \text{point}_b = \text{results}_\text{data}\{j+3\}; \]

  **Finding the second point in** \( \text{point}_b\{j\} \) **by searching for the closest point** \( xx, yy \) in \( \text{point}_b\{j\} \)

  Second point \((x_2, y_2, u_{x2}, u_{y2})\)

  if \( y_2 \leq 0 \)

  \( \Delta t = t_j - t_i \) record \( \Delta t \) and move to \( i+1 \) and continue

  break

  else

  \[
  \begin{align*}
  xx &= x_2 + u_{x2} \times \Delta t \\
  yy &= y_2 + u_{y2} \times \Delta t 
  \end{align*}
  \]

  Continue to look for the next point in \( \text{point}_b\{j+1\} \) until the \( y < = 0 \) record \( \Delta t \) End

End Finish all the location point \( p = 1: N \) (location number) and create the contour of the det-time quantiles values at different locations.

**Figure 6-1 Flowchart for particle tracking procedure**
The primary numerical tracking model in this chapter is based on the TFM numerical results. The vertical and horizontal components of the gas and particle at 10 s (steady-state) were used as the initial condition in this study. The flowchart for tracking particles is shown in Figure 6-1. The interval time is recorded for the time distribution contour map.

6.3.2 Formation to stripper time distribution contour

This method was select to track statistical distributions of the time for the particles in the formation zone to go to the stripper zone. As shown in Figure 6-3, points are selected in the formation zone: 3363 points for the no baffle fluidized bed reactor and 7014 points for the baffled fluidized bed. Particles at each location were tracked to see how long it took to reach the stripper zone. Once the particle is in the stripper zone, a new tracking will begin at the same location with particle information at the next 0.1s. The cumulative time distribution was collected for each location. The cumulative formation-to-stripper time distributions were a very vital factor in industrial applications. Typically, the longer it takes a particle or agglomerate to go from the formation zone to the stripper zone, the longer it will have to react, and the less liquid it will carry. Therefore, it is essential to minimize the percent of particles with short formation to stripper times.

Because of the large number of locations considered in this study, it is difficult to compare all their cumulative time distributions and specific times are reported instead: \( t_{10} \) is the time at which 10 % of particles have reached the stripper, \( t_{25} \) corresponds to 25% of the particles, and \( t_{50} \) is the median of the formation to stripper time.

This chapter will use dimensionless time distributions, which are the time \( t_i \) divided by the meantime \( t_{average} \).

\[
\tau_i = t_i / t_{average} \tag{1}
\]

\( i = 10, 25, 50 \)
Figure 6-2 Schematic diagram of the lab-scale bubbling fluidized bed with formation zone and stripper zone
Figure 6-3 Schematic diagram of initial points selected in the formation zone
Figure 6-4 An example of sampled particles released from same location
6.3.3 Dispersion of particles

To evaluate the solid mixing quality in the fluidized bed reactor, it is important to quantify how quickly and how well the particles disperse throughout the bed. An ideal method to measure the dispersion of particles is to track particles. The 2-D starting location (0.25,0.306) was chosen as an example shown in Figure 6-4. Every 0.1 s, one particle is released from the same location, and each particle will be tracked for 10 s. One hundred particles were released over a 10 s period. The trajectory of each particle was recorded, and different aspects of how well and quickly particles disperse in the fluidized bed are calculated and compared under different operating conditions.

6.3.3.1 Solids axial and lateral dispersion properties

To measure the solid mixing rate in the fluidized bed reactor, researchers have been used axial and lateral dispersion coefficient or mixing index to represent the mixing rate inside the fluidized bed [38–41]. Based on the tracking particle method mentioned in 6.3.1, the formation zone was divided into 550 different locations. At each location, a particle was released every 0.1 s and tracked for 10 s. To evaluate the axial dispersion, the standard deviation of the 100 final y coordinates was calculated, and, for the lateral dispersion, the standard deviation of the final x coordinates was calculated.

6.3.3.2 Dispersion of final location of each particle

In the fluidized bed formation zone, because of the existence of baffle, it can be separated into three sections. The sections are shown in Figure 6-5, and their heights are reported in Table 6-1: the heights are the same for the top and bottom sections. For 100 particles released from each location simultaneously, the final location of each particle after 10 s track was recorded. The final distance between all possible binary combinations of the 100 particles is calculated, and the standard deviation and averaged distance value of all the distances are recorded for each location. Figure 6-5 shows the location of each section of the formation zone in the fluidized bed reactor.
The final location of the 100 particles, all the possible Euclidean distances are calculated for each pair of particles as shown in Equation (2), and the average and standard deviation of all the distances are calculated by Equation (3) and Equation (4).

Apart from that, the other method is to compare the value of averaged distance and standard deviation of distance with random locations inside the fluidized bed geometry. For the random locations, to validate the independent of the averaged distance value and standard deviation of the random picked location, different number of locations are picked, which include 100, 1000, 2000, 5000, and 10000. For each number level, 10000 times of picking process are running and averaged distance and standard deviation value for each running are recorded.

For each number level, the averaged value \(d_{\text{random}}\) of 10000 averaged distance values, and standard deviation \(R_{d_{\text{random}}}\) of 10000 averaged distance values are compared in Table 6-2. From Table 6-2, it is easy to obvious that for the averaged distance value, it does not change a lot with more locations picked. To avoid the effect of random picked location on the comparison, the value from the number level of 10000 will be chosen to do the calculation. Similarly, the standard deviation calculation results are also shown in Table 6-2. The dispersion index by averaged distance \(R_{(a,b)}\) and standard deviation \(R_{stv(a,b)}\) are shown in Equation (5) and Equation (6). Both will be applied to qualify and quantify the particle dispersion condition.

For each location \((a, b)\) the 100 released particles

\[
d_k = \sqrt{(x_i - x_j)^2 + (y_i - y_j)^2}
\]  
(2)

\((x_i, y_i)\) and \((x_j, y_j)\) are the final location of the 100 particles

\[1 \leq i \leq 100, 1 \leq j \leq 100\]

\[1 \leq k \leq 5050, \text{ which include the repeat calculation of distance}\]
\[
d_{\text{average}} = \frac{1}{n} \sum_{k=1}^{n} d_k, \quad n = 5050
\]  

(3)

\[
d_{\text{stdv}} = \sqrt{\frac{\Sigma (d_k - d_{\text{average}})^2}{n}}
\]

(4)

\[
R_{(a,b)} = \frac{(d_k)_{(a,b)}/(d_{k,\text{random})_{a,b}}}{(d_k)_{(a,b)}/(d_{k,\text{random})_{a,b}}}
\]

(5)

\[
R_{stv(a,b)} = \frac{(S_k)_{(a,b)}/(S_{k,\text{random})_{a,b}}}{(S_k)_{(a,b)}/(S_{k,\text{random})_{a,b}}}
\]

(6)

**Table 6-1 The dimension of different section**

<table>
<thead>
<tr>
<th></th>
<th>Top section</th>
<th>Baffle section</th>
<th>Bottom section</th>
</tr>
</thead>
<tbody>
<tr>
<td>Location-y</td>
<td>0.0-0.4 m</td>
<td>0.4 -0.7 m</td>
<td>0.7-1.1 m</td>
</tr>
<tr>
<td>Height</td>
<td>0.4 m</td>
<td>0.3 m</td>
<td>0.4 m</td>
</tr>
<tr>
<td></td>
<td>100</td>
<td>1000</td>
<td>2000</td>
</tr>
<tr>
<td>------------------------------</td>
<td>-----------</td>
<td>-----------</td>
<td>-----------</td>
</tr>
<tr>
<td>Average value of the distance ( \overline{d_{k,random}}_{a,b} ) ( \text{m} )</td>
<td>0.748668</td>
<td>0.748546</td>
<td>0.748485</td>
</tr>
<tr>
<td>Standard deviation of averaged distance ( S_{(d_{k,random})_{a,b}} )</td>
<td>0.009963</td>
<td>0.006969</td>
<td>0.00576</td>
</tr>
<tr>
<td>Average value of the standard deviation, ( \overline{S_{k,random}}_{a,b} )</td>
<td>0.458022</td>
<td>0.459418</td>
<td>0.459454</td>
</tr>
<tr>
<td>Standard deviation of averaged standard deviation ( S_{(\overline{S_{k,random}})_{a,b}} )</td>
<td>0.022041</td>
<td>0.006854</td>
<td>0.004772</td>
</tr>
</tbody>
</table>
Figure 6-5 Schematic diagram of the lab-scale bubbling fluidized bed with different section
6.4 Results and discussion

6.4.1 Particle time distribution

6.4.1.1 Particle time distribution under different inlet gas distributor configurations

This section used the method described section 6.3.2 to compare the formation-to-stripper time distribution map for different configurations. We have five different inlet gas distributor configurations, as shown in Chapter-3.

Figure 6-6 (a) compared the $\tau_{10}$ distribution map for five different inlet configurations, where $\tau_{10}$ is the dimensionless time at which 10% of particles released at a given formation location have reached the stripper. For all the configurations, $\tau_{10}$ becomes smaller as the initial formation location becomes closer to the stripper zone, as expected. For initial formation locations above 0.8 m, the Even distributor configuration is preferable as it does have the larger $\tau_{10}$ associated with the other inlet configurations. On the other hand, for initial formation locations below 0.4 m, the Even distributor configuration seems the least attractive as, with all the other inlet configurations, small values of $\tau_{10}$ could be avoided by restriction formation to the central region.

Figure 6-6 compares the dimensionless times $\tau_{10}$, $\tau_{25}$ and $\tau_{50}$ for the different inlet gas distributor configurations. It confirms that the Even configuration is best for initial locations about 0.8 m. The two center-inlet configurations are best for the initial locations below 0.4 m, if the formation is located in the central region.

As the time distribution contours for $\tau_{10}$, $\tau_{25}$ and $\tau_{50}$ are quite similar to each other, the following discussion will focus on the time distribution contour for $\tau_{10}$. The dimensionless time $\tau_{10}$ is especially crucial as it characterizes the fastest 10% particles, most likely to contain unconverted liquid.
(a) $\tau_{10}$

(b) $\tau_{25}$
6.4.1.2 Effect of superficial gas velocity on the time distribution

Figure 6-7 confirms that the best inlet gas distributor configurations identified for a superficial gas velocity of 0.4 m/s (Figure 6-6) remain the best configurations for gas velocities ranging from 0.4 to 1 m/s. For formation heights above 0.8 m, the Even configuration is best while the Eastern and Western configurations are best for formation heights below 0.4 m, especially if the formation can be confined to the central region.

There is no clear trend on the effect of superficial gas velocity $\tau_{10}$. The minimum turbulent velocity is 0.6 m/s, which means the fluidization transitions to the turbulent regime around 0.6 m/s. This transition may explain why the differences between the 0.4 m/s, 0.6 m/s, 0.8 m/s and 1.0 m/s are not noticeable, since the structure of the bubble wakes, which are
responsible for solids mixing, changes when the regime transitions from bubbling to turbulent [42].

(a) Western inlet configuration
(b) Even inlet configuration

(c) Eastern inlet configuration

Figure 6-7 $\tau_{10}$ time distribution map for different gas inlet configurations under different superficial gas velocities
6.4.1.3 Impact of baffle and baffle modification

The purpose of adding a baffle to fluidized beds, as in the Fluid Coking process, is to eliminate short residence times between formation zones above the baffle to the stripper, which is located well below the baffle. This section, thus, investigates how a baffle can increase $\tau_{10}$.

Figure 6-8 $\tau_{10}$ time distribution map with asymmetrical baffle and fluxtube, with the even gas inlet configuration (Western side is left-hand side) under superficial gas velocity of 0.4 m/s.

The purpose of adding a baffle to fluidized beds, as in the Fluid Coking process, is to eliminate short residence times between formation zones above the baffle to the stripper, which is located well below the baffle. This section, thus, investigates how a baffle can increase $\tau_{10}$.

Figure 6-8 shows the effect of the asymmetrical baffle, with and without fluxtube, on the time distribution from formation zone to stripper zone. With the Even inlet gas distributor configuration, both asymmetrical baffles and fluxtube can improve the solids time distribution on the western side (left-hand side on Figure 6-8), but worsens it on the Eastern
Adding a flux tube to the asymmetrical baffles appears to enhance its effects on solids' time distribution.

Figure 6-9 shows that the symmetrical baffle, with and without fluxtube, does not improve the solids time distribution. The rest of this study will, therefore, focus on the asymmetrical baffle.

![Figure 6-9: Time distribution maps with symmetrical baffles and asymmetrical fluxtube.](image)

<table>
<thead>
<tr>
<th>Height, m</th>
<th>Lateral location, m</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.1</td>
<td>0.1</td>
</tr>
<tr>
<td>0.2</td>
<td>0.2</td>
</tr>
<tr>
<td>0.3</td>
<td>0.3</td>
</tr>
<tr>
<td>0.4</td>
<td>0.4</td>
</tr>
<tr>
<td>0.5</td>
<td>0.5</td>
</tr>
<tr>
<td>0.6</td>
<td>0.6</td>
</tr>
<tr>
<td>0.7</td>
<td>0.7</td>
</tr>
<tr>
<td>0.8</td>
<td>0.8</td>
</tr>
<tr>
<td>0.9</td>
<td>0.9</td>
</tr>
<tr>
<td>1.0</td>
<td>1.0</td>
</tr>
<tr>
<td>1.1</td>
<td>1.1</td>
</tr>
</tbody>
</table>

Figure 6-9 $\tau_{10}$ time distribution map with symmetrical baffles and symmetrical fluxtube, with the even gas inlet configuration (Western side is left-hand side) under superficial gas velocity of 0.4 m/s

6.4.1.4 Fluxtube modification - Effect of fluxtube length

This section compares three different fluxtube lengths with the asymmetrical baffle. Figure 6-10 shows similar time distribution maps with the three different fluxtube lengths. The strongest impact of the fluxtube length is on the solids residence time distribution of particles formed below the baffle: the short fluxtube reduces the most attractive area with
\( \tau_{10} \) larger than 1 and move it towards the center. For the particles formed above the baffle, the long fluxtube increases the area of the zones with the highest \( \tau_{10} \).

6.3.3.8. Fluxtube modification-Effect of fluxtube diameter

This section compares three different fluxtube diameters. Figure 6-11, it can be easily observed that when the diameter of fluxtube increases. Increasing the fluxtube diameter makes the time distribution above the baffle less even.
In this section, we combine two modifications: inlet gas distributor configurations and baffles. Figure 6-12 shows that combine other inlet gas distributor configurations (the Western and the Eastern) with the asymmetrical baffle does not increase the asymmetrical baffle performance. Moreover, it shortens the formation to stripper time in the part below the baffle. For the part above the asymmetrical baffle, high $\tau_{10}$ value area is on the western side with the Western inlet configuration and moves to close to the east side when the gas inlet changes to the eastern side. It indicates that the inlet gas distributor configuration affects the whole bed fluidization performance.

Figure 6-14 shows the effects of combining different inlet gas distributor configurations symmetrical baffles with three different fluxtubes. Similar solids distribution maps are
obtained with different inlet gas distributor configurations, as observed with the symmetrical baffle without a fluxtube.

Figure 6-12 $\tau_{10}$ time distribution map with asymmetrical baffle combined with different gas inlet configurations under superficial gas velocity of 0.4 m/s
Figure 6-13 $r_{10}$ time distribution map with symmetrical baffle combine inlet gas distributor configuration under superficial gas velocity of 0.4 m/s
Figure 6-14 $\tau_{10}$ time distribution map with different fluxtubes combine gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Short-fluxtube, (b) Regular-fluxtube, (c) Long-fluxtube
6.4.2 Gas velocity vector map

Figure 6-15 $\tau_{10}$ distribution map and local gas velocity vector in the formation zone of western gas inlet configuration under superficial gas velocity of 0.4 m/s
This section investigates the local gas velocity vector to understand how it might affect the solids residence times. Figure 6-15 shows that, with the Western inlet gas configuration, at every height, the zone of highest $\tau_{10}$ corresponds to the zone with strong upward gas velocity. Similar correlations are observed with the Even and the Eastern inlet gas distributor configurations, as shown in Figure 6-16 and Figure 6-17. In Figure 6-18 and Figure 6-19, which are the gas velocity vector and time distribution for both two extra inlet gas distributor configurations: center-inlet-1 and center-inlet-2, in which most of the gas goes up straight in the center part and goes down in the near-wall part. Since the gas inlet opened is in the center of the distributor, which also make them more evenly distributed compared to other distributor configurations.

Figure 6-16 $\tau_{10}$ distribution map and local gas velocity vector in the formation zone of even configuration under superficial gas velocity of 0.4 m/s
Similar correlations between the gas flow and the solids time distribution were also observed with the baffled fluidized bed. Adding a baffle changes the gas flow pattern and induces solid recirculation that enhances the radial mixing, as shown in Figure 6-20 and Figure 6-21. Moreover, with fluxtube, gas bubbles go through the fluxtube, as shown in Figure 6-21, which also explains gas go through the fluxtube in the time distribution contour map. Furthermore, by adding symmetrical baffles, more gas recirculation is found above the baffles, which is also the reason for more evenly distributed high $\tau_{10}$ value above the symmetrical baffles.

**Figure 6-17** $\tau_{10}$ distribution map and local gas velocity vector in the formation zone of eastern configuration under superficial gas velocity of 0.4 m/s
Figure 6-18 $\tau_{10}$ distribution map and local gas velocity vector in the formation zone of center-inlet-1 under superficial gas velocity of 0.4 m/s
Figure 6-19. $\tau_{10}$ distribution map and local gas velocity vector in the formation zone of center-inlet-2 under superficial gas velocity of 0.4 m/s
Figure 6-20 $\tau_{10}$ distribution map and local gas velocity vector in the formation zone of regular-fluxtube under superficial gas velocity of 0.4 m/s
Figure 6-21 \( \tau_{10} \) distribution map and local gas velocity vector in the formation zone of asymmetrical baffle under superficial gas velocity of 0.4 m/s
\( \tau_{10} \)  
gas velocity vector

Figure 6-22 \( \tau_{10} \) distribution map and local gas velocity vector in the formation zone of symmetrical baffle under superficial gas velocity of 0.4 m/s
Dispersion of particle in the lateral direction

The particle dispersion in the lateral direction by releasing 100 particles from each location every 0.1 s and predicting the final location for each of the 100 particles. To get the dispersion in the lateral direction, the standard deviation of the x-coordinate values of the 100 particles.

Figure 6-24 shows the dispersion of particles in the lateral direction under five different inlet gas distributor configurations. The Even configuration (Figure 6-24b) is better than other inlet configurations, with most of the central region with a standard deviation of over 0.07. However, the center-inlet-1 configuration (Figure 6-24c) provides more intense
lateral dispersion of particles formed in some particular areas and may be attractive if the area where particles are formed can be selected. The other inlet gas distributor configurations do not provide as intense a particle dispersion in the lateral direction.

Adding different baffles to the fluidized bed, as shown in Figure 6-25, separates the bed into two zones: the lateral dispersion is always intensified above the baffle. The impact of the asymmetrical baffle on lateral dispersion (Figure 6-25 (2)) greatly depends on the inlet gas distribution. With the even distribution (b), the lateral dispersion is reduced below the baffle while with the other gas inlet distributions ((a), (c)), the lateral dispersion is greatly intensified everywhere in the bed. The symmetrical baffle (Figure 6-25 (3)) and the asymmetrical baffle with regular fluxtube (Figure 6-25 (4)) do not promote lateral dispersion as effectively as the asymmetrical baffle with the Eastern and Western inlet gas configurations.

Lateral dispersions obtained with three different fluxtube lengths, and the asymmetrical baffle, were compared in Figure 6-26. The long fluxtube promotes lateral dispersion better than two other sizes.
Figure 6-24 The standard deviation of x-value after 10 s of the formation zone in fluidized bed under different inlet configurations (a) Western (b) Even (c) Center-inlet-1 (d) Center-inlet-2 (e) Eastern under superficial gas velocity of 0.4 m/s
Figure 6-25 The standard deviation of x-value after 10 s of the formation zone in fluidized bed under different baffle geometry under three different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western (b) Even (c) Eastern
Figure 6-26 The standard deviation of x-value after 10 s of the formation zone in fluidized bed under asymmetrical baffle with different fluxtube sizes under three different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western (b) Even (c) Eastern
6.4.4 Particle dispersion in axial direction

Figure 6-27 The standard deviation of y-value after 10 s of the formation zone in fluidized bed under different inlet configurations (a) Western (b) Even (c) Center-inlet-1 (d) Center-inlet-2 (e) Eastern

The particle dispersion in the vertical direction is discussed in this section. It is different from the mixing quality in the lateral direction, which is shown in Figure 6-24 and Figure 6-27. In Figure 6-27, it can be discovered that by using different inlet gas distributor configurations can improve the solid mixing quality in the vertical direction, especially the center-inlet-2. Moreover, as for the center-inlet-2, it improves the mixing quality more evenly with Even configuration compared to other inlet gas distributor configurations.

The solid mixing in different baffled fluidized bed was also compared in this section. The gas inlet works better on the lower section. However, the baffle works better on increasing the upper section of the fluidized bed. Moreover, among all the baffled fluidized beds, the asymmetrical baffle with the Eastern inlet gas distributor configuration works better than
others on increasing the solid disperse in the upper section, and regular fluxtube works better considering the whole formatting zone.

The mixing quality with different fluxtube geometry was also considered here. Figure 6-29 shows the results with three different fluxtube lengths combine with the even, Western and Eastern inlet gas distributor configuratons. The differences between each other are not visible. In all three different conditions, the short fluxtube works better than the two others

If the solid mixing in both lateral direction and the vertical direction is considered, it is difficult to pick up one best configuration for both. However, it can be concluded that the baffle can improve the solid mixing in the upper section, and the inlet gas distributor configuratons function more on the lower section. It is also reasonable as inlet gas distributor configuraton influence the lower part more than the upper section. After the gas-solid flow goes pass the baffle section. The conclusion is constructive for the applications with the injection location.
Figure 6-28 The standard deviation of y-value of after 10 s the formation zone in fluidized bed under different baffle geometry under three different gas inlet configurations (a) Western case (b) Even case (c) Eastern case
Figure 6-29 The standard deviation of y-value after 10 s of the formation zone in fluidized bed under different fluxtube sizes under three different gas inlet configurations (a) Western case (b) Even case (c) Eastern case
6.4.5 Particle dispersion in different zones

The final distance between all possible binary combinations of the 100 particles is calculated, and the standard deviation of all the distances is recorded for each location is the other way to check the solid dispersion properties in the fluidized bed reactor.

We separate the formation zone into three different sections, as shown in Figure 6-5. In the top section, as shown in Figure 6-30, with the asymmetrical baffle in the fluidized bed, the particle disperses more effectively, especially with the Western and Eastern inlet gas distributor configurations.

The particles from the baffle section (Figure 6-30) are dispersed more effectively than those from other sections. Moreover, it easy to find out that the part under the baffle has lower values compared to other parts. It can be discovered that the baffle blocks the part under that and, at the same time, increased the solid movement above the baffle. Moreover, among all the baffled fluidized beds, the symmetrical baffle with Eastern inlet configuration and the asymmetrical baffle with Western and Eastern inlet configurations work better on increasing the center part of the reactor compared to others.

For the bottom section (Figure 6-32), adding a baffle reduces the particle dispersion. This agrees with the results on particle dispersion in the lateral and vertical directions, as shown in Figure 6-25 and Figure 6-28.
Figure 6-30 Standard deviation contour after 10 s of top section under three different inlet configurations
(a) Western case (b) Even case (c) Eastern case
Figure 6-31 Standard deviation contour after 10 s of baffle section under three different inlet configurations (a) Western case (b) Even case (c) Eastern case
Figure 6-32 Standard deviation contour after 10 s of bottom section under three different inlet configurations (a) Western case (b) Even case (c) Eastern case
6.4.6 Dispersion index by averaged distance $R_{(a,b)}$

The final distance between all possible binary combinations of the 100 particles is calculated, and the dispersion index $R_{(a,b)}$ is applied in this section to check the particle dispersed condition. Figure 6-33 compared the contours of the dispersion index $R_{(a,b)}$ for five different inlet gas distributor configurations, all the values are smaller than one, which means that the dispersed is still worse than the random dispersed. However, if compared between the different inlet gas distributor configurations, it is easy to discover that the $R_{(a,b)}$ in the Even inlet is more evenly distribute. Moreover, center-inlet-1 and Eastern inlet have a more high-value region in the low part of the formation zone, which is similar to the tendency of particle dispersion in the axial direction.

![Figure 6-33 Dispersion index $R_{(a,b)}$ contour after 10s for the formation zone under different gas inlet configurations (a)Western (b) Even (c) Center-inlet-1 (d) Center-inlet-2 (e) Eastern under superficial gas velocity of 0.4 m/s](image)

- (a) Western
- (b) Even
- (c) Center-inlet-1
- (d) Center-inlet-2
- (e) Eastern

Under superficial gas velocity of 0.4 m/s
(1) No baffle

(2) Asymmetrical baffle
Figure 6-34 Dispersion index $R_{(a,b)}$ contour after 10s for the formation zone with and without baffle under different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western case (b) Even case (c) Eastern case
Figure 6-34 shows the dispersion index $R_{(a,b)}$ contour for the formation zone with and without baffle under different gas inlet conditions. By adding asymmetrical baffle, it increases the dispersion index $R_{(a,b)}$ above the baffle and lower the part below the baffle for all three gas inlet conditions shown in Figure 6-34 (a) and Figure 6-34 (b). Same for adding symmetrical baffle to the fluidized bed shown in Figure 6-34 (c). Moreover, among the baffled fluidized bed, combing Eastern inlet configuration with the baffle is better than other conditions. However, for the asymmetrical baffle with fluxtube, combing Western inlet with fluxtube is best compared to two other inlet gas distributor configurations.

Figure 6-35 shows the fluxtube of different lengths combining with three different inlet gas distributor configurations. For the short fluxtube shown in Figure 6-35 (1) shows there are no differences between the three inlet gas distributor configurations. However, for the regular fluxtube and long fluxtube combing with different inlet gas distributor configurations, the results are different. Regular fluxtube combing with the Western inlet is better than two other inlet configurations shown in Figure 6-35 (2). However, Figure 6-35 (3) for long fluxtube, it seems that combing with Eastern inlet configuration is better than two others.

Moreover, the results consistent with the results of the standard deviation of $y$ value. Among all the contours, combine long fluxtube with Eastern inlet has the best particle disperse in the upper zone compared to others. Besides, the fluxtube with different diameter also compares in Figure 6-36, in which three different diameters are compared for the dispersion index $R_{(a,b)}$ contour in the formation zone. In which, it shows that the $D_2=6$ cm is best compares to others.
(1) Short fluxtube (12 cm)

(2) Short fluxtube (12 cm)
Figure 6-35 Dispersion index $R_{(a,b)}$ contour after 10s for the formation zone with baffle with fluxtube of different lengths under different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western case (b) Even case (c) Eastern case
The final distance between all possible binary combinations of the 100 particles is calculated, and the dispersion index $R_{stv(a,b)}$ is applied in this section to check the particle dispersed condition as the second method. For the five different inlet gas distributor configurations, the $R_{stv(a,b)}$ value contour is shown in Figure 6-37. The result shows the same tendency, as shown in Figure 6-33. So as the results for the fluidized bed without and with different baffles show in Figure 6-38.

However, the effect of different fluxtube length and inlet gas distributor configuration on the dispersion index $R_{stv(a,b)}$ is not as apparent as on the dispersion index $R_{(a,b)}$. In the short fluxtube, there is not much difference between the three different configurations shown in Figure 6-38 (a).
For the fluxtube with different diameter, both the dispersion index $R_{(a,b)}$ contour for the fluxtube with $D_2$ and $D_3$ are better than the fluxtube with a small diameter.

In conclusion, both the dispersion index $R_{stv(a,b)}$ and the dispersion index $R_{(a,b)}$ are consistent with each other and independent to show the particle dispersion quality with baffle and gas inlet modification.

![Figure 6-37 Dispersion index $R_{stv(a,b)}$ contour after 10s for the formation zone under different gas inlet configurations (a) Western (b) Even (c) Center-inlet-1 (d) Center-inlet-2 (e) Eastern under superficial gas velocity of 0.4 m/s](image)

Figure 6-38 Dispersion index $R_{stv(a,b)}$ contour after 10s for the formation zone with and without baffle under different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western case (b) Even case (c) Eastern case
Figure 6-39 Dispersion index $R_{stv(a,b)}$ contour after 10s for the formation zone with baffle with fluxtube of different lengths under different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western case (b) Even case (c) Eastern case
Conclusion

The objective of this study is to see how inlet gas distributor configuration and different internal geometry can affect the solid mixing in both lateral and horizontal direction. For the time it takes for the particles to go from formation zone to stripper zone. It is essential to minimize the percent of particles with short formation to stripper times. Center-inlet-1 is the best modification, and the second-best modification is the Eastern configuration. As for the superficial gas velocity, 0.6 m/s is the best compared to 0.4 m/s, 0.8 m/s and 1.0 m/s. When adding baffle in the fluidized bed, fluxtube is better when the injection location is above the baffle. Among all the fluxtube with different sizes discussed in this chapter, the long fluxtube works better than others. Combine inlet gas distributor configurations and adding baffles together does improve the performance but not apparent.

Different inlet gas distributor configurations do not work on improving the particle dispersion in lateral direction. However, baffle can greatly improve the solid dispersion above it and reduce it below. As for the dispersed in the vertical direction, the inlet gas

Figure 6-40 Dispersion index $R_{(a,b)}$ contour after 10s for the formation zone with baffle with fluxtube of different diameters with even gas inlet configuration (a) $D_1 = 2$ cm (b) $D_2 = 6$ cm (c) $D_3 = 8$ cm

6.5 Conclusion

The objective of this study is to see how inlet gas distributor configuration and different internal geometry can affect the solid mixing in both lateral and horizontal direction.

For the time it takes for the particles to go from formation zone to stripper zone. It is essential to minimize the percent of particles with short formation to stripper times. Center-inlet-1 is the best modification, and the second-best modification is the Eastern configuration. As for the superficial gas velocity, 0.6 m/s is the best compared to 0.4 m/s, 0.8 m/s and 1.0 m/s. When adding baffle in the fluidized bed, fluxtube is better when the injection location is above the baffle. Among all the fluxtube with different sizes discussed in this chapter, the long fluxtube works better than others. Combine inlet gas distributor configurations and adding baffles together does improve the performance but not apparent.

Different inlet gas distributor configurations do not work on improving the particle dispersion in lateral direction. However, baffle can greatly improve the solid dispersion above it and reduce it below. As for the dispersed in the vertical direction, the inlet gas
distributor configuration works better on the section below the baffle and the baffle functions well for the upper section above the baffle.

Separate the formation section into three zones are convenient for deciding the secondary injection location. If located in the part above baffle, the asymmetrical baffle combing Eastern and Western inlet gas distributor is best choice. And if located in the bottom section, the configuration without baffle works better than others.

Both the dispersion index \( R_{a,b} \) and the dispersion index \( R_{stv(a,b)} \) are independent enough to show the particle dispersion condition. Center-inlet-1 and Eastern inlet configuration have high index value compared to others. Moreover, the asymmetrical baffle and symmetrical baffle can improve the upper section above the baffle.

6.6 References


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Chapter 7

7 Gas mixing in gas-solid fluidized bed reactor

7.1 Introduction

Fluidized bed technology has been widely applied in industrial processes such as fossil fuel combustion, gasifier, drying, fluid catalytic cracking and others, due to its excellent mass and heat transfer, high efficiency of gas-solid contact [1,2]. In most fluidized beds, all the fluidization gas through a gas distributor that is located at or near the bottom of the bed, while in some processes, part of the gas is added gradually with increasing bed height as secondary injection gas.

Gas mixing can be categorized into vertical mixing and horizontal mixing based on the direction of mixing. Normally, the vertical mixing is two or three times greater than the horizontal mixing due to lacking radial momentum of fluidization gas for a conventional fluidized bed [6]. There are also a lot of studies that focus on the mixing of gas in fluidized bed [5–8].

Mixing behaviours of gas and solid play a significant role in the conversion and selectivity of chemical reactions in fluidized bed reactors [3,4]. In most fluidized bed reactors, reactant conversion and product selectivity are favoured when lateral mixing is maximized, and vertical mixing is minimized. In a fluidized bed gasifier, for example, evenly and well-distributed thermal energy is needed [5]. In a fluidized bed combustor, fuel mixing also has a significant impact on the overall performance: as the horizontal mixing of fuel improves, it reduces incomplete combustion.

In the Fluid Coking™ process, heavy oil is injected into a fluidized bed of hot coke particles that provide enough heat for the heavy liquid to crack and vaporize. Vapors exit at the top the bed while the particles exit at the bottom, through a stripper region where the valuable vapors trapped between the coke particles are displaced by injected steam. Reducing the vapors reaching stripper zone is an essential for the Fluid Coking process to minimize the loss of valuable vapors with the exiting solids and to minimize the fouling of the stripper internals by coke deposits [10–13]. Gas mixing in the vertical direction should,
therefore, be minimized. On the other hand, gas mixing in the lateral direction is beneficial, as it helps prevent any local vapor saturation which would slow the vaporization process of the heavy oil [11–15].

The gas dispersion coefficient is often used to determine the degree of gas mixing. Gas dispersion in a fluidized bed can be determined by injecting tracer gas into the fluidized beds at a steady-state, and monitoring the concentration of tracer at different locations [3,6,7,16]. Both the axial and lateral tracer gas concentration profiles were used to study the gas mixing quality [3,4]. Chyang et al. [16] introduced the radial dispersion coefficient of kurtosis, which is applied to characterize the shape of the tracer gas concentration profile.

The mixing index is another parameter that is often used to represent the degree of gas mixing in fluidized bed. It is usually based on the statistical analysis and the standard deviation of a certain parameter that varies with the tracer concentration. However, the mixing quality may be affected by the injection and the monitoring probe inserted inside the fluidized bed. This is avoided with non-invasive concentration measurement. For example, MRI has been applied to measure the concentration laser-polarized xenon (129Xe) [19]. However, the MRI technique is limited by its cost and cannot be used with large equipment. Dang.et.al [19] applied a novel infrared technique coupled with the Digital Image Analysis for tracer gas detection in both bubbling and turbulent fluidization regimes.

Gas dispersion mixing in all directions tends to increase with superficial gas velocity for any distributor [3] [6] [16][17] [19]. Werther et al. [9] used the Peclet number to correlate the gas dispersion coefficient with other parameters. The response surface methodology (RSM) has been applied to study the effect of operation and geometry parameters on the radial gas mixing in the fluidized bed [7]. A general governing equation is formulated for the gas dispersion in the fluidized bed reactor by Nautiyal et al. [18], and an analytical solution is derived to estimate the dispersion coefficient in various directions.

In the fluidized bed, when the gas bubbles rise in the bed, the surrounding solids will replace the void, which is the main reason for the lateral mixing of solids and gas dispersion. Meanwhile, when the bubbles eruption and collapse occur on the bed surface,
the axial mixing of gas and particles [3,4,6]. However, it is also essential to limit the back mixing and maintain an excellent gas-solid contract. Various methods have been investigated to improve the performance of the fluidization systems. Guo et al. [20] discovered that the gas dispersion coefficient increased with superficial gas velocity first and then decreased gradually after hitting a maximum value. A similar tendency was also discovered by Dang et al. [19] by applying the newly developed IR experimental technique couped with DIA. Chyang et al. [6] found out that the coefficient of horizontal gas dispersion increased with superficial gas velocity ($3.5U_{mf} < U < 6.5U_{mf}$) in specific superficial gas velocity range, and $(U-U_{mf})/U_{mf}$ and $d_p$ have a great impact on the coefficient, so as the design of the gas distributor. A better extent of dispersion of the coefficient can be obtained by decreased the dead zone and lower the open area ratio.

The previous chapter showed that the mixing of fluidized solids can be modified by either adjusting the initial distribution of the fluidizing gas or adding internals such as baffles. Similarly, past studies have shown the gas mixing can be modified by changing the gas distributor [6] or by adding baffles [4,21]. For example, a louver baffle proved to be efficient in suppressing gas backmixing [4,21–23]. In the stripper section, ring baffles can reduce gas back-mixing through the sheds [12]. Cui et al. [24] explored the gas and solid mixing in the stripper with and without baffle and found that the gas mixing is most intense in the stripper core.

### 7.2 Statement

The present chapter is focused on gas mixing in both bubbling and turbulent regimes. The main objectives are to reduce the gas reaching the stripper zone by minimizing vertical gas mixing, and to promote lateral gas mixing. The effect of gas distributor configuration and different internals are investigated. In particular, the following topics will be discussed in this chapter:

1. The correlation between the superficial gas velocity and gas mixing;
2. Effect of different gas inlet configuration;
3. Effect of baffle geometry and location;
4. The correlation between the gas disperses and solid mixing.
7.3 Methodology and Experimental Setup

In this section, the gas-solid two-phase model introduced in Chapter 2 is used to simulate the fluidization. We consider two aspects of gas mixing properties. The first aspect is the dispersion of gas, which will mainly focus on the gas lateral dispersion coefficient. The second aspect is the time for the gas to go from the formation zone to the stripper zone. The MATLAB software is applied for the calculation in this chapter.

7.3.1.1 Flowchart for gas molecule tracking

The primary numerical tracking model in this chapter is based on the TFM numerical results. The vertical and horizontal components of the gas at 10 s (steady state) were used as the initial condition in this study. The flowchart for tracking gas molecule is shown in Figure 7-1.
For \( p = 1: N \) (starting location point) \( N \) is the number of location \((x_p, y_p)\)

Read the location value \( x, y \)

- For \( i = 1: \) number (data files)
  
  \[
  \text{point}_a = \text{results}_\text{data}\{i+3\}; \text{read data from result (time-}i, x, y, u_x, u_y)\]

- Finding the first point in \( \text{point}_a \) by searching for the closest point \( x_1, y_1 \)
  
  First point \((x_1, y_1, u_{x1}, u_{y1})\)

  \[
  \begin{align*}
  xx &= x_1 + u_{x1} \times \Delta t \\
  yy &= y_1 + u_{y1} \times \Delta t
  \end{align*}
  \]

  if \( xx <= 0, \) the point will hit the wall and stop, \( \Delta t = 0.1s \);

  if \( yy <= 0, \) the point already goes to the stripper zone, \( \Delta t = 0.1s \);

- if \((xx>0 \text{ and } yy>0)\) continue to the next time file to looking for the data

  For \( j = i+1: \) number

  \[
  \text{point}_b = \text{results}_\text{data}\{j+3\};
  \]

  Finding the second point in \( \text{point}_b(j) \) by searching for the closest point \( xx, yy \) in \( \text{point}_b(j) \)

  Second point \((x_2, y_2, u_{x2}, u_{y2})\)

  if \( (y_2 <= 0) \)

  \( \Delta t = t_j - t_i \) record \( \Delta t \) and move to \( i+1 \) and continue

  break

  else

  \[
  \begin{align*}
  xx &= x_2 + u_{x2} \times \Delta t \\
  yy &= y_2 + u_{y2} \times \Delta t
  \end{align*}
  \]

  Continue to look for the next point in \( \text{point}_b(j+1) \) until the \( y < 0 \) record \( \Delta t \)

  \text{End}

\text{End}

Finish all the location point \( p = 1: N \) (location number) and create the contour of the det-time quantiles values at different locations.

\textbf{Figure 7-1 Flowchart for gas molecule tracking procedure}
7.3.2 Dispersion of gas

To evaluate the gas mixing quality in the fluidized bed reactor, we consider axial (vertical) mixing and radial (horizontal) mixing separately. Most experimental studies about gas mixing were conducted by injecting a tracer gas into the fluidized bed and subsequently sampling from various points [16]. A similar idea is applied in this study: tracer gas is released from the individual location in the formation zone, and is tracked for 10 s. The 2-D starting location (0.25,0.306) was chosen as an example shown in Figure 7-5. Every 0.1 s, one tracer gas is released from the same location, and tracked for 10 s, i.e. hundred tracer gas releases over a 10 s period. The trajectory of each tracer gas is recorded, and different aspects of how well and quickly tracer gas disperses in the fluidized bed are calculated and compared for various operating conditions and configurations.

7.3.2.1 Percentage of tracer gas go to stripper zone

In the fluidized bed formation zone, for each location in the formation zone, the tracer gas is tracked during 10 s, and whenever the gas goes to the stripper zone (y<0), it will be recorded. And during the 10 s for the tracer gas, for each formation location, the proportion of tracer gas that reaches the stripper zone is recorded. The percentage of tracer gas go down to the stripper zone is recorded for each location and discussed.

7.3.2.2 Tracer gas lateral dispersion properties

To determine the tracer gas mixing rate in the fluidized bed reactor, researchers have used axial and lateral dispersion coefficients, or a mixing index [8,18,20,24]. Based on the tracking gas method mentioned in 7.3.1.1, the formation zone is divided into 550 different locations. At each location, tracer gas is released every 0.1 s and tracked for 10 s. To evaluate the axial dispersion, the standard deviation of the 100 final y coordinates is calculated, and, for the lateral dispersion, the standard deviation of the final x coordinates is calculated.
Figure 7-2 Schematic diagram of the lab-scale bubbling fluidized bed with formation zone and stripper zone
Figure 7-3 Schematic diagram of initial points selected in the formation zone
Figure 7-4 Different gas inlet configurations used in the two-dimensional simulation model
Figure 7-5 A example of sampled gas released from same location
7.3.3 Gas travelling time distribution

This method was selected to track statistical distributions of the time for the gas starting from a specific location in the formation zone to go to the stripper zone. As shown in Figure 7-2, points are selected in the formation zone: 3363 points for the no baffle fluidized bed reactor and 7014 points for the baffled fluidized bed. Gas at each location is tracked to see how long it takes for it to reach the stripper zone. As in the previous chapter for solids mixing, this is repeated every 0.1 s for 10 s. The cumulative time distribution is collected for each location. The cumulative formation-to-stripper time distributions are important for industrial applications such as Fluid Coking, as it is important to minimize bypassing of vapors from formation zone to stripper zone. The maps presented in this chapter can be used to select the best locations for heavy oil injection.

Because of the large number of locations considered in this study, it is difficult to compare all their cumulative time distributions and specific times are reported instead: $t_{10}$ is the time at which 10% of gas has reached the stripper, $t_{25}$ corresponds to 25%, and $t_{50}$ is the median formation to stripper time.

This chapter will use dimensionless time distributions, which are the time $t_i$ divided by the average time $t_{\text{average}}$.

$$\tau_i = \frac{t_i}{t_{\text{average}}}$$  \hspace{1cm} (1)

$i = 10, 25, 50$
7.4 Results

7.4.1 Percent of gas go to stripper section

Reducing the amount of gas going to the stripper zone is also important in the process.

Figure 7-6 shows that most of the gas from the central locations part goes up and only the gas from the lower wall regions has a high probability of reaching the stripper zone. It also shows that with all inlet configurations, increasing the gas velocity increases the regions in the center from which less number of gas reaches the stripper. It also shows that as the gas velocity increases, the part near the wall and in the lower part, from which, there are more number of gas go to the stripper zone also increases. Figure 7-7 shows that the Even inlet gas configuration is best. With the Even configuration, the region from which a low proportion of the gas reaches the stripper is much larger.

![Graph showing gas distribution](image)

(a) Western inlet gas distributor configuration
Figure 7-6 The percent of gas go to stripper zone after 10 s of the formation zone in fluidized bed under different inlet configurations for different superficial gas velocities
Figure 7-7 The percent of gas go to stripper zone after 10 s of the formation zone in fluidized bed under different inlet configurations under superficial gas velocity of 0.4 m/s (a) Western (b) Even (c) Center-inlet-1 (d) Center-inlet-2 (e) Eastern

Figure 7-8 shows that, when compared to the fluidized bed with no baffle, all the baffles work well on reducing the proportion of the gas formed above the baffle that would reach the stripper. On the other hand, a baffle increases the proportion of gas formed below the baffle that would reach the stripper.

Figure 7-9 compared asymmetrical baffles with a fluxtube of different lengths for three different gas inlet configurations. With the short-fluxtube, the three different gas inlet configurations gave similar results, as shown in Figure 7-9 (1). With the regular-fluxtube and long-fluxtube, the Even inlet is better than the other inlet configurations. In general, the short-fluxtube gives better results than the other fluxtubes. As for the diameter of the fluxtube, the fluxtube with the largest diameter is the best, as shown in Figure 7-10.
Figure 7-8 The percent of gas go to stripper zone after 10 s of the formation zone in fluidized bed with different baffle under three different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western (b) Even (c) Eastern
Figure 7-9 The percent of gas go to stripper zone after 10 s of the formation zone in fluidized bed with fluxtube of different lengths under three different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western (b) Even (c) Eastern
Figure 7-10 The percent of gas go to stripper zone after 10 s of the formation zone in fluidized bed with fluxtube of different diameters, with Even inlet gas configuration under superficial gas velocity of 0.4 m/s

7.4.2 Dispersion of gas in lateral direction

The gas dispersion in the lateral direction by releasing tracer gas 100 times from each location every 0.1 s and predicting the final location for each of the 100 injections. To get the dispersion in the lateral direction, the standard deviation of the final x-coordinate values is calculated for the 100 injections. Strong lateral gas dispersion is preferable.

Figure 7-11 shows the standard deviation of the x-coordinate values under different superficial gas velocities for the Western, the Even, and Eastern gas inlet configurations. For all gas velocities, the Western and Eastern gas inlet configurations greatly increase the standard deviation value when compared to the Even gas inlet configuration. And clearly with higher superficial gas velocity, the standard deviation value is much higher than the low superficial gas velocity as shown in Figure 7-11 (a) and (c) separately. However, when the superficial gas velocity is equal to 0.6 m/s, it is worse than with three other velocities.
Figure 7-12 shows results for the five different inlet gas distributor configurations. It confirms that the best lateral dispersion is obtained with the Western and Eastern inlet gas distributor configuration.

(a) Western inlet gas distributor configuration

(b) Even inlet gas distributor configuration
The standard deviation of $x$-value after 10 s of the formation zone in fluidized bed under different inlet configurations for different superficial gas velocities.

Figure 7-13 compares the standard deviation of $x$-value contour for the no baffle fluidized bed, asymmetrical baffle, symmetrical baffle and fluxtube under three different gas inlet configurations. It shows that baffle does not increase the standard deviation of $x$-value in the formation zone. Figure 7-13 (a) shows that both the Western inlet and Eastern inlet are much better than the Even inlet gas configuration. And the Western inlet configuration is more evenly distributed with high standard deviation value.

Figure 7-14 compared fluxtube with different lengths under three different gas inlet configurations. Figure 7-14 (a) shows no differences for the fluidized bed with short-fluxtube under three different gas inlet configurations. However, with baffle insider the fluidized bed, it shows different tendency. For the short fluxtube, there are no obvious differences between three different gas inlet configurations, and combine with Eastern inlet configuration seems a little better than others shown in Figure 7-14 (1).
The long fluxtube with the Even inlet seems better than the one with Western and Eastern inlet configurations shown in Figure 7-14 (3). Fluxtube with different diameters are compared in Figure 7-15. It shows that with diameter of fluxtube increases, the gas disperses is getting better.

Figure 7-12 The standard deviation of x-value after 10 s of the formation zone in fluidized bed under different inlet configurations (a) Western (b) Even (c) Center-inlet-1 (d) Center-inlet-2 (e) Eastern under superficial gas velocity of 0.4 m/s
Figure 7-13 The standard deviation of x-value after 10 s of the formation zone in fluidized bed with baffles under three different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western (b) Even (c) Eastern
Figure 7-14 The standard deviation of x-value after 10 s of the formation zone in fluidized bed under asymmetrical baffle with different fluxtube lengths under three different gas inlet configurations under superficial gas velocity of 0.4 m/s (a) Western (b) Even (c) Eastern
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Figure 7-15 The standard deviation of x-value after 10 s of the formation zone in fluidized bed under different baffle diameter, with the Even inlet gas distributor configuration under superficial gas velocity of 0.4 m/s (a) Western (b) Even (c) Eastern

7.4.3 Gas time distribution

For the gas back-mixing, which we already discussed in 7.4.1, that a fraction of the gas goes down to the stripper section under different operating conditions. So it is also essential to figure out the time for a gas molecule to travel from the formation zone to the stripper zone. As with particle tracking, but restricting to the gas molecules that are known to reach the stripper zone, the distribution of the time taken by a gas molecule from formation to stripper section is obtain by tracking the gas molecules that go down to the stripper zone. The cumulative time distribution for each location in the formation zone is calculated and the dimensionless $\tau_{10}$ will discussed in the following section.
7.4.3.1 Gas time distribution under different inlet gas distributor configurations

This section uses the method mentioned before to compare the gas (vapor) formation-to-stripper time distribution map for different gas inlet configurations. We have five different gas inlet configurations, as shown in Figure 7-4.

Figure 7-16 (a) compared the $\tau_{10}$ distribution map for five different conditions, where $\tau_{10}$ is the dimensionless time at which 10% of gas released at a given formation location have reached the stripper (this is a similar concept to the particle $\tau_{10}$ distribution shown in Chapter-7. As expected, for all the five distributor configurations, $\tau_{10}$ is smaller when the initial location is closer to the stripper zone. For initial formation locations above 0.8 m, the Even distributor configuration is preferable as provides larger $\tau_{10}$ than the other inlet configurations. However, if the initial location is lower than 0.4 m, other inlet configurations are better than the Even configuration as they provide higher $\tau_{10}$ in the central region.

The results obtained with the dimensionless times $\tau_{25}$ and $\tau_{50}$ confirm the results obtained with $\tau_{10}$ (Figure 7-16). Center-inlet-1 and Center-inlet-2 are the best gas inlet configurations for the initial locations below 0.4 m; the next best configuration is the Eastern configuration. The Even distributor configuration is the best for initial formation locations above 0.8 m. In all cases, the best vapor formation location is in the central region.

The trends of the formation-to-stripper time are similar for gas and particles (see Chapter 6). This is because most of the movement of particles and emulsion gas are induced by bubbles or voids. When a bubble rises in the bed, it carries particles in its wake, which are released and move back downward in the emulsion phase.

As the time distribution contours for $\tau_{10}$, $\tau_{25}$ and $\tau_{50}$ are similar, the following discussion will focus on the time distribution contour for $\tau_{10}$. The dimensionless time $\tau_{10}$ is especially important as it characterizes the fastest 10% of the gas, is more likely to contain heavy vapors.
(a) $\tau_{10}$

(b) $\tau_{25}$
Figure 7-16 Time distribution map for different inlet gas distributor configurations under superficial gas velocity of 0.4 m/s

7.4.3.2 Effect of different superficial gas velocity on the time distribution

Figure 7-17 confirms that the best gas inlet configurations identified for a superficial gas velocity of 0.4 m/s (Figure 7-16) remain the best configurations for gas velocities ranging from 0.4 to 1 m/s. Figure 7-17 (b) shows that, for the Even gas inlet configuration, a superficial gas velocity of 0.8 m/s seems to provide more locations with higher $\tau_{50}$ values.
(a) Western case

(b) Even case
Internal baffles provide a relative low-cost measure to improve fluidized bed performance. For the Fluid Coking(TM) process, ExxonMobil patented the use of baffles to limit the flow of wet solids reaching the stripper [11–13]. The aim of this section is to determine whether a baffle can eliminate short formation-to-stripper times when the formation zone is above the baffle.

Figure 7-18 shows the effect of the asymmetrical baffle, with and without fluxtube on modifying the time distribution in the formation zone in the fluidized bed with Even gas inlet configuration. The asymmetrical baffle, with or without fluxtube, improves the time distribution on the western side of the reactor, by increasing \( \tau_{10} \), but worsens the time
distribution on the other side. Slightly better results are obtained when the baffle is equipped with a fluxtube.

**Figure 7-19** shows the impact of the symmetrical baffle with and without fluxtube. It is similar to the impact of asymmetrical baffle with and without fluxtube. The fluxtube has a beneficial effect by increasing the formation locations with larger $\tau_{10}$.

For injection from the western side, an asymmetrical baffle is best. For injection from a central location, a symmetrical baffle with fluxtube is best. Further study will determine the best fluxtube characteristics.

**Figure 7-18** $\tau_{10}$ time distribution map with asymmetrical baffle with and without fluxtube under superficial gas velocity of 0.4 m/s (Western side is left-hand side)
No-baffle  Symmetrical baffle  Symmetrical baffle with fluxtube

Figure 7-19 $\tau_{10}$ time distribution map with symmetrical baffles and symmetrical fluxtube, with the Even inlet gas distributor configuration under superficial gas velocity of 0.4 m/s (Western side is left-hand side)

7.4.3.4 Fluxtube modification-Effect of fluxtube length
This section compares three different fluxtube lengths for the asymmetrical baffle. Figure 7-20 shows similar time distribution maps with the three different fluxtubes. The short fluxtube increases the formation zones with larger $\tau_{10}$ above the baffle.
The other fluxtube parameter considered in this work is its diameter. In Figure 7-21, it can be easily observed that when the diameter of fluxtube increases, the area of high $\tau_{10}$ values above the fluxtube increases.

Asymmetrical baffle  Short-fluxtube (12 cm)  Regular-fluxtube (16 cm)  Long-fluxtube (18 cm)

Figure 7-20 $\tau_{10}$ time distribution map with fluxtubes of different length, with even inlet gas distributor configuration under superficial gas velocity of 0.4 m/s (Western side is left-hand side)

7.4.3.5 Fluxtube modification-Effect of fluxtube diameter
The other fluxtube parameter considered in this work is its diameter. In Figure 7-21, it can be easily observed that when the diameter of fluxtube increases, the area of high $\tau_{10}$ values above the fluxtube increases.
In this section, we combine the two modifications: gas inlet configuration and baffle addition. Figure 7-22 shows results obtained with different gas inlet configurations and the asymmetrical baffle. Focusing on formation above the baffle, with the Western inlet configuration, high $\tau$ areas are on the western side with the Western inlet configuration and shift closer to the east side when the gas inlet is changed to the eastern side. It indicates that the gas inlet configuration affects the whole bed fluidization performance, even in the presence of a baffle.

However, the $\tau$ distribution contour for symmetrical baffled fluidized bed does not have this tendency, as shown in Figure 7-23. There is nearly no impact of the gas inlet configuration on the fluidization performance.
configuration on the time-distribution map above the baffle. This may result from the smaller baffle open area, which weakens the influence of the gas inlet configuration on the region above the baffle.

Figure 7-22. $\tau_{10}$ time distribution map with asymmetrical baffle combine inlet gas distributor configurations under superficial gas velocity of 0.4 m/s (Western side is left-hand side)

Figure 7-24 shows the results for the asymmetrical baffle with three different fluxtubes, for different gas inlet configurations. If one can control the formation zone, then western locations above the baffle with the Western inlet configuration, and the asymmetrical baffle with the short fluxtube would be best.
Figure 7-23 $\tau_{10}$ time distribution map with symmetrical baffle combine gas inlet configurations under superficial gas velocity of 0.4 m/s (Western side is left-hand side)
Figure 7-24 $\tau_{10}$ time distribution map with different fluxtubes combine gas inlet configurations (a) Short-fluxtube (12 cm) (b) Regular-fluxtube (16 cm) (c) Long-fluxtube (18 cm) under superficial gas velocity of 0.4 m/s (Western side is left-hand side)
7.5 Conclusion

The objective of this study is to see how gas inlet configuration and different internal geometry can affect the gas movement.

Changing the gas inlet configuration does not greatly affect the proportion of gas that reaches the stripper. Adding baffles, however, reduces this proportion for formation zones above the baffle and increases it form formation zones below the baffle. Regular fluxtube is better than other fluxtubes to reduce the proportion of gas that reaches the stripper.

Gas dispersion in the lateral direction, can be enhanced by increasing the superficial gas velocity. Eastern and Western gas inlet provide better lateral dispersion than other gas inlet configurations. Adding baffle does not greatly affect the lateral gas dispersion.

For the time for gas to go from formation zone to stripper zone, compared with five different gas inlet configurations, the Even gas inlet is better when the injection location is above 0.4 m; Center-inlet-1 and Center-inlet-2 are the best gas inlet configurations for the formation locations below 0.4 m.

For the time for gas to go from formation zone to stripper zone, in the baffled fluidized bed, if the injection location is above the baffle, the fluxtube improves performance. An asymmetrical baffle is best for injection on the western side, and a symmetrical baffle with fluxtube is best for central formation locations. As for the fluxtube characteristics, changes in length have more impact than changes in diameter. The Long-fluxtube is best. Combining gas inlet configurations and baffles can be used to significantly increase the formation-to-stripper time performance.
7.6 References


Chapter 8

8 Conclusions and Recommendations

8.1 Conclusions

This thesis developed a numerical model to study a gas-solid fluidization system. The model predicted the impact of various fluidization gas distributors and baffles on gas bubbles distribution, solid mixing, and gas mixing under different operating conditions.

In this study, the multi-phase Eulerian-Eulerian two-fluid method (TFM) coupled with the kinetic theory of granular flow (KTFG) was used to investigate the hydrodynamics of particle flows (Geldart Group B) in a lab-scale fluidized bed operating in bubbling and turbulent regimes. The proposed numerical model was validated with experimental data from two different methods under twelve different operating conditions. The hydrodynamics and flow patterns under different superficial gas velocities were studied numerically. Predictions obtained with different gas and particle properties were compared to verify that results obtained with sand and air in the laboratory column were similar to what would be obtained with coke particles and the gases present in fluid cokers, for example.

The model predictions showed that gas bubble distributions can be modified by changing the gas distributor angle, and the effect of the inlet configuration on gas distribution is stronger when the distributor has a large incline. If changing the gas bubble distribution in the upper regions of the bed is the objective, an inclined gas distributor with a large angle is preferable.

Baffles can also be used to change the gas bubble distribution inside the fluidized bed. Baffles can induce strong bed recirculation currents. For baffles with a fluxtube, the fluxtube length and diameter affect the gas distribution, with a stronger effect obtained by varying the length.

Two aspects of solids mixing were studied: solids backmixing to the lower part of the bed, i.e. the “stripper zone”, and solids lateral dispersion. Solids backmixing can be greatly
reduced and lateral dispersion can be greatly enhanced by adjusting the fluidization gas
distribution or inserting a baffle of optimal geometry. A symmetrical baffle, with and
without fluxtube, does not reduce solids backmixing while an asymmetrical baffle, with
and without fluxtube, can reduce solids backmixing but only for solids formed in specific
regions: above the baffle and near the wall where gas inlet is concentrated. Generally, a
baffle enhances lateral dispersion above the baffle and reduces it below.

Two aspects of gas mixing were studied: gas flow to the stripper zone, and gas lateral
dispersion. The gas inlet configuration can be adjusted to prevent gas from reaching the
stripper zone too quickly. The use of a baffle can also prevent gas from reaching the stripper
zone quickly, but only for gas originating at certain locations: in central regions above the
baffle for the symmetrical baffle, and, for asymmetrical baffles, in regions near the wall
above the baffle, with inlet gas injected on the same side as the baffle. Gas dispersion in
the lateral direction can be enhanced by increasing the superficial gas velocity and
adjusting the gas inlet but is not greatly affected by baffles.

8.2 Recommendations

First, predictions of the gas bubble distribution could be improved by segregating the
bubble gas from the emulsion gas, which would require the development of new
algorithms. More accurate methods such as the discrete element method (DEM) or direct
numerical simulation (DNS) approaches could also be tested.

Second, the predictions made for gas and solids mixing in the laboratory column could be
validated by performing dedicated experiments to measure gas and solids mixing. Solids
mixing could be studied with thermal or radioactive tracers. Gas mixing could be studied
with both steady-state and unsteady-state gas tracer injections, with sampling through the
column wall at different locations.

Finally, the most relevant extension of this work would be the application to commercial
scale fluidized beds of the simulation methods validated in this study, and the methods
developed to characterize gas and solids mixing. A potential application, for example,
would be the modelling of large commercial fluid cokers.
Appendices

A video of the fluxtube zone. Take from the regular fluxtube fluidized bed operated under superficial gas velocity of 0.4 m/s.

From which you can find out that most of the gas bubble carried particles go up through the fluxtube, and only a small part of the gas downward through the tube near the western side.
# Curriculum Vitae

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