Mitigation of fouling and liquid losses in a Fluid Coker™: Influence of operating conditions and internals on wet-agglomerates contribution to liquid carry-under

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A thesis submitted in partial fulfillment of the requirements for the Doctor of Philosophy degree in Chemical and Biochemical Engineering

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Abstract

In Fluid Cokers, heavy oil is sprayed into a fluidized bed of hot particles through rings of feed nozzles located at different heights. Unreacted or partially reacted feedstock, trapped in wet-agglomerates, reaches the reactor bottom where it causes fouling or is lost to the system burner. This thesis aims to determine whether modifying the feed distribution between rings or adding baffle(s) can reduce the amount of liquid reaching the reactor bottom.

A model was developed to predict the amount of liquid reaching the reactor bottom. It integrates models for agglomerate formation and re-wetting in spray jets, agglomerate drying through heat transfer from the hot bed particles, and agglomerate breakage from shear in the turbulent bed. The model requires accurate agglomerate trajectories through the reactor to predict re-wetting, drying time, and shear.

Radioactive Particle Tracking (RPT) uses radioactive tracers to provide agglomerate trajectories in a 0.25 m diameter cold model with solid circulation. This thesis shows how RPT accuracy can be enhanced by correcting systematic errors from radiation absorption and random signal fluctuations due to the stochastic nature of radioactive emissions.

The model shows that agglomerates between 10 and 13 mm in diameter are most likely to bring liquid to the reactor bottom. Smaller agglomerates dry quickly. Larger agglomerates are more likely to be broken through shear.

Redistributing feed between rings can reduce the amount of liquid reaching the reactor bottom. Wet-agglomerates formed in the upper regions have more time to dry and break. For example, with five feed rings, redistributing the feed from the lower ring to the upper four rings reduces how much liquid reaches the reactor bottom by about 70-80%.

Adding a ring-baffle in the bed can also reduce the amount of liquid reaching the reactor bottom by about 40-50%. The baffle prevents wet-agglomerates from reaching the reactor bottom quickly, providing more time for drying. The baffle also creates a zone of high shear between the baffle tips, enhancing agglomerate breakage. Combining baffle addition and feed
redistribution from the lower ring to the upper feed rings reduces the amount of liquid reaching the reactor bottom by about 85-95%.

Keywords

Recirculating Fluidized bed, Fluid Coker, Radioactive Particle Tracking, wet-agglomerates, ring baffle, empirical model, fouling, drying model, breakage model, re-wetting, taper
Summary for Lay Audience

Oil sands, or more specifically the bitumen they contain, require a conversion from heavy oil to valuable lighter hydrocarbons. This thesis aims to improve the performance of Fluid Cokers, a type of fluidized thermal cracking unit used for this type of oil conversion.

A fluidized unit consists of gas injected into a bed of solid particles to impart a liquid-like behaviour to the gas-solid mixture. In a Fluid Coker, heavy oils are mixed with steam and sprayed into a fluidized bed of hot coke (a coal derivative) to be converted into lighter products. In Fluid Cokers, one major challenge is wet-agglomerates formed when coke and bitumen clump together. When these wet-agglomerates reach the unit bottom, the valuable liquid they carry is destroyed, increasing how much raw materials are required per barrel of refined oil. Therefore, optimizing the process would minimize its environmental impact. Wet-agglomerates also stick and accumulate on the inner walls of the unit bottom, significantly limiting the unit operational lifecycle.

A model is developed to predict how much liquid is carried by wet-agglomerates to the bottom of commercial Fluid Cokers. This model bundles several key components of an agglomerate lifecycle: formation, drying due to liquid vaporization, possible re-wetting, and possible destruction from interactions with the bed. Specific focus is given to possible industrially suitable solutions, such as redistributing the heavy oil injection profile or adding a specific insert (“ring-baffle”) inside the bed. A radioactive tracer mimicking an agglomerate is used in a cold gas-solid pilot unit to acquire data for the model. A new mathematical method converting the radiation measurements into tracer coordinates is proposed to improve the accuracy of calculated wet-agglomerate trajectories.

The model shows that redistributing how bitumen is injected into the Fluid Coker, by shifting the injection toward the upper part of the bed, can reduce the amount of liquid reaching the unit bottom. This improvement is explained by a change in the bed hydrodynamics which slows wet-agglomerates travelling down. Adding a “ring-baffle” is also helpful: in addition to slowing down agglomerates travel, it also locally promotes their destruction. Combining both methods is beneficial.
Co-Authorship Statement

Chapter 1: Yohann Cochet prepared the original manuscript, including collecting references and writing the manuscript. Prof. Cedric Briens, Dr. Jennifer MacMillan & Prof. Franco Berruti reviewed and revised the manuscript.

Chapter 2: Yohann Cochet prepared the original manuscript, including conducting experiments, writing any code necessary to analyze the data, processing the data, and writing the manuscript. Prof. Cedric Briens aided for some sections of the code design. Prof. Cedric Briens, Dr. Jennifer MacMillan & Prof. Franco Berruti reviewed and revised the manuscript.

Chapter 3: Yohann Cochet prepared the original manuscript, including collecting references and writing the manuscript. Prof. Cedric Briens, Dr. Jennifer MacMillan & Prof. Franco Berruti reviewed and revised the manuscript.

Chapter 4: Yohann Cochet prepared the original manuscript, including conducting experiments, writing any code necessary to analyze the data, processing the data, and writing the manuscript. Prof. Cedric Briens helped for some sections of the code design. Prof. Cedric Briens, Dr. Jennifer MacMillan & Prof. Franco Berruti reviewed and revised the manuscript.

Chapter 5: Yohann Cochet prepared the original manuscript, including conducting experiments, processing the data, and writing the manuscript. Prof. Cedric Briens, Dr. Jennifer MacMillan & Prof. Franco Berruti reviewed and revised the manuscript.

Chapter 6: Yohann Cochet prepared the original manuscript, including conducting experiments, processing the data, and writing the manuscript. Prof. Cedric Briens, Dr. Jennifer MacMillan & Prof. Franco Berruti reviewed and revised the manuscript.

Chapter 7: Yohann Cochet prepared the original manuscript, including collecting references and writing the manuscript. Prof. Cedric Briens, Dr. Jennifer MacMillan & Prof. Franco Berruti reviewed and revised the manuscript.
Acknowledgments

During the period of my Ph.D. I met many people who helped me make progress through their constructive criticism, and offered helpful assistance and expertise. This thesis would have never been possible without them.

First, I would like to first thanks both my supervisors, Prof. Cedric Briens and Prof. Franco Berruti. Thank you for inviting me to join the Institute for Chemicals and Fuels from Alternative Resources (ICFAR) when I was just finishing my undergraduate degree at the École Nationale Supérieure des Industries Chimiques (ENSIC) in France, for encouraging me to transfer into the doctoral program, and for your support, guidance, and suggestions during the course of this research project.

I want to thank the entire team of engineers at Syncrude Canada Ltd. who work with students and researchers at ICFAR, especially Dr. Jennifer McMillan. Your guidance and insights proved critical, ensuring this research would be as close as possible to industrial needs.

I would like to express my gratitude to Dr. Francisco Javier Sanchez Careaga, the best postdoc that one could hope for. Thank you for your invaluable help and advice during my long hours in the lab and analyzing data.

I would like to thank Syncrude Canada Ltd. and the Natural Science and Engineering Research Council (NSERC) of Canada for their financial support.

I would like to thank Prof. Dominic Pjontek, Prof. Shahzad Barghi, Prof. George Knopf and Dr. Allan Issangya for accepting to serve as examiners for my Ph.D. thesis examination.

I would like to thank all the wonderful people at the University Machine Shop (UMS) at the University of Western Ontario, especially Cody Ruthman, for always going above and beyond to help me build and modify my experimental unit.

I would like to acknowledge the team at the McMaster Nuclear Reactor for their help in creating radioactive tracers for my experiments.
I would like to express my appreciation to my friends and colleagues in the ICFAR and the Department of Chemical and Biochemical Engineering at the University of Western Ontario (Western University). Yuan, Erfan, Ghazaleh, Tim, Cher, Anthony, and so many others: thanks for the help and the laughs. I would also like to express my appreciation to my friends from France. Our discussions and jokes are always a refreshing source of joy, and I cannot wait to see you all again.

I would like to express my gratitude to my parents and brothers for their love, support, and encouragement during all these long years of study.

Finally, I would like to express my deepest gratitude and love to my girlfriend, Rebecca Doyle, for her love, support, and care while we were both working on the arduous path toward a Ph.D. With you, Miki, Stella, Peach and Olive, we form the best fluffle which could ever be.
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\text{s}^{-1})………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………………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= 0.45 kg·s⁻¹) a) Fraction of agglomerates wet & rewetted at stripper for agglomerates formed at Bank i, b) Location of last re-wetting bank for agglomerates formed at Bank i .......... 514

Figure L-7. Fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid & rewetted, with the drying & re-wetting equations only, for various initial agglomerates properties (Short Taper, Baffle @ B, No Flux-Tubes, L_Nozzle = 0 cm, L_jet = 9 cm, u_G ∈ [0.3; 0.9] m·s⁻¹, F_S = 0.45 kg·s⁻¹) a) Fraction of agglomerates wet & rewetted at stripper for agglomerates formed at Bank i, b) Location of last re-wetting bank for agglomerates formed at Bank i ................................................................................................................................. 515

Figure L-8. Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, with the drying and re-wetting equations only (Short Taper, Baffle @ B, No Flux-Tubes, L_Nozzle = 0 cm, L_jet = 9 cm, u_G ∈ [0.3; 0.9] m·s⁻¹, F_S = 0.45 kg·s⁻¹) a) Fraction of liquid initially in agglomerate reaching stripper (from model), b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates) ................................................................................................................................................................................................. 517

Figure L-9. Fraction of liquid carried by agglomerates towards the stripper (for the whole bed), with the drying and re-wetting equations only, for various initial agglomerates properties (Short Taper, Baffle @ B, No Flux-Tubes, L_Nozzle = 0 cm, L_jet = 9 cm, u_G ∈ [0.3; 0.9] m·s⁻¹, F_S = 0.45 kg·s⁻¹) a) Fraction of liquid initially in agglomerate reaching stripper (from model), b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates) ................................................................................................................................................................................................. 518

Figure L-10. Fraction of agglomerates formed at Bank i reaching the stripper with liquid, with the drying and shear destruction equations only (Short Taper, Baffle @ B, No Flux-Tubes, L_Nozzle = 0 cm, L_jet = 9 cm, u_G ∈ [0.3; 0.9] m·s⁻¹, F_S = 0.45 kg·s⁻¹) ................................................................................................................................. 519

Figure L-11. Fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid, with the drying and shear destruction equations only, for various initial agglomerates properties (Short Taper, Baffle @ B, No Flux-Tubes, L_Nozzle = 0 cm, L_jet = 9 cm, u_G ∈ [0.3; 0.9] m·s⁻¹, F_S = 0.45 kg·s⁻¹) ................................................................................................................................................................................................. 521
Figure L-12. Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, with the drying and shear destruction equations only (Short Taper, Baffle @ B, No Flux-Tubes, \( L_{\text{Nozzle}} = 0 \) cm, \( L_{\text{jet}} = 9 \) cm, \( u_G \in [0.3; 0.9] \) m·s\(^{-1}\), \( F_S = 0.45 \) kg·s\(^{-1}\)) a) Fraction of liquid initially in agglomerate reaching stripper (from model), b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)........................................................................................................522

Figure L-13. Fraction of liquid carried by agglomerates towards the stripper (for the whole bed), with the drying and shear destruction equations only, for various initial agglomerates properties (Short Taper, Baffle @ B, No Flux-Tubes, \( L_{\text{Nozzle}} = 0 \) cm, \( L_{\text{jet}} = 9 \) cm, \( u_G \in [0.3; 0.9] \) m·s\(^{-1}\), \( F_S = 0.45 \) kg·s\(^{-1}\)) a) Fraction of liquid initially in agglomerate reaching stripper (from model), b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)..................................................524

Figure L-14. Fraction of agglomerates formed at Bank i reaching the stripper with liquid & rewetted, with the drying, re-wetting and bed shear (Short Taper, Baffle @ B, No Flux-Tubes, \( L_{\text{Nozzle}} = 0 \) cm, \( L_{\text{jet}} = 9 \) cm, \( u_G \in [0.3; 0.9] \) m·s\(^{-1}\), \( F_S = 0.45 \) kg·s\(^{-1}\)) a) Fraction of agglomerates wet & rewetted at stripper for agglomerates formed at Bank i, b) Location of last re-wetting bank for agglomerates formed at Bank i..............................................................................................525

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Figure L-16. Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, with the drying, re-wetting and shear destruction equations (Short Taper, Baffle @ B, No Flux-Tubes, \( L_{\text{Nozzle}} = 0 \) cm, \( L_{\text{jet}} = 9 \) cm, \( u_G \in [0.3; 0.9] \) m·s\(^{-1}\), \( F_S = 0.45 \) kg·s\(^{-1}\)) a) Fraction of liquid initially in agglomerate reaching stripper (from model), b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)........................................................................................................529
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Figure L-19. Repartition of agglomerates positions in-between formation and stripper or destruction, with drying, re-wetting & bed shear equations (Short Taper, Baffle @ B, No Flux-Tubes, \( L_{\text{Nozzle}} = 0 \text{ cm}, L_{\text{jet}} = 9 \text{ cm}, u_G \in [0.3; 0.9] \text{ m} \cdot \text{s}^{-1}, F_S = 0.45 \text{ kg} \cdot \text{s}^{-1}, \rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3}, (L/S)_0 = 0.31, D_{\text{Agg}} = 1\text{ cm}, C_{\text{Rewet}} = 1, d_b = 4.0 \text{ cm}, M = 7 \) ) a) Zones, b) Repartition of agglomerates reaching stripper with liquid, b) Repartition of agglomerates dried or destroyed ................................................................................................................................. 534

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## Nomenclature

**Roman letters:**

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>A, A_i, A'_i,</td>
<td>Constants used in correlation (miscellaneous units)</td>
</tr>
<tr>
<td>B, B_i, B'_i</td>
<td>Constants used in correlation (miscellaneous units)</td>
</tr>
<tr>
<td>AVG_{MovStd}</td>
<td>Numerical criterion assessing the amount of impossible velocities, an average of the moving standard deviations of Cartesian coordinates (with a window size of ( \Delta t_{MOV} )) (m)</td>
</tr>
<tr>
<td>Br_{Agg,0}</td>
<td>Simplified initial agglomerate yield strength (m(^2))</td>
</tr>
<tr>
<td>Cap_{MEAS}</td>
<td>Measured capacitance (F)</td>
</tr>
<tr>
<td>Cap_{REF}</td>
<td>Reference capacitance (F)</td>
</tr>
<tr>
<td>C_i</td>
<td>Number of radioactive counts measured by detector i during a specified time period ( \Delta t_M ) (Counts)</td>
</tr>
<tr>
<td>CC_i</td>
<td>Calculated number of radioactive counts at detector i (Counts)</td>
</tr>
<tr>
<td>CC_{Grid,i}</td>
<td>Calculated number of radioactive counts at detector i, from the position ([x; y; z]) (Counts)</td>
</tr>
<tr>
<td>CC_{Cal1D,i}</td>
<td>Calibrated number of radioactive counts at detector i, from the position ([x; y; z]) (from 1D CARPT calibration) (Counts)</td>
</tr>
<tr>
<td>CCN_i</td>
<td>Normalized calculated number of radioactive counts at detector i</td>
</tr>
<tr>
<td>CCR_i</td>
<td>Calculated radioactive counts rate at detector i (Counts( \cdot s^{-1} ))</td>
</tr>
<tr>
<td>C_{Rewet}</td>
<td>Variation of Liquid-To-Solid ratio before/after re-wetting</td>
</tr>
<tr>
<td>C_{SteelWall,i}</td>
<td>Relative steel wall counts correction</td>
</tr>
<tr>
<td>CR_i</td>
<td>Measured radioactive counts rate by detector i during a specified time period ( \Delta t_M ) (Counts( \cdot s^{-1} ))</td>
</tr>
<tr>
<td>CN_i</td>
<td>Normalized number of radioactive counts at detector i</td>
</tr>
<tr>
<td>C_{p,G}</td>
<td>Liquid bitumen heat capacity (J( \cdot )kg(^{-1})( \cdot )K(^{-1}))</td>
</tr>
<tr>
<td>C_{p,L}</td>
<td>Product vapour heat capacity (J( \cdot )kg(^{-1})( \cdot )K(^{-1}))</td>
</tr>
<tr>
<td>C_{ppm,mol}</td>
<td>Gas concentration (ppm(_{mol}))</td>
</tr>
<tr>
<td>C_{0,MEAS}</td>
<td>Original strength of the tracer used for measurement (Counts)</td>
</tr>
<tr>
<td>C_{0,CAL}</td>
<td>Original strength of the tracer used for calibration (Counts)</td>
</tr>
<tr>
<td>d_b</td>
<td>Bubble diameter (m)</td>
</tr>
<tr>
<td>d_{p,SM}</td>
<td>Sauter-mean diameter of the solids (Fluid Coke) population (m)</td>
</tr>
<tr>
<td>Symbol</td>
<td>Description</td>
</tr>
<tr>
<td>--------</td>
<td>-------------</td>
</tr>
<tr>
<td>$D_{Agg}$</td>
<td>Tracer-Agglomerate &amp; Agglomerate diameter (m)</td>
</tr>
<tr>
<td>$D_{Agg,0}$</td>
<td>Initial agglomerate diameter (m)</td>
</tr>
<tr>
<td>$D_{Agg,Rewet}$</td>
<td>Agglomerate diameter after re-wetting (m)</td>
</tr>
<tr>
<td>$D_{Baffle}$</td>
<td>Baffle diameter (m)</td>
</tr>
<tr>
<td>$D_{Baffle,Bottom}$</td>
<td>Bottom baffle diameter (m)</td>
</tr>
<tr>
<td>$D_{Baffle,Top}$</td>
<td>Top baffle diameter (m)</td>
</tr>
<tr>
<td>$D_{Bed}$</td>
<td>Bed diameter (m)</td>
</tr>
<tr>
<td>$D_{Bed,Bottom}$</td>
<td>Bottom bed diameter (m)</td>
</tr>
<tr>
<td>$D_{Bed,Top}$</td>
<td>Top bed diameter (m)</td>
</tr>
<tr>
<td>$D_{Downcomers}$</td>
<td>Baffle downcomers (flux-tubes) diameter (m)</td>
</tr>
<tr>
<td>$D_{Nozzle}$</td>
<td>Nozzle diameter (m)</td>
</tr>
<tr>
<td>$D_S$</td>
<td>Depth of search volume used for iterative convergence of absorption correction (m)</td>
</tr>
<tr>
<td>$F_G$</td>
<td>Gas mass flowrate (kg·s$^{-1}$)</td>
</tr>
<tr>
<td>$F_S$</td>
<td>Recirculated solids mass flowrate (kg·s$^{-1}$)</td>
</tr>
<tr>
<td>$F_{LS}$</td>
<td>Recirculated solids mass flux (kg·s$^{-1}$·m$^{-2}$)</td>
</tr>
<tr>
<td>$F_{V,G}$</td>
<td>Gas volumetric flow (m$^3$·s$^{-1}$)</td>
</tr>
<tr>
<td>$F_{V,S}$</td>
<td>Solids volumetric flow (m$^3$·s$^{-1}$)</td>
</tr>
<tr>
<td>$F_{pP}$</td>
<td>Particle Froude number</td>
</tr>
<tr>
<td>$g$</td>
<td>Local gravitational field of Earth (m·s$^{-2}$)</td>
</tr>
<tr>
<td>$H_{Baffle}$</td>
<td>Baffle height (m)</td>
</tr>
<tr>
<td>$H_{Bed}$</td>
<td>Bed height (m)</td>
</tr>
<tr>
<td>$H_S$</td>
<td>Height of search volume used for iterative convergence of absorption correction (m)</td>
</tr>
<tr>
<td>$H_{Taper}$</td>
<td>Taper height (m)</td>
</tr>
<tr>
<td>$I$</td>
<td>Radiation received at a distance $L_i$ from the radioactive source (J·s$^{-1}$·m$^{-2}$)</td>
</tr>
<tr>
<td>$I_0$</td>
<td>Radiation received at a distance $L_{4,0}$ from the radioactive source (J·s$^{-1}$·m$^{-2}$)</td>
</tr>
<tr>
<td>$k_{HB}$</td>
<td>Apparent viscosity for the Herschel-Bulkley model (Pa·s)</td>
</tr>
<tr>
<td>$k_S$</td>
<td>Thermal conductivity of coke layers (W·m$^{-1}$·K$^{-1}$)</td>
</tr>
<tr>
<td>Symbol</td>
<td>Description</td>
</tr>
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<td>Symbol</td>
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<td>Agglomerate radius (m)</td>
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<td>$R_{Agg,Rewet}$</td>
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<td>Temperature ($K$)</td>
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<td>$T_{Bed}$</td>
<td>Bed temperature ($K$)</td>
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</table>
Temperature at the reaction front (K)

Bubble velocity (m·s⁻¹)

Superficial gas velocity (m·s⁻¹)

Superficial gas velocity below an agglomerates formation zone (m·s⁻¹)

Bottom superficial gas velocity (m·s⁻¹)

Top superficial gas velocity (m·s⁻¹)

Minimum fluidization velocity (m·s⁻¹)

Chemiresistor input voltage (V)

Output voltage of the capacitance measurement system (V)

Chemiresistor output voltage (V)

Response voltage of the pressure transducer (V)

Chemiresistor amplified & measured output voltage (V)

Input voltage of the capacitance measurement system (V)

Number of γ-rays emitted per disintegration

Agglomerate volume (m³)

Solids (Fluid Coke) volume in agglomerate (m³)

Volume in-bed (m³)

Total volume with an artificially low tracer presence in the bed (m³)

Search volume used for iterative convergence of absorption correction (m³)

Normalized weight determining the contribution of sorted jth node position in the reconstructed tracer position (∑ₖ₌₁ w_j = 1)

Cartesian coordinates of the position of the tracer-agglomerate (m)

Cylindrical coordinates of the position of the tracer-agglomerate (m, rad)

Cartesian coordinates of the position of detector i (m)

Cylindrical coordinates of the position of the detector i (m, rad)

Coke yield

Numerical criterion assessing the proportion of artificially low tracer presence inside the bed

Numerical criterion assessing the proportion of impossible positions
### Greek letters:

<table>
<thead>
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<th>Symbol</th>
<th>Description</th>
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<td>Enthalpy of vaporization &amp; reaction of liquid bitumen at temperature $T$ (J·kg$^{-1}$)</td>
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<td>$\epsilon_{\text{Agg,BeforeRewet}}$</td>
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<td>$\epsilon_S$</td>
<td>Liquid (Bitumen) mass fraction in agglomerate</td>
</tr>
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<td>$\epsilon_{\text{mf}}$</td>
<td>Bed voidage at minimum fluidization</td>
</tr>
</tbody>
</table>
\( \varepsilon_S \)  
Solids (Fluid Coke) mass fraction in agglomerate

\( \zeta \)  
Radiation attenuation coefficient over time

\( \eta(t) \)  
Normalized radial position of the reaction front at time \( t \)

\( \theta_b, \varphi_b \)  
Hilligardt-Werther coefficients for bubble velocity calculations

\( \kappa \)  
Parameter to minimize for CARPT convergence

\( \lambda \)  
CARPT weighting coefficient

\( \mu_L \)  
Liquid viscosity (Pa·s)

\( \xi \)  
Dead-time per recorded pulse (s)

\( \rho_{\text{Agg}} \)  
Tracer-Agglomerate & Agglomerate density (kg·m\(^{-3}\))

\( \rho_{\text{bed}} \)  
Bed density (kg·m\(^{-3}\))

\( \rho_{\text{bulk}} \)  
Bulk agglomerate density (kg·m\(^{-3}\))

\( \rho_G \)  
Gas (Steam & Hydrocarbon vapour mix) density (kg·m\(^{-3}\))

\( \rho_i \)  
Absorbing medium \( j \) density (kg·m\(^{-3}\))

\( \rho_L \)  
Liquid (Bitumen) density (kg·m\(^{-3}\))

\( \rho_{\text{mf}} \)  
Minimum fluidization density (kg·m\(^{-3}\))

\( \rho_S \)  
Solid (Fluid Coke) density (kg·m\(^{-3}\))

\( \sigma_{\text{Rad}} \)  
Standard deviation of radioactive counts measured (Counts)

\( \tau \)  
Mean solid residence time (s)

\( \tau_{\text{agg}}(t) \)  
Agglomerate yield strength (or “yield stress”), at a given time \( t \) (Pa)

\( \tau_{\text{agg}}^*([x, y, z], t) \)  
Critical agglomerate yield strength, at a given position \([x, y, z]\) and a given time \( t \) (Pa)

\( \tau_{\text{COK}} \)  
Mean solid residence time in a commercial Fluid Coker (s)

\( \tau_{\text{EXP}} \)  
Mean solid residence time in the experimental unit (s)

\( \tau_{\text{granule}} \)  
Agglomerate characteristic granule stress (Pa)

\( T \)  
Temporal scale-up criterion

\( \varphi \)  
Peak-to-total (photopeak) ratio

\( \Phi \)  
Angle in-between downcomers (flux-tubes) (rad)

\( \Phi_S \)  
Final sub-incident flux, at \( L_i \)

\( \Phi_0 \)  
Initial incident flux, at \( L_{i,0} \)

\( \Phi_{S,j} \)  
Sub-incident flux, after crossing the medium \( j \)

\( \Phi_{0,j} \)  
Incident flux, before crossing the medium \( j \)
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Strength of the radiation source
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Chapter 1

1 Introduction

The purpose of this thesis is to identify strategies that will help improve the performance of Fluid Cokers, an upgrading process unit used to convert heavy hydrocarbon feedstocks, such as bitumen from Canada’s oils sands, into valuable liquid hydrocarbons. A cold pilot unit is used to obtain information about gas and solids hydrodynamics sensitivity to modifications such as feeding strategy or addition of internals. A model is developed to allow for the transposition of experimental results into predictions for the commercial Fluid Coker. The general research strategy can be broken down as: identification of the most important performance criteria for the target process; identification of potential modifications (applicable to commercial Fluid Cokers) to improve their performance; identification and development of models that will help evaluate the potential modifications; identification of required experimental tools (first to provide the information required by the model, and then to validate the model predictions).

1.1 Bitumen & coking

In the last two decades, the crude oil market has experienced extreme price variations, ranging from historical highs of US$ 167 per barrel in 2008 to relatively low valuation, with variations between US$ 40 and US$ 65 per barrel in the 2016-2019 period (MacroTrends, 2021). To stabilize their rentability, refiners need to maximize stable-income high-value petroleum products output, such as middle distillate, gasoline, and lube oil, over volatile low-value products, such as fuel oil residua-based products (Cross, 2013). This shift towards high-value products is even more crucial for Canadian oil producers: their ongoing difficulties in transporting heavy crude through pipelines or rail to refineries in the U.S. is exacerbated by growing US production. Lacking other market output, they are forced to sell at a significant discount (The Economist, 2018). Simultaneously, environmental concerns have increased, resulting in more rigorous specifications for petroleum products, including fuel oils. Therefore, optimizing the yield of valuable liquid products from various processes and getting more value from residues is of immediate
attention to refiners (Rana et al., 2007). Consequently, since the 1950s, refiners have developed numerous processes to transform “heavy oil” and “extra-heavy oil”, such as bitumen, into conventional oils. Upgrading heavy oil increases the natural resource value, eases its transportation and allows it to be further processed into fuels and lubricants at existing refineries and to be used as feedstock in petrochemical plants (Nikiforuk, 2008; Percy, 2012; Little, 2015).

### 1.1.1 Oil sands & bitumen

Oil sands are composed of a mix of clay material, water, and a form of heavy oil called “bitumen.” In its raw form, bitumen is a dark-coloured, asphalt-like oil and represents about 10 – 12 wt.% of the mined oil sands, depending upon the location. Compared to conventional crude oils, bitumen has higher concentrations of high molecular weight species and heteroatomic species such as nitrogen, sulphur, and metals (Soundararajan, 2001). Thus, it requires upgrading before being usable. The optimization of this upgrading

![Figure 1-1. Athabasca Oil Sands map (Einstein, 2006)](image_url)
process is particularly crucial for Canada, favoured with the third-largest oil reserves in the world, behind Saudi Arabia and Venezuela (The Canadian Association of Petroleum Producers (CAAP), 2019). An estimated 170 billion barrels of bitumen from oil sands are recoverable with current technologies and primarily located in Alberta (The Canadian Association of Petroleum Producers (CAAP), 2019), as shown in Figure 1-1.

In this context, most Canadian refineries have invested in complex configurations that allow them to use cheaper feedstock, such as these heavy oils. Even the recent and sharp production hike of sweet, light oils from shale in the United States is not decisive enough to push companies to refit these existing facilities. The reconfiguration of these numerous facilities adapted to heavy oils would be costly, and consequently, the optimization of bitumen upgrading operations is still crucial for petroleum industries (Eberhart, 2014; Shahandeh and Li, 2016; Shahandeh, 2018).

Bitumen upgrading may consist of hydrogen addition (hydrogenation) or carbon rejection through thermal cracking (Percy, 2012). Thermal cracking is less expensive and is, thus, used on its own or in combination with hydrogenation.

1.1.2 Thermal cracking

Coking is a thermal process for converting heavy hydrocarbons into synthetic crude oil through carbon rejection in the form of coke and with permanent gases as by-products. Several processes have been used to thermally crack bituminous materials (Robinson, 2006; Speight, 2014):

- **Visbreaking (or “Viscosity Breaking”):** Process developed to reduce the viscosity of highly viscous hydrocarbons by introducing them into a furnace to achieve “mild” thermal cracking to meet fuel oil required specifications.
- **Delayed Coking:** Semi-continuous process in which vacuum residues are heated and then introduced into a coking drum, which provides very long residence times that enable more severe thermal cracking.
- **Fluid Coking™ & Flexicoking™:** Continuous processes where vacuum residues are sprayed into a fluidized bed of hot coke particles for thermal cracking into more
valuable products. Fluidized beds have been commonly used in various industries for decades because they provide effective gas and solids mixing and rapid mass and heat transfer. This process decreases the yield of undesirable coke, increasing the yield of liquid products but with lower quality.

1.1.3 Fluid Coker

Fluid Cokers have been developed by Esso Research and Engineering Company (now owned by ExxonMobil) since 1958 (Downing and North, 1958) and are employed by the petroleum industry for thermal conversion of heavy hydrocarbon molecules into distillate products (Matsen, 1985; Kunii and Levenspiel, 1991). In Fluid Cokers, heavy oil such as bitumen is cracked into lighter distillable products (which are later upgraded into synthetic crude oil) through a thermal, non-catalytic conversion process. Fluid Coking™ is preferred for bitumen cracking due to its high flexibility, reliability, continuous production, and low greenhouse gas emissions. (Speight, 2014). Syncrude Canada Ltd., one of the largest producers of synthetic crude oil in Canada, operates the three largest Fluid Cokers in the world (Kamienski et al., 2009). ExxonMobil uses Fluid Coking in many of its refineries. Its Canadian affiliates, Imperial Oil Ltd. and ExxonMobil Canada, also operate Fluid Cokers. More generally, between 30% and 40% of Canada's heavy oil production is processed in Fluid Cokers (Natural Resources Canada - Oil division, 2005; Giove and Sciarrabba, 2019; National Energy Board - Government of Canada, 2019).

The Fluid Coker system is a two-vessel system with a fluidized bed reactor connected to a fluidized bed burner, as shown in Figure 1-2. The process can be described as follows. In the fluidized bed burner, coke particles are partially combusted with air and heated up to 600 to 680 °C. Coke used in the Fluid Coker can be classified as a Group B powder (Geldart, 1973). These “hot” coke particles are then pneumatically transported to the top of the reactor section, in which bitumen feed is sprayed. Coke particles and bitumen come into contact at a temperature ranging from 500 to 550 °C. Bitumen thermally cracks on the surface of the hot coke particles at a temperature ranging from 510 to 550 °C. Since the coking process is endothermic, the hot particles provide the heat required by the thermal cracking reactions and the vaporization of the main cracking products. “Cold” coke
particles will eventually move down the reactor and then back to the burner to start a new cycle (Downing and North, 1958).

The Fluid Coker reactor can be divided into three sections: the reaction section, the scrubber section (located above the reaction section), and the stripper section (located below the reaction section).

- The reaction section is where the bitumen is atomized in the bed through spray nozzles, using high-pressure steam as atomization gas. The goal is to promote a quick and uniform spray of the bitumen over the individual particles for a stable cracking reaction rate. The products consist mainly of hydrocarbon vapours (about 60 wt%), solid coke (about 30 wt%), and permanent gases (about 10 wt%) (ExxonMobil Research and Engineering, 2014).
Product vapours flow to the top of the reactor section and go through the cyclones located at the top of the reactor to remove the entrained coke particles that flow back to the dense bed into the reaction section. The vapours then go through a scrubber section that removes any remaining fine particles and condenses any partially cracked heavy residues. These condensed residues are then combined with fresh feed and recycled back to the reaction section via the feed nozzles. The average superficial velocity of the rising gases and vapours ranges from 0.3 m·s\(^{-1}\) to 1.0 m·s\(^{-1}\), depending on elevation within the fluidized bed and coke size distribution (Pfeiffer et al., 1959; Li et al., 2012a). The variation of the superficial velocity with elevation results from the combination of the distribution of the steam-bitumen injection points along the height of the reactor, the decrease in pressure along the height of the reactor, and the cumulative contribution of the vapours created by cracking. Many Fluid Cokers are tapered to mitigate the increase of the superficial velocity with elevation. After exiting the scrubber section, the product vapours flow through a fractionator and are processed in downstream units.

Coke particles will lose heat in the cracking process and eventually move down the reactor. They will then circulate first to the stripper section and then back to the burner to start a new cycle. The role of the stripper is to enhance the removal of hydrocarbon vapours and liquid moving with the down-flowing fluid coke. Stripping is usually accomplished in a dense, moving, fluidized bed. Steam is injected at the bottom of the stripper, and bubbles rise counter-currently to the down-flowing coke stream (Yang, 2003). Baffles called “shed decks” or “sheds” (multiple rows of chevron-style internals) are placed in the stripping section of the reactor to enhance the interaction between the steam and the coke stream and prevent gas back-mixing (Sanchez Careaga, 2013).

Some attention should be given to the tapered section of the reactor. Conventional straight-sided fluidized beds can be relatively easily modelled with a bed structure analogous to a bubbling liquid, with particle-free voids moving upwards through a uniform emulsion phase (Toomey and Johnstone, 1952). The same approach is often used with tapered beds: they are considered uniformly fluidized and therefore structurally homogenous. This assumption allows for the use of 1D models to describe their behaviour.
(Maruyama et al., 1984). Consequently, the superficial gas velocity decreases with height due to the increasing cross-sectional bed area (Biswal et al., 1982; Shi et al., 1984; DiFelice et al., 1991; Kwant et al., 1995; Deiva Venkatesh et al., 1996; Coulson et al., 1999) and mixing of the homogenous bed is enhanced by favouring the formation of larger bubbles in the upper part of the bed (Peeler and Huang, 1989). However, several studies have demonstrated that tapered fluidized beds are commonly heterogeneous, with different regions in the bed fluidized to different degrees. Since bubbles concentrate in a central region as a result of coalescence, tapered sections can develop a well-defined structure comprising a fully fluidized core region above the gas inlet and a less well-fluidized peripheral region, such as presented in Figure 1-3 (Toyohara and Kawamura, 1990; Toyohara and Kawamura, 1991; Gernon et al., 2007; Gernon et al., 2009; Wormsbecker, Pugsley, and Tanfara, 2009; Wormsbecker, van Ommen, et al., 2009; Gernon and Gilbertson, 2012).

![Figure 1-3. Schematics of the heterogenous structure within a tapered fluidized bed.](image)

(Gernon and Gilbertson, 2012)

Past studies contradict themselves about this structure. For some, the core region is typically uniform in extent and approximately equal in width to the diameter of the bottom of the taper. In addition, when the gas flowrate exceeds a critical value, this structure would
break down, and the entire bed becomes fluidized (Toyohara and Kawamura, 1990; Toyohara and Kawamura, 1991). For others, the fluidized core region grows steadily with increasing gas flowrate and even at very high gas flowrates, thin unfluidized regions are developed along the sloping walls (Gernon et al., 2009; Gernon and Gilbertson, 2012). It should be noted that the high cross-sectional superficial gas velocities used in Fluid Cokers coupled with the solids recirculation between the reactor and the burner make unlikely the persistence of unfluidized region along the sloping walls.

ExxonMobil also developed Flexicoking™, an improvement to the Fluid Coking™ process in which a third major vessel is added to gasify the coke product. In the gasifier, over 95% of the gross product coke is gasified to produce either a low heating-value fuel gas or synthesis gas to make liquid fuels or chemicals. In this manner, the net coke yield is substantially reduced (Saxton, 1972; Furimsky, 2000).

1.2 Losses to burner & fouling in Fluid Coker

The two-vessel design of Fluid Coker, with a recirculation loop between the reactor, where the cracking occurs, and the burner, where the coke is heated, creates a significant risk for vapour and liquid losses of valuable products to the burner. Suppose hydrocarbon vapours, in the emulsion phase, or unreacted liquid, in the emulsion phase or trapped in wet-agglomerates, reach the bottom of the reactor. In that case, they will be lost to the burner, reducing the overall Fluid Coker yield (Sanchez Careaga et al., 2013).

Another challenge in Fluid Coker operation is fouling: a build-up of unwanted material on the surfaces of process equipment (Cui et al., 2006; Sanchez Careaga, 2013). The fouling rate is usually defined as the difference between the rate of deposition and the rate of removal. When fouling occurs in a process, the two most likely scenarios are: an equilibrium between build-up and removal is reached after some time, and the thickness of the fouling layer reaches a maximum, or the deposition rate always outweigh the removal rate, and in the end, a solid barrier will be formed. In Fluid Cokers, the second scenario occurs. Coke accumulates on all inside surfaces of the unit, especially around the area of the cyclones, around the banks of feed injection nozzles located along the height of the reaction section of the bed, and near the stripper (as shown in Figure 1-4), finally leading
to regular and necessary maintenance stops that have a consequential financial cost. In commercial Fluid Cokers, coke deposits may grow to occupy 85% of the original open volume between stripper sheds (Bi et al., 2004). The bed temperature is increased to mitigate fouling, but this can result in over-cracking of bitumen and consequently decreasing the yield of the process (Sanchez Careaga, 2013; Gray, 2015).

![Figure 1-4. Stripper section fouling in a commercial Fluid Coker (modified from Cochet et al., 2019)](image)

1.2.1 Vapour carry-under

Most of the valuable oil vapours trapped in the interstitial voids in-between the downflowing coke particles are recovered in the stripper region of the reactor by countercurrent contact with steam. This stripping occurs before the coke particles are sent to the burner (Cui et al., 2006). The addition of “sheds” type of internals enhances this stripping (Sanchez Careaga, 2013).

1.2.2 Liquid carry-under

In Fluid Cokers, an additional and essential role of the stripper section is to minimize bypassing of unreacted liquid, carried on coke particles within the emulsion phase or
trapped in wet-agglomerates. This unreacted liquid continues to react within the stripper and releases product hydrocarbon vapours which flow up through the stripper sheds (Rose et al., 2005; Cui et al., 2006). As mentioned earlier, this stripping limits the amount of valuable hydrocarbons recirculated to and destroyed in the burner. The drawback is that these vapours can form vapour phase coke that binds coke particles to the surface of the stripper sheds, especially on top shed rows. Extensive fouling alters the sheds shapes, making them thicker and reducing the free space between adjacent sheds. This fouling decreases the cross-sectional area of the coke flow channels and reduces stripping efficiency, and in the end, may cause a premature shutdown of the reactor. (Sanchez Careaga, 2013).

When sprayed into the Fluid Coker, the bitumen feed is sprayed into very fine droplets, significantly increasing the liquid-solid contact area to perform rapid and effective cracking without significant heat and mass transfer limitations (Base et al., 1999). Due to this fast reaction kinetics of cracking when the liquid is at bed temperature (Gray et al., 2004), most of the valuable liquids in the emulsion phase react quickly. Therefore, their contribution to losses to the burner and stripper shed fouling is limited. On the other hand, after the bitumen injection, some of the valuable liquids reaching the stripper are trapped in wet-agglomerates made of coke particles and partially converted heavy oil (Weber et al., 2006; Ali et al., 2010). Due to heat and mass transfer limitations within the agglomerates, liquids trapped cannot reach the bed temperature quickly, limiting their effective reaction rate (Gray et al. 2004; House 2007) significantly.

1.3 Agglomerates

Agglomerates can be defined as aggregates of several individual particles forming an apparent particle size significantly larger than any individual particle size. The agglomeration process in fluidized bed reactors results from several simultaneously occurring elementary steps (Parveen, 2011).

Agglomeration of solids occurs in many fluidization processes. It can be desirable, such as in pharmaceutical and fertilizer industries where agglomeration reduces process problems like dustiness (Weber, 2009). In thermal cracking processes, such as coking, agglomeration
is not desirable since it alters the production yield (agglomerates leave the Coker with a considerable amount of highly valuable un-cracked hydrocarbons, only to be burned in the burner) and contributes significantly to fouling of the reactor internals and surfaces (Rose et al., 2005; Sanchez Careaga et al., 2015).

![Figure 1-5](image.jpg)

**Figure 1-5.** Picture of bitumen-coke agglomerates from commercial Fluid Coker (Photo by Yohann Cochet, 2021)

Typical wet-agglomerates have a diameter ranging from 1 to 20 mm in volume-equivalent diameter (Weber, 2009), with a voidage ranging from 0.30 to 0.50 (Masuda et al., 2006). The densest agglomerates, with a voidage of 0.30, are filled with liquid, giving an agglomerate density of about 1340 kg·m⁻³ for a liquid feedstock density of 1087 kg·m⁻³ (McFarlane, 2007). The lightest agglomerates have a density of around 870 kg·m⁻³, corresponding to a voidage of 0.50, with all their original liquid converted to coke with a coke yield of about 20 wt.% (Sanchez Careaga et al., 2015). Agglomerates recovered from a commercial Fluid Coker are shown in Figure 1-5.
1.3.1 Agglomerate formation

In most granulation processes, the scientific consensus is to model agglomeration as a grouping of the following three steps (Tardos et al., 1997; Iveson et al., 2001; Litster and Ennis, 2004): wetting & nucleation, consolidation & growth/coalescence, attrition & breakage.

The agglomerate formation mechanism in Fluid Cokers is different: it occurs due to the imperfect mixing of dry solids entrained into the low-pressure jet cavities and liquid droplets. Based on experimental research (Ariyapadi, 2004), agglomerate formation occurs near the tip of the jets because all the liquid is not instantaneously vaporized upon injection into the fluidized bed. Indeed, a significant fraction of the feedstock must first crack to smaller vaporizable molecules. It should be noted that some studies have found evidence of agglomerates even when the bed temperature is higher than the boiling point of the injected liquid (Bruhns and Werther, 2005) as liquid injection cools the bed locally.

Figure 1-6. X-ray screenshot of gas-liquid injections in a fluidized bed, with agglomerates formation. Images have been artificially enhanced for clarity. (Ariyapadi et al., 2003)
In Fluid Coker, the injected liquid and its associated atomization gas create a low voidage jet cavity in the bed (Figure 1-6). Dry particles enter the jet cavity from the bed due to suction into the low-pressure zone just downstream of the spray nozzle tip or with the gas bubbles rising into the jet. Entrained particles and high-velocity liquid droplets mix imperfectly, coalescing at the end of the jet cavity to form agglomerates.

The formed wet-agglomerates are periodically removed from the jet cavity by rising gas bubbles detaching from the jet tip and are carried through the fluidized bed, where they may break up into smaller agglomerates or individual wetted particles. The secondary growth of wet-agglomerates can occur by capturing dry particles, especially just after a breakage into smaller wet-agglomerates that have exposed wetter internal surfaces (Weber, 2009). Agglomerate stability depends on liquid content and properties, particle properties, jet and nozzle properties, and local bed hydrodynamics. The agglomerate liquid content is affected by the initial liquid-solids mixing in the jet cavity, subsequent liquid reaction and evaporation, and agglomerate breakup.

1.3.1.1 Liquid & particle properties

Previous studies found that the viscosity of the liquid and contact angle are essential variables in the formation and breakup of agglomerates (McDougall et al., 2005). With a liquid that wets the particles well (low contact angle between the liquid and the solid surface), agglomeration occurs only when the liquid has a high viscosity. On the other end, for liquids that do not wet the particles well (high contact angle), agglomerate formation always occurs. These results were validated experimentally for coke-bitumen agglomerates (Weber, 2009). A moderate increase in liquid viscosity (viscosity increase up to 3 mPa·s above the expected value of commercial feed) makes the agglomerates more stable, which is not desirable for Fluid Cokers. Aminu et al. (2004) noted that even if preheated Fluid Coker feedstocks have a low viscosity of around 2 mPa·s, the liquid viscosity after injection increases rapidly as the reaction occurs, bitumen cracks, volatile products vaporize, and residual liquid polymerizes.

Particle properties also affect agglomerates stability and hence their formation and breakup processes. For instance, agglomerates with poorly wettable solids are less stable
(McDougall et al., 2005; Weber et al., 2006) and do not grow as much through the capture of dry bed particles (Hemati et al., 2003; Saleh and Guigon, 2007). In Fluid Cokers, coke particles are nearly ideally wettable by the injected liquid (Aminu, 2003; Aminu et al., 2004), which promotes stable agglomerates. Besides, if the particle population has a wide size distribution, bridging mechanism by the small particles will promote agglomerate formation (Schaafsma et al., 1998; Weber et al., 2006; Idowu et al., 2018). On the other hand, the presence of larger monosize particles reduces agglomerates formation (Weber et al., 2006). Finally, porous particles lead to less stable agglomerates: their pores absorb some liquid, reducing available binding liquid on the particle surface (Bruhns and Werther, 2005; Ahmadi Motlagh et al., 2019). Therefore, since coke used in Flexicokers is more porous than coke in Fluid Cokers (Jack et al., 1979), Flexicokers form less stable agglomerates.

1.3.1.2 Jets & nozzle properties

In industrial processes such as Fluid Coking™ and Flexicoking™, atomization gas flowrate must be minimized. In the Fluid Coking process, the flowrate of atomization steam must be minimized to reduce energy costs, increase the Fluid Coker processing capacity, and reduce wastewater flows (Farkhondehkavaki, 2012). Therefore, the ratio of the atomization gas mass flowrate to the liquid mass flowrate injected, or “Gas to Liquid Ratio” (GLR), is typically set to range from 0.5 to 2 wt%, with a typical value of 0.86 wt% in Fluid Cokers (Wormsbecker et al., 2016). A typical adjusted value of 2 wt% has been used in most academic studies (Farkhondehkavaki, 2012). This 2 wt% used in academic studies, where air or nitrogen are mostly used as atomization gas, was selected to match the volumetric flow ratio of the commercial nozzles, which use steam as atomization gas. Scaling in this way allows the nozzle pressure, nozzle velocity and the void fractions to match the commercial case.

To achieve a stable operation and adequate atomization at these low GLR values, most early commercial units used the so-called “Terence E. Base” (TEB) nozzle, patented by Base et al. (1999). After a premixer and a nozzle conduit (not represented here), gas-liquid bubbly flow is established, with long ligaments of liquid bitumen flowing within steam. Then, the TEB nozzle uses a succession of restriction/expansion/restriction of diameter to
breaking down these liquid ligaments into progressively smaller droplets through elongation and shear stress while accelerating them. This geometry results in the injection of a fine horizontal spray into the bed, reducing the agglomerate formation rate (Prociw et al., 2018). A TEB nozzle is presented in Figure 1-7.

![Figure 1-7. “Terence E. Base” (TEB) nozzle (modified from Base et al., 1999)"

Nowadays, most commercial units use the more advanced GEN3 feed nozzles, consisting of the TEB nozzle modified with a cloverleaf-shaped disperser (Chan et al., 2015; McMillan et al., 2019). It is presented in Figure 1-8. For instance, these nozzles equip all Syncrude Canada Ltd. Fluid Cokers and most ExxonMobil Fluid Cokers. In addition, the way steam and bitumen are pre-mixed is also optimized (Wormsbecker, Wiens, et al., 2021).

![Figure 1-8. GEN3 nozzle (McMillan et al., 2020)"

For the range of GLR values used in commercial Fluid Coker, the liquid distribution is imperfect, and agglomerates formation occurs. Previous research also investigated a nozzle with a high GLR (up to 50%) and showed no agglomerates were formed when the liquid was injected into the fluidized bed (House, 2007; Farkhondehkavaki, 2012). It should also be noted that significant solids entrainment into the jet cavity is desired as it reduces the liquid concentration of the liquid-solid mixture that accumulates at the tip of the jet cavity (Ariyapadi et al., 2005; Briens et al., 2008).
1.3.1.3 Local bed conditions near jets

Previous research demonstrated that the local bed hydrodynamics affect the original distribution of the injected liquid on the particles (Mohagheghi Dar Ranji, 2015; Li, 2016). It was demonstrated that moving liquid injection to more turbulent regions of the fluidized bed improves liquid distribution. This improvement is due to the gas carried into the jet cavity by the bubbles, which shortens the jet expansion-contraction cycle time, and the solids carried in their wakes decrease the liquid content of the wet solids concentrated at the tip of the jet cavity (Mohagheghi Dar Ranji et al., 2013; Pardo Reyes, 2015). Since entrainment of wet particles into the jet cavity increases the formation of wet-agglomerates, to reduce agglomerate stability, the bed hydrodynamics should also promote a rapid renewal of bed particles surrounding the jet cavity (Mohagheghi Dar Ranji, 2015; Pardo Reyes, 2015).

1.3.1.4 Reducing agglomerates formation in commercial units

Based on the previous discussion, several strategies can reduce the agglomerate formation rate. These strategies are separated into two categories: 1) enhancing nozzle sprays, 2) creating better local bed hydrodynamics near sprays. These categories are not exclusive and should even be complementary: even though better local bed hydrodynamics near sprays can be used to compensate for poor spray nozzle operation, the ideal goal is to achieve a low agglomeration formation rate by combining favourable nozzle operation and favourable bed hydrodynamics (Li, 2016).

1.3.1.4.1 Enhanced spraying

As mentioned earlier, an increase of gas and solids entrainment into the jet cavity reduces the agglomeration formation rate. This entrainment can be promoted by increasing the atomization gas flowrate, which increases the GLR value (Ariyapadi et al., 2005; Briens et al., 2008). Such flowrate increase would come at the expense of increasing the operation cost of the Fluid Coker (energy and wastewater costs) while reducing the Fluid Coker processing capacity. It should be noted that no significant reduction in agglomerate stability reduction was noticed when the GLR value was increased but remained below 1-1.5 wt% at commercial Fluid Coker conditions (Portoghese et al., 2008; Prociw et al., 2018). Since
suitable liquid spraying is more crucial where the fluidization is not as turbulent (i.e., for the lower injection rings), a possible strategy could be to use a different GLR for different injection rings: the GLR could be increased for the lower nozzle rings and reduced for the upper nozzle rings (Wormsbecker et al., 2016).

Liquid distribution can also be improved by pulsing the atomization gas flowrate at a well-controlled frequency of 1 to 5 Hz (Sabouni et al., 2011; Leach et al., 2013). It should be noted that this method requires robust operational control as poorly controlled pulsations lead to an increase in agglomerate formation rate (House et al., 2008).

Finally, since increased viscosity promotes agglomeration formation (McDougall et al., 2005), another strategy to enhance spraying could be a gradual variation of liquid viscosity depending on the injection rings vertical position within a Fluid Coker. The viscosity could be decreased for the lower nozzle rings to reduce the agglomerate formation rate and increased for the upper nozzle rings.

1.3.1.4.2 More favourable local bed hydrodynamics near the spray

An approach discussed in the previous section aimed to move liquid injection locations toward more turbulent regions of the fluidized bed. Instead of redirecting the injections, an alternative approach considered was to make the fluidized bed more turbulent around the current injection locations. This approach showed promising results. For instance, a significant reduction in agglomerate stability can be achieved by surrounding the spray nozzle with small high-velocity gas jets, creating a more turbulent environment (House et al., 2008; Morales, 2013). Li et al. (2020) also found that agglomerate stability can be reduced by increasing the fluidization velocity either locally near the spray zone or in the whole of the bed.

Local bed conditions can also be modified by adding a ring-shaped baffle. The baffle may include downcomers (or “flux-tubes”), as shown in Figure 1-9, to allow downward solids flow and upward gas flow, improving baffle performance (Wyatt Jr. et al., 2011). Internals, such as baffles, are used in the petroleum and chemicals industries to improve fluidization by breaking and re-distributing the bubbles (Chitnis et al., 2013; Issangya et al., 2013;
Kamienski et al., 2013). Previous research showed that baffles could modify local bubble distribution (Jahanmiri, 2017) and solid patterns in a fluidized bed (Sanchez Careaga, 2013). It significantly improved the liquid spray distribution into the fluidized bed above the baffle (Jahanmiri, 2017; Li et al., 2018). Figure 1-9 shows a picture of a commercial ring baffle used in a Fluid Coker. ExxonMobil initially proposed baffles to limit solids mixing between the reaction zone and the stripping zone of a Fluid Coker (Wyatt Jr. et al., 2011).

Figure 1-9. Picture of an industrial Fluid Coker baffle (modified from Kamienski et al., 2013)

1.3.2 Agglomerate growth & erosion

After wet-agglomerates are formed and leave the jet cavity, two opposite mechanisms can happen: growth and erosion. Weber (2009) experimentally identified a critical agglomerate liquid content where the processes of agglomerate growth and erosion were balanced. Agglomerates with liquid contents greater than this value will aggregate dry particles, while agglomerates with less liquid will lose material due to erosion by the surrounding bed. It should be noted that for any agglomerate in growth mode, any dry solids coalescing at the surface will reduce the liquid-to-solids (L/S) ratio in the agglomerate, ultimately
switching the agglomerate from a growth mode to an erosion mode. In Fluid Cokers, the liquid is cracked into lighter vapour hydrocarbon. The reaction and vaporization of liquid trapped within wet-agglomerates are slowed down by heat and mass transfer within the agglomerate (House, 2007). This process reduces the (L/S) ratio in the agglomerate over time, accelerating the switch from growth mode to erosion mode.

1.3.2.1 Drying & bypassing

As mentioned before, in Fluid Cokers, contact of free liquid and wetted particles with hot, dry bed particles is required to provide enough heat for rapid oil reaction and vaporization (Downing and North, 1958). While the cracking reaction is almost instantaneous when the liquid reach the reaction temperature, there is a significant heat transfer limitation in real agglomerates (Sanchez Careaga, 2013). Several models have been proposed in the literature to describe this drying mechanism, ranging from a simple heat transfer to a complex multi-phase combined heat and mass transfer model (Gray et al., 2001; House, 2007; Sanchez Careaga, 2013). Due to a limited effective reaction rate (from the heat and mass transfer limitations within the agglomerates), wet-agglomerates can reach the bottom of the Fluid Coker reactor before they have a chance to dry up (House, 2007). As mentioned earlier, they will then flow to the burner or contribute to stripper shed fouling (Rose et al., 2005; Sanchez Careaga et al., 2015). The larger or wetter agglomerates will take longer to dry (House, 2007; Sanchez Careaga et al., 2015). This short travel time of agglomerates carrying valuable liquid will be referred to as “agglomerate bypassing” in this thesis.

The impact of agglomerates bypassing was validated experimentally in a scaled-down cold flow model of a Fluid Coker: the shortest time for solids to go from the top to the bottom of the reactor bed is about 1 to 2 s (Wormsbecker et al., 2012). This travel time corresponds to 10 to 30 s when scaled up to typical commercial Fluid Coker conditions, a much shorter time than the minimum duration required for complete drying of typical bitumen-coke agglomerates (Ali et al., 2010). Therefore, minimizing agglomerates bypassing is a crucial step to reduce liquid losses in Fluid Coker. It should be noted that if agglomerates bypassing needs to be minimized, solids mixing needs to be maximized to ensure good mixing of wet and hot dry particles to provide efficient cracking. This contradiction is resolved by the wide dispersion of liquid injection locations within the Fluid Coker: by
using up to 50 spray nozzles, efficient solids dispersion is needed on a much smaller scale than the overall bed (Briens and McMillan, 2021).

1.3.2.2 Re-wetting

Previous research used both model predictions and experimental results to show that the agglomerates liquid concentration is higher when solids previously wetted come back into a jet cavity: the agglomerates are rewetted (Elkolaly, 2015; Mohagheghi Dar Ranji, 2015; Pardo Reyes, 2015). Consequently, re-wetting significantly increases the time required for drying and should therefore be limited.

1.3.2.3 Mitigate agglomerates bypassing & re-wetting in commercial units

Based on the previous discussion, several strategies have been investigated to mitigate agglomerates bypassing and agglomerates re-wetting. These strategies are separated into two categories: 1) redistribute liquid distribution between banks, 2) baffle ring(s) addition.

1.3.2.3.1 Redistribute liquid distribution between banks

The time required for an agglomerate to travel from its formation zone (at the tip of the spray jet) to the stripper zone should be reduced in order to reduce agglomerate bypassing. Based on previous studies using thermal drying models, the most critical agglomerates to slow down are the larger and wetter ones, which are more likely to cause fouling and liquid losses. As this will be discussed in this research, this not necessarily the case anymore when possible agglomerate breakages, due to bed shear, are considered. These larger and wetter agglomerates travel faster than bed particles (Sanchez Careaga et al. 2015; Yan et al. 2020) and require more time to dry (Ali, 2002; House, 2007). Previous experiments showed that an increase in the overall gas velocity of the Fluid Coker reduces the residence time of the fastest agglomerates (Ayatollahi, 2016; Hofer et al., 2019; Cochet et al., 2020).

In typical fluidized beds, the Peclet number is about 0.2 (Yang et al., 2019; Zhao et al., 2020), suggesting intense back-mixing, which could lead to a wide distribution of the time required for an agglomerate to travel from its formation zone, at the tip of the spray jet, to the stripper zone. Previous research showed that solids mixing in the Fluid Coker is
primarily induced by gas bubbles, which convey particles in their wake. Solids mixing can be enhanced by increasing the gas velocity (Du et al., 2002) or modifying the gas distribution. Other studies showed that modifying the spray jets modify the gas flow patterns (Li et al., 2012a) and, hence, solids mixing (Sanchez Careaga, 2013). Therefore, a spray jet could also be used to limit agglomerates bypassing. This research conducts experiments to quantify this time distribution. It also investigates how agglomerate properties, bed hydrodynamics, and the location of the spray jets influence this time distribution.

1.3.2.3.2 Baffle ring addition

Adding a baffle just above the stripper section was found to increase the time agglomerates spend in the reactor, above the stripper, reducing stripper shed fouling (Sanchez Careaga et al., 2018).

Preliminary CFD simulations (Xing, 2020) suggest that a ring baffle would enhance solids dispersion above the baffle and reduce it below the baffle. This research conducted experiments to verify if baffle(s) addition within the reactor, between nozzle banks, can also reduce solids bypassing without hindering local solids dispersion.

1.3.3 Agglomerate breakage

Agglomerates formation inside Fluid Cokers cannot be eliminated; therefore, their destruction is significant to operate the unit successfully. Agglomerates can be destroyed through two distinct mechanisms: erosion or fragmentation. Erosion occurs when the agglomerates undergo diffuse microcracking and produce very fine fragments. Fragmentation occurs when the agglomerates experience a brittle fracture and produce much bigger fragments (Salman et al., 2004).

Both destruction mechanisms are always at work, but the agglomerates go from an erosion dominated destruction mode at low gas velocity toward a fragmentation-dominated destruction mode as the fluidization velocity increase. In the range of fluidization velocities used in Fluid Cokers, the fragmentation destruction mode is dominant (Boyle et al., 2005; Weber, 2009).
1.3.3.1 Erosion

Erosion is the agglomerate destruction mechanism that dominates at low fluidization velocities. Shamlou et al. (1990) suggested that it is most likely caused by low energy impacts in the bed core. It was also demonstrated that when erosion is the dominant mechanism of destruction, larger and denser agglomerates are more stable than smaller and lighter ones (Weber et al., 2011).

1.3.3.2 Shear fragmentation

Agglomerate fragmentation occurs when the agglomerate strength becomes lower than the shearing force of the surrounding fluidized bed (Tardos et al., 1997). Stronger agglomerates (i.e., agglomerates with a larger liquid-to-solid (L/S) ratio) can survive at certain fluidized bed conditions where weaker agglomerates would be destroyed (Weber et al., 2011). Besides, the agglomerates drying mentioned in the previous section reduces agglomerates (L/S) ratio over time, making them more brittle and therefore more susceptible to shear fragmentation over time.

Increasing fragmentation of wet-agglomerates reduces Fluid Coker fouling and carries a positive retro-action loop, as the fragments created would dry significantly quicker than the original unbroken agglomerates (Weber, 2009; Sanchez Careaga, 2013). These smaller fragments would also undergo erosion at a much faster rate than the original agglomerates, also designated as “secondary erosion” (Weber, 2009).

1.3.3.3 Promoting agglomerates destruction in commercial units

Based on the previous discussion, several strategies have been investigated to promote agglomerate destruction. These strategies are separated into two categories: 1) change of bed hydrodynamics and agglomerate trajectories, 2) addition of baffle ring(s).

1.3.3.3.1 Change of bed hydrodynamics and agglomerate trajectories

Weber (2009) showed that when the overall Fluid Coker superficial gas velocity increases, fragmentation increases, resulting in smaller fragments and increased secondary erosion of these fragments. Studies have also shown that in a fluidized bed with different levels of
injections, bubbles from the bottom spray jet could be redirected and used to promote agglomerates breakage as they rise in the fluidized bed (Elkolaly, 2015).

Instead of redirecting gas to enhance shear in a specific zone of the Fluid Coker, another strategy could be to redirect solids towards a region with high potential shear fragmentation. As mentioned in previous sections, experiments showed that modifying the spray jets modify the gas flow patterns (Li et al., 2012a) and, hence, solids mixing (Sanchez Careaga, 2013). Therefore, spray jets could redirect agglomerates toward zones within the Fluid Coker where shear fragmentation is likely.

1.3.3.3.2 Baffle ring(s) addition

Baffles can modify local solid patterns in a fluidized bed considerably (Sanchez Careaga, 2013); therefore, they can be used to redirect agglomerate towards a region of high shear fragmentation. For instance, Jahanmiri (2017) found that the turbulence in the region between the baffle tip and the opposite wall increased after a baffle was added in a rectangular cross-section bed. This enhanced turbulence increased the shear on wet-agglomerates.

1.4 Towards a simple experimental-based model for losses in the lower section of Fluid Cokers

This section provides an overview of the modelling and experimental needs to predict valuable product losses in the bottom section of a commercial Fluid Coker. These hydrocarbons losses correspond to the loss of unreacted or partially reacted feedstock due to:

- recirculation to burner vessel,
- agglomerates formation and fouling of internals. This fouling impairs stripping and may cause premature shutdown of the reactor.

These valuable hydrocarbons losses can occur through three main paths: trapped liquid in wet-agglomerates, liquid carried-under in the emulsion, or vapour carried-under.
Therefore, the goal of this research is to measure and combine data obtained in an experimental cold bed unit in two areas:

- Fluidized bed hydrodynamics. This topic will deal with gas/liquid/solid holdups, gas and solids residence time, agglomerate formation rate near lateral injection jets, agglomerate destruction due to shear, and the impact of internals (ring baffles) addition.
- Cracking and vaporization kinetics. This topic will cover liquid and vapour cracking, liquid vaporization, and agglomerate drying.

1.4.1 Previous research

This section discusses previous research conducted to study valuable hydrocarbons losses towards the bottom of commercial Fluid Cokers.

1.4.1.1 Hydrodynamics

Fluid Coker hydrodynamics have been investigated through experimental cold/warm model studies in lab/pilot scale units and by Computational Fluid Dynamics (CFD) simulations of small to commercial scale Fluid Cokers. Cold studies were conducted at room temperature (20 °C), and warm studies were conducted in fluidized beds no hotter than 150-200 °C, considerably colder than commercial Fluid Cokers temperatures of 500-550 °C. CFD models were validated against experimental data (Li et al., 2012b).

Experimental Fluid Coker hydrodynamics studies have to take into account the following points:

- The test unit has to be properly scaled in order to be able to transpose results to commercial units. Scaling between pilot units and commercial units can be performed using dimensionless analysis (Glicksman et al., 1994). This research used the following scaling rules: same solids to gas volumetric flow ratio $F_{V,S}/F_{V,G}$, same recirculated solids flux $F_{L,S}$, same bottom and top superficial gas velocity $u_{G,Bottom}$ & $u_{G,Top}$, and same bed height to bottom reactor diameter ratio.
Consequently, the particle Froude number $Fr_p$ is also similar between the experimental and the commercial unit.

- Systems at ambient & warm conditions have to properly simulate hydrodynamics at commercial Fluid Coker conditions. Theoretically, particle properties should be adjusted to account for differences in gas properties between commercial units and cold simulators (Glicksman et al., 1994). However, for Fluid Cokers, comparing CFD and experimental results showed that axial and radial profiles of the phase holdups and velocities are not very sensitive to changes in gas and particle properties (Song et al., 2006; Xing et al., 2019).

- The effect of spray jets on hydrodynamics has to be considered. In experimental units, spray jets are usually simulated with gas jets (Song et al., 2004; Song et al., 2006). This is an important limitation since it assumes nearly instantaneous vaporization.

CFD simulations of scaled-down Fluid Cokers ($1/38^{th}$ of the commercial scale) were performed with the Ansys Fluent software, using a multi-fluid Eulerian-Eulerian model and Granular Kinetic Theory equations for the solid phase (Li et al., 2012b). The predicted phase holdups agreed with experimental data obtained in scaled-down cold units ($1/19^{th}$ of the commercial scale) (Li et al., 2012b). After this validation, CFD studies were used to overcome experimental limitations. For instance the CFD simulation added the study of the impact of feed vaporization, which could not be studied in the experimental cold units (Li et al., 2012a). CFD (Boyce et al., 2017; Ahmadi Motlagh et al., 2019) or CFD-DEM (Computational Fluid Dynamics - Discrete Element Method) (Girardi et al., 2016) can also be used to model how the local bed hydrodynamics impact breakup of agglomerates of given properties. However, DEM is computationally intensive, and an Eulerian treatment of the solid phase has been found to provide reasonable predictions (Li et al., 2012a; Xing, 2020).

CFD simulations are not a turnkey solution to obtain an in-depth understanding of the physics underpinning fluidized bed behaviours. The closure equations need to be selected to obtain accurate results as a function of the different flow regimes and bed voidage in the fluid. Therefore, experimental measurements and a good understanding of the system are
needed (Lettieri and Mazzei, 2009). The importance of experimental validation is demonstrated by looking at results obtained with 2D CFD simulations, much more common in the literature than 3D CFD simulations, and comparing them with experimental data. It was demonstrated that 2D CFD simulations failed to capture significant qualitative and quantitative trends detected in the experiments (Bakshi et al., 2018). On the other hand, 3D CFD simulations give accurate results but require significantly higher computing resources. Even for 3D CFD, much work still remains to be done to develop correct mathematical expressions for the solid stress tensors and the particle-particle interactions to achieve realistic multi-component systems based on some fundamental science (Lettieri and Mazzei, 2009). This is an ongoing work, which already demonstrated successful predictions of experimental results in a scaled-down version of a Fluid Coker (Song et al., 2006; Lakghomi et al., 2011; Li et al., 2012b; Xing, 2020). Nonetheless, this can only be achieved by comparing the simulation results with an extensive range of experimental studies. For example, the fraction that vaporizes in the spray jet region is not precisely known, impairing the accuracy of Fluid Coker CFD models.

1.4.1.2 Cracking and vaporization kinetics

Due to the challenging operating conditions of commercial Fluid Cokers (> 500 °C), only few studies investigated cracking and vaporization kinetics from experimental fluid bed measurements. Most often cracking experiments were not conducted in a fluidized bed with hydrodynamics representative of commercial Fluid Cokers. Therefore, numerous models have instead been proposed to describe cracking and vaporization kinetics. These models were usually validated against a few selected experimental data points.

The simplest model for cracking and vaporization kinetics considered the coke-bitumen wet-agglomerate as a single solid particle covered by a thin film of bitumen. This model (Gray et al., 2001) correlated the thickness of the reacting thin films with a reduction of the diffusional resistance and, consequently, increased yields of liquids.

As a first approximation to reaction and mass transfer in a thin film, a pseudo-steady-state condition was considered, with products formed in the liquid phase by cracking and evolving by diffusive transport or by bubble formation. This approximation is valid for the
initial cracking reactions so long as the concentration gradients are established rapidly compared to the reaction time. For the initial reaction and transport rates, the concentrations and transport properties are the same as those of the initial reaction mixture. The two main reaction mechanisms are: cracking of the residue to form products and reverse reactions, which trap volatile products in the liquid phase. The product concentration within the reacting liquid phase will fall in two regimes depending on the concentration of the cracked products relative to the concentration required for bubble formation. Considering the liquid phase as two zones, the inner zone (interior of thick films of liquid) will undergo transport by diffusion and gas bubbling while the outer zone (the outermost layer of thick films and to the entire film thickness for the case of thin films), would only undergo transport by diffusion. In this model, there is no driving force for bubble formation.

Cracked products will range from methane, the single most abundant product on a molar basis, to components with boiling points over 650 °C that are stripped out of the liquid phase during the reaction. Nevertheless, as a simplification bubbling process is approximated as pure methane evolution due to cracking.

Another simple model for cracking and vaporization kinetics, the “shrinking core model”, was developed to predict how, under reaction conditions, the liquid inside the agglomerate would be cracked into vapours. The research investigated where the moving agglomerate releases vapours in the fluidized bed to obtain the flow rate of hydrocarbon vapours flowing at any axial location of the bed (Sanchez Careaga, 2013). The model is derived from Crank’s (1975) equations on diffusion through a sphere and has the same mathematical structure as the model for diffusion through ash layers model presented by Levenspiel (1999).

The model was developed using these simplifying assumptions:

- Thermal cracking reactions are instantaneous as soon as the oil reaches the reaction temperature. This is a reasonable assumption for the study of hydrocarbon carry-under and fouling, which are primarily due to large agglomerates where the reaction time does not play a critical role (contrary to small agglomerates).
• The thermal cracking reaction in agglomerates is only limited by conduction heat transfer from the agglomerate outer surface to the reaction front: mass transfer limitations of the vapours to the agglomerate surface are assumed to be negligible.
• The surface temperature of the agglomerate is equal to the bed temperature: any external heat transfer resistance is negligible.
• Stationary conditions: as the reaction front moves, the temperature profile from the outer surface to the reaction front reaches steady-state faster than the reaction front moves.
• The heat capacity of coke is neglected: the heat required to heat the agglomerate solids to the reacting temperature is much smaller than the heat of reaction of the liquid trapped within the agglomerate.
• At the beginning (t = 0), the liquid is uniformly distributed throughout the agglomerate.

Some more complex and complete mass and heat transfer models have also been developed. For instance, House (2007) considers the heat and mass transfer processes in a reacting agglomerate. It combines: a kinetic model to predict the local rate of reaction (adapted from Gray et al., 2004), a model for inter-phase mass transfer to predict the rate of vapour evolution from the liquid phase, a mass and momentum balance performed on the vapour phase to establish the boundary layer condition for the prediction of inter-phase mass transfer, and a heat balance performed to predict the heat transfer through the agglomerate. Finally, a spatial discretization is presented to allow numerical simulation. This model also takes into account that the reacting material is a complex multi-component solution and uses a reaction scheme composed of five lumped species (properties given in Gray et al., 2004): heavy residue (lump 1, 650 °C+), light residue (lump 2, 524-650 °C), gas oil (lump 3, 343-524 °C), distillates (lump 4, 343 °C-), and coke precursors and coke.

1.4.2 Model requirements

This section discusses the requirements to build a simple, scalable model to predict liquid and vapour losses in the bottom section of commercial Fluid Cokers, resulting in stripper fouling or undesired hydrocarbons recirculation to the burner.
First the model objectives will be summarized, then the required simulations steps will be detailed and finally the required measurements will be listed.

1.4.2.1 Model objectives

Combining all the mechanisms and parameters discussed in previous sections, the sum of all unreacted valuable liquid & vapour losses in the bottom section of commercial Fluid Cokers can be separated into three categories, by order of importance:

- Unreacted liquid trapped in wet-agglomerates. The amount of liquid trapped in wet-agglomerates is also controlled by:
  o agglomerates formation rate,
  o agglomerates initial liquid content,
  o agglomerates drying and bypassing,
  o agglomerates destruction in bed
- Unreacted liquid carried by emulsion phase.
- Cracked vapours carried by emulsion phase.

These three categories are inter-connected. A modification of the Fluid Coker, such as a baffle addition for instance, could reduces the impact of one of them while increasing the impact of another category.

Therefore, there is a clear need for a model that integrates these different mechanisms and parameters to predict valuable product losses in commercial Fluid Coker accurately. This model needs to be based on experimental data, scalable, and transposable to commercial units. In addition, it should remain relatively simple and should be a complementary tool to complex CFD simulations. The objectives of this model are to characterize:

1) Wet-agglomerates (composed of trapped liquid & coke)
   a. Quantify residence time & bypassing as a function of bed hydrodynamics (including internals addition) and agglomerate formation location.
   b. Quantify the impact of re-wetting on overall liquid carried-under to stripper and recirculation.
c. Quantify the impact of agglomerate destruction on overall liquid carried-under to stripper and recirculation.

d. Study location of agglomerate destruction, as a function of bed hydrodynamic and agglomerate formation location.

e. Quantify liquid carried-under to stripper and recirculation, as a function of bed hydrodynamics (including internals addition) and agglomerate formation location.

2) Free liquid on individual emulsion particles (“free liquid” behind defined as liquid not trapped in wet-agglomerates)

a. Estimate qualitative carry-under variation as a function of bed hydrodynamics (including internals addition).

3) Vapours in emulsion

a. Estimate qualitative carry-under variation as a function of bed hydrodynamics (including internals addition).

b. Model gas transfer from emulsion to bubble.

1.4.2.2 Required simulations

There are inherent limitations in the experimental unit, that make impossible to directly measure all mechanisms and parameters required to quantify valuable product losses in the bottom section of commercial Fluid Coker. Therefore, these mechanisms and parameters are obtained from an empirical model developed to overcome the experimental unit limitations.

The first experimental limitation is using a gas-solid fluidized bed to study a gas-liquid-solids system. This limitation has several consequences:

- It is not possible to directly measure liquid cracking and agglomerate drying. A cracking and vaporization kinetics model is required. This research selected the shrinking core model based on its simplicity and previous success to estimate liquid losses from agglomerate drying in the stripper region (Sanchez Careaga, 2013).
- It is not possible to directly measure liquid trapping when an agglomerate is formed or re-wetted. Since this is a key component to assess the amount of liquid carried...
under, results from previous research (Pardo Reyes, 2015; Li, 2016) were used to realistically predict initial liquid trapping for a given bed condition.

Another parameter not measurable experimentally in the unit is agglomerate destruction due to shear forces within the bed. This destruction mechanism plays a significant role in mitigating liquid losses to the stripper and burner recirculation. Therefore, a “pass or fail” shear test model, function of agglomerate properties and local bed conditions, is continuously applied to detect agglomerate breakage (Weber, 2009). This also allows for a mapping of liquid release location within the bed.

Finally, the experimental unit is significantly smaller than a commercial Fluid Coker. Therefore, scaling is an important parameter to consider. First, some design choices were selected to ensure reasonable agreements between the experimental unit and commercial Fluid Coker. These were: the ratio of bed height to bottom diameter, the taper angle, the number and positions of nozzles, and the use of Fluid Coke as fluidized solids. Operating conditions were selected to ensure similar hydrodynamics between the experimental and commercial units. Therefore, using the respective solids residence time distribution of each unit to go to dimensionless time, the experimental results can be transposed reasonably to the commercial scale. This is discussed in more detail in Chapter 2 and Appendix I: Spatial and Temporal scaling of the model.

1.4.2.3 Required measurements.

Three main categories of experimental data are required to develop the model: simulated agglomerates tracking data, local bed hydrodynamics, and vapour & liquid carry-under.

The first category of required experimental data is connected to agglomerate tracking. Since the experimental unit of this research is a gas-solid system, it requires the use of simulated agglomerates. For the sake of simplicity, these simulated agglomerates will be referred to as “agglomerates”. The data required includes location and time of agglomerate formation, location and time of agglomerate re-wetting, the time required to travel from agglomerate formation zone to stripper, time since formation for an agglomerate at a given position (required to estimate agglomerate drying), agglomerate repartition within the bed,
and agglomerate flow patterns (required to estimate shear fragmentation). As mentioned in Section 1.3.1, in commercial Fluid Cokers, agglomerates are formed near the lateral injection jet tips. Therefore, in the experimental unit of this research, the simulated agglomerate formation and re-wetting zones are defined as a 3D volume around each of the experimental lateral injection jet tips. More information is provided in Chapter 2.

The second category of required experimental data is connected to global and local bed hydrodynamics. It includes the bubbling and turbulent regime identification and the local voidage (or bubble flux) profile. This data is required to estimate the local shear force at any given position within the bed.

The third category of required experimental data is connected to vapour & liquid carry-under measurements. These data include the minimum amount of liquid injected to detect carry-under (either in vapour or liquid form), differentiation between vapour and liquid carry-under, and steady-state concentration of liquid and vapour carried-under for a given amount of liquid injected. Since the experimental unit of this research is a gas-solid system, these measurements required the use of vaporizable tracer injections in the system. The measurement methods also require to be able to differentiate between vapour and liquid carry-under.

1.4.3 Measurements tools

This section will provide a quick overview of available measurement methods for each category of measurements required for the model. The selected tools are then listed.

1.4.3.1 Global and local bed hydrodynamics

This section presents an overview of measurement methods used to study global and local fluidized bed hydrodynamics. The methods are separated into non-intrusive (global and/or local) methods and intrusive (local) methods.

1.4.3.1.1 Non-intrusive methods

Pressure measurements are often used to characterize fluidized bed hydrodynamics (Yates and Simons, 1994) since they are easy to perform, cheap, and reliable. Measurements
include the transition between bubbling and turbulent regimes (Rhodes, 1996), phases holdups (van Ommen et al., 1999), or slugging detection (Satija and Fan, 1985). One can either use pressure drop measurements over specific sections of the bed to obtain local data or overall bed pressure measurements for more global measurements.

Acoustic emission methods are based on pressure waves emitted from a fluidized bed with a specific frequency range. They can be considered as a shift of pressure measurements to a higher frequency range. Acoustic emissions methods have been used to characterize the fluidization state, such as particle size (Leach et al., 1978), bubbling behaviour (Żukowski, 1999), and fluidization quality (Book et al., 2011). Since other acoustic sources can affect the signal, the measuring equipment and location must be carefully selected to avoid irrelevant sound emissions.

Direct visualization through a column wall is the most straightforward method to determine fluidized bed internal structures. Methods based on digital image analysis (Lim et al., 1990) or laser sheets (Horio and Kuroki, 1994) obtained conclusive results about bubble radial profiles in the upper parts of fluidized beds. Semi-cylindrical beds have also been used to observe jets in fluidized beds and spouted beds, with only minor wall interference (Hatate et al., 1985). However, direct visualization is limited to the outer section of dense three-dimensional fluidized beds, pseudo-two-dimensional beds, or very lean gas-solid beds. Therefore, these methods cannot study internal bed characteristics in the bed interior (Karimipour and Pugsley, 2011). Fluid coke makes the optical method harder as it tends to coat surfaces, increasing the opacity of walls and light emitters.

Another method commonly used for fluidized bed internal flow structure investigation is based on the use of X-rays. For instance, Rowe and Partridge (1966) and Matsuno and Rowe (1970) were able to study bubble shapes, bubble concentration variations, and coalescence mechanisms. However, X-rays images cannot accurately represent multiple, simultaneous overlapping bubbles: they can only visualize 2D projection of 3D objects. Tomography methods need to be implemented to obtain an enhanced observation of multiple bubbles (van Ommen and Mudde, 2008). Using 1D X-ray densitometry imaging, Wormsbecker, Pugsley, van Ommen, et al. (2009) were
able to extract average solids concentration profiles and standard deviation profile of the measured solids along beam path. Similar 1-D analysis can also be performed with γ-rays emission and scintillation detectors located on both sides of the bed.

X-ray tomography has been used extensively to study fluidized bed internal structure. Mudde (2011) used a double X-rays tomographic scanner to measure solid distribution in a fluidized bed to determine bubble size, volume, and velocity for bubbles greater than 2.5 cm. The vertical dimensions of bubbles were obtained from bubble velocity allowing for an estimation of the volume of each bubble detected. According to numerous studies (Verma et al., 2014; Chandrasekera et al., 2015; Maurer et al., 2015), X-ray tomography is a reliable tool for designing bubbling fluidized bed reactors. However, X-ray tomography cannot detect small bubbles and was therefore not selected for this study, in which most of the bed is in the turbulent fluidization regime.

Electrical capacitance tomography (ECT) is another visualization method based on multiple capacitance measurements. It has been successfully used by Li et al. (2016) to determine the averaged bubble rising velocity in a bubbling regime. Chandrasekara et al. (2015) used ECT and measured sizes of gas bubbles moving within a fluidized bed. ECT applications are restricted to small-scale units. In large units, the image resolution decreases significantly. Lancia et al. (1988) also proposed a method to detect the transition from slugging to turbulent regime using capacitance probes. The experimental unit used in this research uses a steel wall to limit erosion. The wall would act as a Faraday cage preventing capacitance measurements from capacitance sensors located around the vessel. Therefore, capacitance measurements would require the insertion of multiple insulated plane electrodes within the vessel, which would be complex and time-consuming. For this reason, the capacitance measurements method was not used.

Radioactive Particle Tracking (RPT) can also be used as an indirect method to acquire information about internal bed structures. For instance, since particles or agglomerates that move up quickly are in bubble wakes, RPT data can be processed to extract local information about bubble flow (Sanchez Careaga, 2013).
This research used pressure measurements to obtain information about the bubbling/turbulent transition, global and local phase holdups along the bed height, and for bed mass monitoring. γ-rays radiation transmission measurements were conducted to verify bubbling/turbulent transition and local phases holdup along bed height. It was also used to measure radial phase holdup.

1.4.3.1.2 Intrusive methods

Optical probes can be viewed as an extension of the direct visualization methods mentioned earlier. They are used for the investigation of local fluidized bed structure in a region of the dense bed. Optical probes were used to measure bubble properties in the core region of a fluidized bed (Mainland and Welty, 1995) and changes in bubble size with gas velocity (Rüdisüli et al., 2012). A critical limit of the method is that when the quantity of bubbles increases, it becomes impossible to accurately determine the bubble size or bubble velocity (Rüdisüli et al., 2012). Optical probes tend to be fragile and not adapted to fluidized beds with high gas velocity. Fluid coke also makes the optical method harder to use as it tends to coat surfaces, increasing the opacity of probes and light emitters over time.

Capacitance probes were introduced by Werther and Molerus (1973) to study the spatial distribution of bubbles to determine fluidization regime transition in gas-fluidized beds. Similar to optical probes, when the quantity of bubbles increases, the method cannot accurately determine the bubble size or bubble velocity (Farag et al., 1997; Karimipour and Pugsley, 2011). Besides, with some high-velocity beds, triboelectric charges induced by the friction of solid particles on the walls and internals become so crucial that electrostatic discharges can destroy capacitance measurement circuits.

The triboelectric effect (or “triboelectric charging”) is a type of electric charge between certain materials after they come in physical contact. Tribo-electric probes use this triboelectric effect to get information about bed hydrodynamics. Tribo-electric probes is a sturdy measurement system that has been used to detect fines in a fluidized bed (McMillan et al., 2011), to measure the moisture content of solids in fluidized beds (Portoghese, 2007), to monitor solid flow in fluidized beds (Lee and Wang, 1995), in injection jets in fluidized beds (Berruti et al., 2009), and in cyclones (Da Silva et al., 2003). In the dense phase, a
triboelectric current is generated by the collisions of particles induced by the bubble motion around the probe (Li et al., 2019). The metal triboprobe has the advantage of being affordable without any maintenance requirement. The intensity of the generated electric current will rely on the fluidization velocity, the bubble size, and the particle size (Ireland, 2010; McMillan et al., 2011; Fotovat et al., 2017). Jahanmiri (2017) used triboprobes to measure bubble gas distribution in silica sand fluidized beds with high fluidization velocities. This work was extended by Li et al. (2019) to investigate how baffle(s) can be used to modify bubble gas distribution. During their research, it was identified that the recorded current is not solely triboelectric but includes current induced by the motion of the charged bed particles. Therefore, the probes were renamed “E-probes”.

This research used E-probes for targeted local validation measurements of the internal bed structure. It should be noted that the solids used in this research, fluid coke, made the use of E-probes more challenging than with silica sand. This is due to the higher electrical conductivity of fluid coke (Helmenstine, 2019), making it less suitable for local measurements of triboelectric and induced currents in fluidized beds. Significant extra shielding and grounding were required to limit electrical noise.

1.4.3.2 Vapour & liquid carry under measurements

Several experimental methods are available to quantify vapour and liquid carry-under. They can be divided into tracer injected and carried-under as gas, and tracer injected as a liquid and carried-under as either liquid or vaporized liquid. The second category requires an additional measurement to differentiate between liquid and vapour signals.

1.4.3.2.1 Gas

An ideal gas tracer could be detected at small concentrations with inexpensive equipment, be safe, and not adsorbed by coke particles. It should be noted that gas carry-under is evaluated at steady-state, and fast response measurement methods are not required. Previous work was conducted to compare different stripper configurations in a recirculated geometrically- and dynamically-scaled half-column of a Fluid Coker (Rose et al., 2005). To study the vapour phase, Helium was injected in the reaction section of the fluidized bed. A gas-sampling probe was positioned in the standpipe below the stripper, and a Thermal
Conductivity Detector (TCD) was used to analyze the sample and quantify the gas carry-under. Another work within the same research group used the same experimental tool to develop a method to calculate the total gas carry-under towards a cold-model Fluid Coker stripper, accounting for non-uniform tracer concentration and gas velocity profiles within the stripper standpipe (Rose et al., 2005).

This research uses MQ-135 gas sensor (Abbas et al., 2020) to detect tracer vapours carried toward the bottom of the fluidized bed and recirculated. This sensor is described in more detail in Chapter 3.

1.4.3.2.2 Liquid & liquid vaporized

Varsol™ is a liquid hydrocarbon with a significant vapour pressure at room temperature (Mohagheghi Dar Ranji, 2015). Varsol™ also does not promote agglomeration when injected into an air-coke fluidized bed at room temperature and atmospheric pressure (McDougall et al., 2005). The initial viscosity of heavy oil used in commercial Fluid Cokers is between 1 and 2 mPa·s. (Aminu, 2003), therefore Varsol™ and its viscosity of 1.2 mPa·s at room temperature provide a good simulation of the industrial conditions. The Varso™ flashpoint is at 43 °C, and the concentration of Varsol™ vapour in air at temperatures below 30 °C is below the lower explosive limit, allowing for safe operation in the experimental unit.

Varsol was used in previous research to perform cold simulations of a Fluid Coker, replacing bitumen liquid feed in a fluid coke-air bed at room temperature (Mohagheghi Dar Ranji, 2015). Capacitance probes were used to measure the evolution of the free liquid in the fluidized bed. A similar method was applied to study the breakage kinetics of wet-agglomerates of coke whereby capacitance probes were used to measure the mass of Varsol™ that was not trapped in wet-agglomerates (Hamidi, 2015).

In another study (Zirgachianzadeh, 2012) electric conductance was used to assess liquid-gas injection (water) into a large gas-solid fluidized bed (silica sand and air) in terms of the quality of distribution of the liquid feed on the bed particles.
Another approach to investigate liquid vaporization with a fluidized bed used an oxygen probe at the cyclone inlet to track water vaporization within a heated silica sand fluidized bed (Silitonga, 2020). Nevertheless, this method has not been tested yet for carry-under measurements. It presents two significant challenges: a risk of water condensation before the measurement zone and low sensor sensitivity for small tracer concentrations.

The MQ-135 gas mentioned above can also work with some liquid tracer. If this liquid tracer vaporizes, either within the bed (vapour recirculation) or after gas addition in the recirculation line (liquid recirculation), it can be detected by the gas sensor. Capacitance probes are then used to differentiate between liquid and vapour tracer carried-under. The method is discussed in more detail in Chapter 3.

1.4.3.3 Agglomerate tracking

Particle tracking tools are a set of non-intrusive methods tracking a tracer-agglomerate motion within a fluidized bed. They are used to: map agglomerates repartition within the bed, map agglomerates flow patterns (and, by extrapolation, gas flow patterns), obtain the time required for the agglomerate to travel from their formation zone to the stripper, and study agglomerate ejection at the bed surface. The most common particle tracking method is Radioactive Particle Tracking (RPT), based on a radioactive tracer-agglomerate and radiation sensors disposed around the bed (Khanna et al., 2008; Tamadondar et al., 2012; Sanchez Careaga, 2013; Ayatollahi, 2016; Tebianian et al., 2016; Bhatti, 2017). The RPT method was initially developed using a single tracer, but recent developments allowed multiple tracers (Rasouli et al., 2015). Another method using a similar principle is Magnetic Particle Tracking (MPT), where the radioactive tracer is replaced by a magnet (Sette et al., 2015). The main advantage of this method is its ability to detect the orientation of a non-spherical tracer. Another variation on the RPT method is Positron Emission Particle Tracking (PEPT). This method replaces small individual radiation detectors used for RPT with large radiation-sensitive panels (Parker and Fan, 2008). This modification allows for more accurate position determination, producing an exact flow pattern map, but at the cost of a significant limitation on the maximum bed volume which can be investigated. An innovative approach using Radio Frequency Identification (RFID) technology was developed to detect individual agglomerates' fragmentation inside a
fluidized bed (Parveen, 2011). This allowed the study of how much time a single agglomerate would survive at a given bed condition and would indicate where this agglomerate would break in the bed. A slightly different approach consists of coating actual bed solids particles with a coloured dye. These coated particles are injected into the recirculated fluidized bed, and the recirculated solids are sampled continuously. The samples are later analyzed to obtain the proportion of dyed solids passing through the recirculation line over time (Zhao et al., 2020). This measurement allows for solids Residence Time Distribution (RTD) measurements. Alternatively, the solids tracer can be made by applying a fluorescent coating. This coating can then be activated with lights located at strategic positions. Finally, light intensity measurements are made with optical probes located at positions of interest, either in the recirculation line or within the bed (Tayebi, 1998).

This research uses Radioactive Particle Tracking (RPT) to track agglomerates trajectories. This method was chosen because it was previously successfully used to investigate agglomerate tracking in a fluidized bed with (Sanchez Careaga, 2013) and without recirculation (Ayatollahi, 2016; Bhatti, 2017). A more detailed description of the benefits and limits of the method is shown in Section 1.5. Validation experiments were conducted with coloured dye.

1.5 Radioactive Particle Tracking (RPT)

Radioactive Particle Tracking (RPT) is the primary measurement tool in this research. Its principles, benefits, and limits are summarized in this section.

Used in fluidized beds, the Radioactive Particle Tracking (RPT) technique can be described as using emitted gamma rays (γ-rays) radiations, from one or more radioactive tracer-particle, to obtain the distances between the radioactive source and the detector from an array of sensors located externally to the bed. The obtained matrixes of distances are then solved to calculate the source particle position inside the bed (Chaouki et al., 1997).

1.5.1 RPT principle

A complete RPT system requires, at least:
• A single radioactive tracer emitting $\gamma$-rays.
• Three or more $\gamma$-rays sensitive detectors. It usually is scintillation detectors.
• One computer per detector, connected in a network to a server. They will receive, process, synchronize, and store the data from each detector.

For safety reasons, radioactive gold ($^{198}\text{Au}$) is one of the preferred isotopes for RPT experiments as it decays very quickly (half-life of 2.697 days) into a stable isotope of Mercury ($^{198}\text{Hg}$) (Moreira et al., 2010). However, this short half-life can also limit the scope of available experimental measurements. Therefore, to perform measurements over a long time, a radioactive source with a significant half-life was needed. For this reason, radioactive Scandium ($^{46}\text{Sc}$) isotope with a half-life of 83.79 days was chosen. Another advantage of $^{46}\text{Sc}$ is that it disintegrates into a stable isotope of Titanium ($^{46}\text{Ti}$), mitigating safety and environmental risks.

To study agglomerate behaviour, the tracer particle must be prepared to ensure similar aerodynamic properties, such as shape, size, and density. However, regardless of the method of tracer preparation and the materials used, any tracer will not be of the same material as the fluidized medium. Therefore, the tracer will not behave exactly like the studied medium, and a compromise between size and density might be necessary, especially for smaller tracers.

1.5.2 RPT benefits

The main advantage of the RPT method is its non-intrusive nature. Data can be obtained without disrupting the gas-solid flow inside the bed. Thus, it is particularly well-suited to study the 3D motion of liquids or solids in multiphase flows (Chaouki et al., 1997). It also allows simultaneous measurements of the 3D position and velocity components over a vast measuring volume with a wide velocity range. Furthermore, RPT relies on small, independent, and easily movable detectors allowing measurements in a unit of arbitrary size and shape. Finally, the RPT method can provide time-averaged full-scale 3D concentration and velocities profiles for the whole system.
1.5.3 RPT limits

Obtaining time-averaged 3D concentration and velocities profiles requires accumulating a significant amount of representative flow information by accumulating data acquired over a few hours to a couple of days (Chaouki et al., 1997).

It has been observed in past research (Sanchez Careaga, 2013) that the number of counts detected by any scintillation detector for a given statistical tracer position located inside the vessel can vary significantly. Since radioactive decay is described by a Poisson distribution (Leo, 1994), the variance $\sigma$ of the measured counts rate $CR_i$ (the number of counts measured $C_i$ by detector $i$ during a specified time period $\Delta t_M$) will increase with decreasing source strength and with decreasing sampling time (Holbert, 2002; Windows-Yule, 2020). The decreasing source strength can correspond to: an increase of source-detector distance (inverse square law), an increase of medium absorption in-between the source and the tracer (inverse exponential decrease with increasing medium density), and reduced radioactive source strength (inverse exponential decay). For these reasons, a random error in the location of the tracer-agglomerate is always expected. In Chapter 3, this analysis is expanded and presents the variation of the fraction of error in the measured value as a function of the source strength and sampling time. Besides, varying bed density (i.e., phase holdup) also adds random noise in the recorded signal. In this research, several processing steps have been developed, tested, and deployed to limit the impact of these errors on the calculated positions.

1.5.4 Coordinates rendition methods

Several rendition techniques exist to determine the coordinates of the tracer inside the reactor at a given time using the combined radiation signals obtained from the set of scintillation detectors. The two most common methods are:

- Computer Automated Radioactive Particle Tracking (CARPT)
- Monte Carlo simulation
1.5.4.1 CARPT

The CARPT method was initially developed by Lin et al. (1985). It is based on the principle that the number of $\gamma$-rays counted by a detector within a given time interval is unequivocally proportional to the distance between the tracer-agglomerate position and a virtual center in this detector surface.

A calibration curve relating $\gamma$-rays counts to distance is established for each detector. Next, the calibration data obtained are processed to produce a curve fit of the $\gamma$-rays raw data. Various fitting methods are used to describe the different domains of distance versus $\gamma$-rays counts relationships. The overall bed is described with a general polynomial or power-law 1D fitting, while short range-distance is described with fitting equations accounting for the tracer-detector solid angle. Finally, more accurate results will require tedious in situ calibration, at the operating condition of interest, to properly account for the varying bed density (Chaouki et al., 1997).

Once the tracer-detector distance $L_i$ is obtained for each detector, by defining an arbitrary reference frame, the CARPT formula can be written as shown in Equation (1.1):

$$L_i^2 = (x - x_i)^2 + (y - y_i)^2 + (z - z_i)^2$$  \hspace{1cm} (1.1)

where:

- $L_i$ is the distance between the tracer and the detector $i$, obtained through a numerical fitting of calibration data.
- $x_i$, $y_i$, $z_i$ are the coordinates of the position of detector $i$.
- $x$, $y$, $z$ are the coordinates of the position of the tracer.

Therefore, the system of equations requires only three detectors to be solved. The availability of more distance $r_i$, by using more than three scintillation detectors, results in data redundancy for the location determination. This planned redundancy is used to improve the location accuracy by using a weighted least-square method (Lin et al., 1985). The main advantages and disadvantages of the CARPT method are presented in Table 1-1.
Table 1-1. Main advantages and disadvantages of the CARPT rendition method

<table>
<thead>
<tr>
<th>Advantages</th>
<th>Disadvantages</th>
</tr>
</thead>
<tbody>
<tr>
<td>- Short processing time. The time required to process a recorded set of counts into a position is down to 3.0 $10^{-2}$ ms per event recorded (using a Lenovo T410, with an Intel® Core™ i5-520M CPU @ 2.40GHz). An event is defined as a time interval of counts recorded. - The simplicity of the mathematics involved.</td>
<td>- Requires a substantial in-situ calibration. - Ignore the effect of the angle between the tracer and a horizontal plane through the virtual center of the detector. - If there is significant radiation attenuation due to the bed particles, the method accuracy decreases significantly (the solids holdup, and therefore the signal attenuation between the source and the detector will vary with time) (Devanathan, 1991).</td>
</tr>
</tbody>
</table>

Sanchez Carrega (2013) used the CARPT method to study agglomerate behaviour in a fluid coke bed, but with a experimental system with a smaller diameter than the current study. He obtained a frequency map of occurrence (spatial distribution), the local residence time in different bed sections, and velocity plot arrow (agglomerate trajectories). Bhatti (2017) developed, in a 2D system, a modified version of the CARPT method based on a set of empirical equations obtained by multilinear regression of calibration experiments. Good results were obtained on the statistical tracer location, but it required a very time-consuming calibration phase and gave poor dynamic results, i.e. tracer velocities and patterns.

1.5.4.2 Monte Carlo simulations

To limit extensive in-situ calibration, École Polytechnique de Montréal research group developed a phenomenological approach to account for geometry and radiation attenuation effects in RPT (Chaouki et al., 1997). They created a Monte-Carlo-based rendition technique. This method requires the construction, for each detector, of one map of
theoretical counts detected as a function of any possible coordinates of the particle. These theoretical counts recorded by a detector, during a sampling time interval $\Delta t_M$, from a point radioactive source of strength $A$ placed at a location $[x, y, z]$, are obtained using Equation (1.2). Using this map, the system can obtain the actual tracer position from the measured detectors counts. Since a certain fraction of the $\gamma$-rays is absorbed by the fluidized material and by the vessel walls, a new map generation is needed whenever the medium density to be studied changes (Chaouki et al., 1997).

$$CC_i(x, y, z) = \frac{\Delta t_M \cdot v \cdot \chi \cdot \varphi \cdot \varepsilon(L_i)}{1 + \xi \cdot \chi \cdot \varphi \cdot \varepsilon(L_i)}$$

(1.2)

where:

- $CC_i$: calculated number of counts at detector $i$, located at $[x_i, y_i, z_i]$.
- $\Delta t_M$: sampling time (s).
- $v$: number of $\gamma$-rays emitted per disintegration.
- $\varphi$: peak-to-total (photopeak) ratio.
- $\varepsilon(L_i)$: total efficiency (the probability that $\gamma$-rays will emerge from the reactor without scattering and interact with the detector), a function of the distance source-detector $i$ $L_i$.
- $\xi$: dead-time per recorded pulse (s).
- $\chi$: strength of the radiation source.
- $[x, y, z]$: position of the radiation source.

A least-squares approach is then used to search for the best position within the generated map by minimizing, for each detector $i$, the difference between the measured counts $C_i$ to the calculated counts $CC_i(x, y, z)$ obtained by Monte Carlo simulation at each of the grid points.
Table 1-2. Main advantages and disadvantages of the Monte Carlo rendition method

<table>
<thead>
<tr>
<th>Advantages</th>
<th>Disadvantages</th>
</tr>
</thead>
<tbody>
<tr>
<td>- Requires less in-situ calibration.</td>
<td>- Require an accurate description of the physical properties of the measurement volume and its surroundings.</td>
</tr>
<tr>
<td>- Considers the angle at which the $\gamma$-rays enter the sensor.</td>
<td>- Mathematics far more complex</td>
</tr>
<tr>
<td>- Less sensitive to radiation attenuation due to the bed particles (a set of reconstruction map can be used to correct for bed density variations)</td>
<td>- Significantly longer computer time required to obtain the position. This time is up to 100 ms per event, 2380 times longer than CARPT (using a Lenovo T410, with an Intel® Core™ i5-520M CPU @ 2.40GHz).</td>
</tr>
</tbody>
</table>

To speed up the process, the grid search is limited to neighbouring nodes. Finally, to provide adequate resolution for particle tracking, the search is continued in the selected grid point neighbourhood, investigating inter-node positions. No Monte-Carlo calculations are necessary to obtain the theoretical counts $CC_i(x, y, z)$ at the inter-node locations investigated during the neighbourhood search; with these slight variations, $CC_i(x, y, z)$ can be extrapolated from the $CC_i$ value at the selected grid position. The main advantages and disadvantages of the Monte-Carlo method are presented in Table 1-2.

1.5.4.3 Neural network

Recently Artificial neural networks (ANN) have been introduced to reconstruct tracer positions from measured counts rate (Dam et al., 2019; Dam et al., 2021). ANN are mathematical models inspired by the human neuronal function. Their main feature is to be able, if an appropriate set of data was selected, to recognize and identify patterns (Dam et al., 2019). This implies that after a training (or “learning”) phase, where the ANN learn to recognize specific patterns in a controlled dataset, the ANN is used to identify these patterns in new uncharacterized datasets. A limitation of the method is related to the great number of measurements required to perform the ANN training. Alternatively, equations similar to ones used with the Monte-Carlo-based methods can be used to describe very
accurately the physical properties of the measurement volume and its surroundings and simulate the required dataset for the ANN training (Dam et al., 2019). The accuracy achievable with well-trained ANN is very high, with values of relative error on calculated positions around 0.03-0.08 (Dam et al., 2021).

1.6 Thesis Objectives, Overview of Experimental Approach and Thesis Outline

The main objective of the research described in this thesis is to develop a model to quantify the valuable hydrocarbons losses towards the bottom section (the reactor stripper section and recirculated to the burner) of a Fluid Coker and to identify potential mitigating strategies. The valuable hydrocarbons losses correspond to the loss of unreacted or partially reacted feedstock. These losses can be carried by wet-agglomerates (containing trapped liquid), liquid carry-under in the emulsion, or vapour carry-under in the emulsion. The stripper exit losses are recirculated and combusted in the burner. Vapours flowing in the stripper (mostly released by wet-agglomerates) foul the stripper internals. This fouling impairs stripping and may cause premature shutdown of the reactor. This research investigates how bed hydrodynamics can be globally or locally modified to mitigate this valuable product loss. The impact on bed hydrodynamics of the location and injection pattern used for the feed nozzles and baffle(s) addition is given specific attention. The study focuses on the behaviour of simulated agglomerates, from a formation zone to the stripping section of a cold flow scaled Fluid Coker and liquid & gas carry-under. To achieve this objective, model agglomerates were manufactured with a radioactive source to be tracked when they move in the fluidized bed unit. A model is developed to integrate the results and transpose them to industrial units. The liquid & vapour carry-under is investigated using a liquid tracer which vaporized within the bed before being carried-under and recirculated in either liquid or vapour form. A gas sensor located in the recirculation line is used to detect this carry-under.

The main objective of this research was divided into the following sub-objectives:

Chapter 2: Development and testing of a model to quantify liquid losses to the stripper due to agglomerates. This model includes the whole life cycle of agglomerates, from formation
to drying, to potential re-wetting, to shear fragmentation. The model is tested against initial agglomerates diameter and liquid content. A scaling analysis is also conducted.

Chapter 3: Design and construction of a lab-scale tapered cylindrical cold flow recirculating fluidized bed. The experimental unit allows to adjust the nozzle penetration of its lateral injectors and incorporates replaceable ring baffle internal(s). Design and calibration of a scintillation detector network to track the motion of a single radioactive tracer-agglomerate, simulating a wet-agglomerate of chosen properties, placed in a recirculating flow of fluid coke particles. Design of an injection and measurement system for vapour & liquid carry-under. Development of validation method for RPT (using dyed coke).

Chapter 4: Radioactive Particle Tracking (RPT). It is essential in the analysis of the amount of liquid carried to the stripper by wet agglomerates. Due to the significant amount of data recorded, this research used the CARPT method over the more precise but much slower Monte Carlo simulation method. Steps were taken to improve the quality of the data. This included: signal denoising to reduce the random variation of the radioactive signal and absorption correction, to reduce the impact of varying bed absorption.

Chapter 5: The effect of the feed injection patterns is studied. The number of lateral injection banks used, the repartition of gas injected between banks, and the nozzle tip position is varied. The following parameters are compared: transition from the bubbling to turbulent fluidization regime, agglomerates flow pattern, formation-to-stripper time, agglomerates formation/destruction, and liquid carryunder.

Chapter 6: The effect of baffle(s) is studied. The number of the baffle(s), their position within the bed, and the use of downcomers are investigated. The following parameters are compared: transition from the bubbling to turbulent fluidization regime, agglomerates flow pattern, formation-to-stripper time, agglomerates formation/destruction, and liquid carryunder.
1.7 References


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Chapter 2

2 Predicting the liquid carryunder to the stripper section of a Fluid Coker™

The model developed in this chapter aims to predict how much valuable liquid is carried by wet-agglomerates to the bottom of a commercial Fluid Coker (also referred to as the "stripper zone"). This transport of liquid to the stripper zone is referred to as "liquid losses". The model integrates different mechanisms impacting wet-agglomerates in a commercial Fluid Coker to give an overall picture of the lifespan of an agglomerate, from formation to (potential) destruction. Each mechanism is discussed extensively, and the equations and assumptions used are presented.

Typical results produced by the model are then presented and discussed. The impact of the different mechanisms is investigated and the sensitivity of the results to important model assumptions is discussed.

2.1 Overview of the proposed model

In the model, the whole reactor of a commercial Fluid Coker is simplified and divided into three zones: the production zones (corresponding to the lateral jets where a mix of steam and bitumen is sprayed in the bed), the stripper zone (located at the bottom of the Fluid Coker), and the rest of the bed (referred to as the "general zone").

The model includes the main mechanisms involved in the lifespan of a wet-agglomerate:

- Formation: wet-agglomerates are formed and released at the tip of the Fluid Cokers gas-liquid spray jets (Weber, 2009). This location of agglomerates release is referred to as the “production zone”.
- Thermal drying: occurs in the general zone. Heavy hydrocarbons trapped in wet-agglomerates are cracked and vaporized due to heat, making the wet-agglomerate dryer (Sanchez Careaga, 2013).
- Breakage: happening in the general zone. Due to interactions with bed hydrodynamics, the agglomerate can break down into smaller fragments and eventually disappear (Weber, 2009).
- Re-wetting: takes places in a production zone. A wet-agglomerate can re-enter a production zone and re-capture liquid (Pardo Reyes, 2015).

**Figure 2-1.** Overview of the liquid losses model developed.

Liquid losses occur when the wet-agglomerates reach the stripper zone: any liquid still carried by the wet-agglomerate will either be contributing to the stripper fouling or being recirculated to the Fluid Coker burner. Therefore, this model does not consider possible recovery in the stripper region, as identified in previous study (Sanchez Careaga, 2013),
and uses the conservative assumption that any liquid reaching the bottom of a Fluid Coker
is lost. Similarly, no distinction is established between the liquid contributing to fouling or
the liquid recirculated to the burner: the model only calculates the overall liquid loses.

A flowchart presenting an overall view of the model is shown in Figure 2-1.

The modular structure of this model will facilitate future upgrades and adaptations. For
example, in this work, required data about agglomerates trajectories were acquired from
experiments in a pilot gas-solids circulating fluidized bed. Nonetheless, the model is
agnostic in this regard. CFD simulations could have been used instead to simulate
agglomerate trajectories, and the model would work in the same way.

Another important component of the model is related to scaling. The objective is to ensure
that the predictions are obtained for commercial Fluid Cokers and not for the experimental
unit or CFD mesh, which can both be scaled-down, and where agglomerates data are
acquired. For example, using data from a 1/10th scaled-down experimental unit and a 1/5th
scaled-down experimental unit should allow for similar model predictions for the
commercial scale.

2.2 Model scaling

As mentioned above, data on agglomerates trajectories were acquired from experiments in
a pilot gas-solids circulating fluidized bed (presented in Chapter 3). This section discusses
how the scaling rules were applied to scale up the relevant agglomerates experimental data
to the commercial Fluid Coker scale.
SCALING

A flowchart presenting an overall view of the scaling component of the model is presented in Figure 2-2.

2.2.1 Agglomerate location coordinates & production zone location coordinates

The pilot gas-solids circulating fluidized bed was designed in such a way as to achieve a constant fixed bed height to bottom diameter ratio (see Chapter 3).

\[
\frac{H_{\text{EXP}}}{R_{\text{EXP}}} = \frac{H_{\text{COK}}}{R_{\text{COK}}} = F \tag{2.1}
\]

Let us define the dimensionless radial location (\(r/R\)). When this dimensionless position is the same between the experimental unit and the commercial Fluid Coker, then:

\[
R_{\text{COK}} = \delta_1 \cdot R_{\text{EXP}} \\
\tau_{\text{COK}} = \delta_1 \cdot \tau_{\text{EXP}} \tag{2.2}
\]

Similarly, let us define the dimensionless height location (\(h/R\)). When this dimensionless position is the same between the experimental unit and the commercial Fluid Coker, then:
\[ H_{\text{COK}} = \delta_2 \cdot H_{\text{EXP}} \]
\[ h_{\text{COK}} = \delta_2 \cdot h_{\text{EXP}} \]

(2.3)

In addition, the angular position \( \theta \) is conserved when scaling up the positions.

Therefore, when scaling from the experimental position \([x, y, z]\) in the pilot-scale experimental unit towards the position \([x', y', z']\) in the commercial-scale Fluid Coker, we have:

\[
\begin{bmatrix}
  x' \\
  y' \\
  z'
\end{bmatrix} =
\begin{bmatrix}
  \delta_1 \\
  \delta_1 \\
  \delta_2
\end{bmatrix}
\cdot
\begin{bmatrix}
  x \\
  y \\
  z
\end{bmatrix}
= \delta \cdot
\begin{bmatrix}
  x \\
  y \\
  z
\end{bmatrix}
\]

(2.4)

This position scaling can be used for the agglomerates and the boundary of the production zone between a pilot scale and a commercial scale vessel. More details are given in Appendix I: Spatial and Temporal scaling of the model.

### 2.2.2 Agglomerate residence time & drying time

The pilot gas-solids circulating fluidized bed was operated in such a way as to achieve the same top and bottom cross-sectional superficial gas velocity as the commercial Fluid Coker (see Chapter 3). It was also operated with a similar solids recirculation flux.

There are two different times, the "time in pilot" \( t \), measured in the pilot-scale unit, and "time in Fluid Coker" \( t' \), required in the model. Similarly, there are two different average solids residence times in the reactor, the "average solids residence time in pilot" \( \tau_{\text{EXP}} \), measured in the pilot-scale unit (as discussed later, this was found to be around 2-2.5 minutes on average), and "average solids residence time in Fluid Coker" \( \tau_{\text{COK}} \), required in the model (this was reported to be approximately 15 minutes according to Song et al. (2006) and Sanchez Careaga (2013)).

By assuming that the downward velocity of the agglomerate is the same in the commercial (full scale) and experimental (scaled down) units, then:

\[
t' = \frac{\tau_{\text{COK}}}{\tau_{\text{EXP}}} \cdot t
\]

(2.5)

Therefore, the dimensionless time in the reactor \((t/\tau)\) is the same between a pilot scale and a commercial scale vessel.
Wormsbecker, Wiens, et al. (2021) demonstrated the effectiveness of similar scaling strategies, with an excellent agreement between the dimensionless break-through time of a cold flow model and a commercial unit.

2.2.3 Wet-agglomerate thermodynamics (drying & strength)

All the thermodynamic equations are supposed to simulate wet-agglomerates behaviour in a commercial Fluid Coker. Therefore, the values used are the ones relevant to a commercial Fluid Coker.

These values include the initial properties of agglomerates: \(D_{\text{Agg},0}\), \(\rho_{\text{Agg},0}\) and \((L/S)_0\). The values used are detailed in Table 2-1. These values also include the agglomerates and reaction properties relevant to drying or agglomerate strength equations \((T, \rho_L, \mu_L, \ldots)\). The values used and the references from which they were obtained are detailed in Table 2-1 and Table 2-2.

2.2.4 Bed hydrodynamics (bed shear & agglomerate breakage)

As mentioned earlier, the pilot gas-solids circulating fluidized bed was operated with the same top and bottom cross-sectional superficial gas velocity as the commercial Fluid Coker (see Chapter 3). It was also operated with a similar solids recirculation flux.

Therefore, there are similar cross-sectional superficial gas velocities \(u_G\) (ranging from \(u_{G,\text{Bottom}}\) to \(u_{G,\text{TOP}}\), identical for the two scales) between the experimental unit and the commercial Fluid Coker. This assumption is discussed in more detail in Chapter 3.

To scale up thermodynamic gas properties \((\rho_G, \mu_G)\) the model uses the relevant values from the commercial Fluid Coker. The values used are detailed in Table 2-2.

To scale up bubbles properties, the model also uses the relevant values from a commercial Fluid Coker. Bubble properties selection, especially the bubble size, is discussed in detail in Section 2.5.1.2.1 and Section 2.9.1.4.

The bed voidage profile (both axial and radial) is assumed to be similar between the experimental and commercial Fluid Coker. The voidage profile is measured in the
experimental unit and scaled up using the same coefficient as for the agglomerate position scaling. This assumption is discussed in more detail in Appendix K: Measurement of the bed voidage profile in the experimental unit.

2.3 Wet-agglomerates formation & re-wetting

**Figure 2-3.** Flowchart of wetting and re-wetting process

As discussed in Chapter 1 – Section 1.3.1, agglomerate formation corresponds to the aggregation of solids and unreacted liquid occurring due to imperfect mixing in the jet cavity near the lateral injection jets.
The model used past studies from the literature to characterize the wet-agglomerates formed. This characterization includes the wet-agglomerates initial properties, such as their diameter, density, and wetness. This characterization also includes information about how much of the total fraction of liquid injected is captured by agglomerates.

The model also considered re-wetting and used past studies to estimate when and how it affects wet-agglomerates.

A flowchart presenting an overall view of the wet-agglomerate production component of the model is shown in Figure 2-3. This flowchart includes the original production of an agglomerate ("wetting") and possible re-wetting.

2.3.1 Production zones

Previous studies showed that wet-agglomerates are formed near the tip of the low voidage jet cavity created by the injected liquid and its associated atomization gas in the bed (Ariyapadi et al., 2003). This is discussed in more detail in Chapter 1 – Section 1.3.1. Therefore, the wet-agglomerate production zones were defined as volumes centred around the tip of the lateral injection jets, as shown in Figure 2-4.

Three size parameters were considered to define the production zones in the model: the height $\Delta z$, the radial depth $\Delta r$, and the angular range $\Delta \theta$ (with $\Delta z = \Delta r$). Preliminary tests were conducted to assess the sensitivity of the model results relative to these parameters. Using the results of these tests and the dimension of the jets in the experimental unit, these model parameters were set to $\Delta z = \Delta r = 4$ cm & $\Delta \theta = 22.5^\circ$.

The position of the jet tip was calculated from selected literature correlations for horizontal jets in gas-solids fluidized beds (Merry, 1971; Benjelloun et al., 1995). These correlations were validated by previous experimental work in a similar system (Ariyapadi et al., 2004). In addition, experiments were conducted with E-probes (see Chapter 3), which validated the results from the correlations.
An important model assumption is related to the jets scale up from the experimental unit to the commercial Fluid Coker scale. It was assumed that the scaling rule defined in Section 2.2.1. would apply. When compared, jet penetrations normalized by the bed radius, are similar between the experimental pilot scale and the commercial Fluid Coker scale. Detailed values are presented in see Chapter 3 - Table 3-1. The same is true for nozzle penetrations. Detailed values are also presented in see Chapter 3 - Table 3-1.

One important model limitation is the non-consideration of potential secondary growing of wet-agglomerates. This mechanism consists of wet-agglomerates capturing independent solids particles after their formation (Weber, 2009). It was assumed that the solids captured represent only a minor fraction of the wet-agglomerate mass and the mechanism quickly disappears as the wet-agglomerate dries. Therefore, it was neglected.

### 2.3.2 Wet-agglomerate formation (initial wetting)

This section discusses wet-agglomerate characteristic properties when they form. It also discusses how to calculate the fraction of liquid injected trapped into a given wet-agglomerate.
2.3.2.1 Initial wet-agglomerate parameters

The initial properties of a wet-agglomerate can be described using three parameters: the agglomerate diameter $D_{\text{Agg}}$, the agglomerate density $\rho_{\text{Agg}}$, and the initial Liquid-to-Solid ratio $(L/S)_0$. These three values were obtained using previous measurements from the literature (Weber, 2009; Sanchez Careaga, 2013; Bhatti, 2017). The values used and their references are presented in Table 2-1.

An assumption is made in the model that agglomerates are generated with a single set of initial agglomerate properties: all agglomerates are generated with the same initial diameter $D_{\text{Agg},0}$, the same initial density $\rho_{\text{Agg},0}$, and the same initial Liquid-to-Solid ratio $(L/S)_0$. In reality, a commercial Fluid Coker contains a population of wet-agglomerates, that are continuously generated and destroyed, with variable properties.

To overcome this limitation and obtain a more accurate prediction for the overall wet-agglomerate population, the model needs to process the same dataset of wet-agglomerate trajectories with several different agglomerate initial parameters. Future development of the model could include integrating the simulation of a population of agglomerates, but this requires more experimental work to realistically assess the distribution of such population in a commercial Fluid Coker.

The model uses agglomerate trajectories starting from any production zone to calculate possible liquid losses. In this thesis, such trajectories were obtained using a tracer-agglomerate in the experimental unit. Therefore, agglomerate production is detected when the tracer-agglomerate is detected as going through a production zone. If the tracer-agglomerate is not already registered in the model as carrying liquid before entering the production zone, it becomes "wetted" and is now considered as a wet-agglomerate in the model. If it is already "wetted", it is re-wetted (see Section 2.3.3).

Table 2-1. Parameters used for agglomerate wetting

<table>
<thead>
<tr>
<th>Parameters</th>
<th>Values</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td>Agglomerate initial diameter</td>
<td>$[5.0 \times 10^{-3}; 1.5 \times 10^{-1}]$</td>
<td>Masuda (2006), Weber (2009), Sanchez Careaga (2013), Pardo</td>
</tr>
</tbody>
</table>
2.3.2.2 Fraction of liquid injected initially trapped in wet-agglomerates

Only a fraction of the liquid hydrocarbons injected into Fluid Coker is trapped into wet-agglomerates. In addition, wet-agglomerates of different sizes or formed at different locations will capture a variable fraction of the liquid injected (Li, 2016).

Figure 2-5 gives a simplified flowchart of liquid transfer after injection into a Fluid Coker.

The rest of the model developed in this chapter focuses on the fraction of the liquid initially trapped in wet-agglomerates that reaches the stripper zone. Therefore, this step focusing on the fraction of liquid injected initially trapped in the wet-agglomerates is critical. Without this correction, the model could predict that some wet-agglomerates carry significant liquid to the stripper while their impact is limited because, in the first place, they do not capture much of the liquid injected. Therefore, this step is essential to accurately estimate the fraction of liquid injected carried by wet-agglomerates to the stripper zone.

<table>
<thead>
<tr>
<th>Agglomerate initial density $\rho_{\text{Agg},0}$ (kg·m$^{-3}$)</th>
<th>[870; 1340]</th>
<th>Reyes (2015), Bhatti (2017), Joness (2019)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Agglomerate initial Liquid-to-Solid ratio $(L/S)_0$</td>
<td>[0.10; 0.62]</td>
<td></td>
</tr>
</tbody>
</table>
An equation was derived from previous experimental measurements to calculate this fraction of liquid injected initially trapped in the wet-agglomerates (Li, 2016). This equation varies with the initial agglomerate size $D_{\text{Agg},0}$ and the cross-sectional superficial gas velocity below the injection production zone considered $u_{G,\text{BELOW}}$. It should be noted that the equation does not consider the radial variation of the gas velocity, a potential path of improvement for future work.

Figure 2-6 presents the fraction of liquid injected trapped in agglomerate as a function of the agglomerate size $D_{\text{Agg}}$ & the gas velocity below the production zone $u_{G,\text{BELOW}}$ as calculated with the equation used. It clearly shows that the agglomerate diameter is the most critical of the two parameters. Smaller agglomerates capture much more of the liquid injected. This result agrees with experimental and industrial observations as much more small agglomerates are formed overall. The gas velocity below the formation zone is not as critical but still significant: the lower the gas flow under the injection jet, the more it captures liquid. This result is again in agreement with the literature, as lower gas velocities promote the imperfect mixing causing the wet-agglomerate to form at the tip of the jet cavity.
Figure 2-6. Fraction of liquid injected trapped in an agglomerate as a function of the agglomerate size $D_{Agg}$ and gas velocity below the production zone $u_{G,BELOW}$ (adapted from Li, 2016)

The dataset acquired by Li (2016) did not extend above an initial diameter $D_{Agg} = 1$ cm. Therefore, values for larger agglomerates will be extrapolated.

2.3.3 Wet-agglomerate re-wetting

Previous studies have shown that wet-agglomerates, already formed and released in the bed, can move back into a production zone (Pardo Reyes, 2015). This phenomenon is referred to as the re-wetting of wet-agglomerates and modifies wet-agglomerates properties.

It should be noted that re-wetting can occur in the same production zone as the one that originally produced the wet-agglomerate or in totally different production zones, near another spray jet of the Fluid Coker.

As mentioned earlier, the model tracks when the tracer-agglomerate goes through a production zone. If the tracer-agglomerate is already registered in the model as carrying liquid, it is re-wetted. If not, it becomes "wetted" (see Section 2.3.2).
Experimental measurements presented in Figure 2-7 (Pardo Reyes, 2015) indicate that re-wetting modifies the liquid-to-solid ratio of agglomerates by a coefficient of re-wetting $C_{Rewet}$ such as presented in Equation (2.6).

$$1 \leq \frac{(L/S)_{Rewet}}{(L/S)_0} \leq 2 \leftrightarrow 1 \leq C_{Rewet} \leq 2$$  \hfill (2.6)

![Graph showing typical agglomerate size in commercial Fluid Coker](image)

**Figure 2-7.** Evolution of liquid concentration in wet-agglomerate after re-wetting compared to the original liquid concentration (adapted from Pardo Reyes, 2015)

(Air/Silica sand fluidized bed, Gum Arabic/Water/Silica Sand agglomerates, $u_G = 0.3 \text{ m·s}^{-1}$, $T_{Bed} = [111; 116] \, ^\circC$)

The following assumptions are made in the model regarding the other wet-agglomerates properties. First, the wet-agglomerate diameter $D_{Agg}$ is kept constant before/after the re-wetting. Second, the wet-agglomerate voidage $\varepsilon_{Agg}$ is kept constant before/after the re-wetting.

$$D_{Agg, rewet} = D_{Agg,0} \quad \& \quad \varepsilon_{Agg, rewet} = \varepsilon_{Agg, BeforeRewet}$$  \hfill (2.7)

Due to the liquid-to-solid ratio ($L/S$) modification, the wet-agglomerate density $\rho_{Agg}$ is modified and re-calculated after a re-wetting.
Another model limit is the use of a single value of the coefficient of re-wetting $C_{\text{Rewet}}$ for the whole bed. The value used for most cases is $C_{\text{Rewet}} = 1$, which is a safe side assumption. A more detailed analysis of this topic is conducted in Section 2.9.

### 2.3.4 Scaling summary

This section is a quick summary of the procedures for the agglomerate formation and re-wetting scaling between the experimental unit and the commercial Fluid Coker.

First, for the wet-agglomerates and the production zones boundaries position, the scaling procedure presented in Section 2.2.1 is used. This implies a similar dimensionless radial location ($r/R$) and a similar dimensionless radius to bed height ($r/h$) between the experimental unit and the commercial Fluid Coker.

Second, the initial properties of wet-agglomerates ($D_{\text{Agg},0}$, $\rho_{\text{Agg},0}$, $(L/S)_0$) are defined for a realistic representation of a commercial Fluid Coker, following recommendations of Section 2.2.3.

### 2.4 Wet-agglomerate drying kinetics

This section presents the component of the model calculating the wet-agglomerate drying. The drying refers to the progressive cracking and vaporization of hydrocarbons trapped in wet-agglomerates. Therefore, the mass fraction of liquid trapped in the wet-agglomerate decreases with time until reaching zero or until the agglomerate exits the reactor.

One of the most important parameters controlling the level of drying of a wet-agglomerate is the time since its formation or latest re-wetting. Therefore, the first step was to obtain the “production-to-stripper” (short for “production zone to stripper zone”) time distributions of wet-agglomerates. This corresponds to the distribution of the time required for wet-agglomerates to move from a production zone to the stripper zone.

Using the data from these production-to-stripper time distributions (and possible re-wetting), a thermal drying model used with proper scaling can calculate the equivalent liquid carried by a given wet-agglomerate in a commercial Fluid Coker.
A flowchart presenting an overall view of the wet-agglomerate drying component of the model is presented in Figure 2-8.

![Flowchart of the drying process in the model](image)

**Figure 2-8.** Flowchart of the drying process in the model

### 2.4.1 Agglomerate transport in the bed

Figure 2-9 shows the agglomerate travel from a production zone to the stripper zone, including possible re-wetting.

As mentioned in previous sections, the model requires information about agglomerates trajectories in order to compute their wetting, drying, potential re-wetting and potential breakage. In this thesis, such trajectories were obtained using a tracer-agglomerate in the experimental unit. Similar results could be obtained with other means, such as CFD simulations.

In this research, agglomerate production and re-wetting are implemented when the tracer-agglomerate is detected as going through a production zone. The differentiation between production and re-wetting is established based on the wetness of the agglomerate before entering the production zone, as registered in the model (see Section 2.3.2 & 2.3.3).
Similarly, the tracer-agglomerate is used to determine the travel time from each production zone to the stripper zone. With enough trajectories, one can calculate the production-to-stripper time distributions of wet-agglomerates for given bed hydrodynamics and given tracer-agglomerate properties.
2.4.1.1 Production-to-stripper time distributions

Typical production-to-stripper calculated time distributions are presented in this section. Each lateral injection jet is associated with its own production zone and, therefore, its own production-to-stripper distribution. To reduce random fluctuations due to a limited number of data points in the distribution, data for all the jets of the same injection bank, i.e. ring of jets located at the same height, are combined. Since each jet of a given bank is fed with the same feedstock and injected within similar bed regions, they should have a production-to-stripper time distribution similar to the other jets and similar to the overall production-to-stripper time distribution of the bank.

For the overall bed, the contribution of each bank is equal to the ratio of the mass flowrate of gas injected in the bank divided by the total mass flowrate injected in all active banks. Their respective weight will be the same for a configuration where the same mass flowrate is injected in each bank. Chapter 3 provides a more detailed discussion about how the gas-liquid feed jets of the commercial Fluid cokers were substituted and simulated by pure gas jets injections in the experimental unit.

Figure 2-10 presents a typical production-to-stripper time distribution. The distributions are made dimensionless using the average solids residence time in the bed, which is by definition much longer than the production-to-stripper times. This difference explains why almost all the agglomerates go through the bed in less than $t/t_{\text{EXP}} = 1$. It matches previous experimental and industrial observations about agglomerates reaching the stripper significantly faster than the average particle population (Wormsbecker, Wiens, et al., 2021).

It is also clear that the lower the bank, the shorter the residence time. This result is expected as the distance between a production zone and the stripper zone is an important parameter controlling how fast wet-agglomerates can reach the stripper zone. It should be noted that the figure uses a logarithmic scale for the horizontal x-axis.
**Figure 2-10.** Production-to-stripper time distributions

(5 Banks distribution as shown in Figure 2-9, $D_{\text{Tracer}} = 0.9$ cm, $\rho_{\text{Tracer}} = 1100$ kg·m$^{-3}$, $F_S = 0.44$ kg·s$^{-1}$, $m_S = 59.5$ kg)

Using similar methods to the one used to obtain the production-to-stripper time distributions, the model can obtain the "age" of a wet-agglomerate, defined as the time since production or last re-wetting, at any time.

### 2.4.2 Thermal drying model

Once the "age" of wet-agglomerates is known at any time, a thermal model can be used to calculate the wet-agglomerate liquid content at any time.

While the cracking reaction is almost instantaneous when the liquid reach the reaction temperature, there is a significant heat transfer limitation in real agglomerates (Sanchez Careaga, 2013). Internal heat transfer through the agglomerate is limiting, while external heat transfer resistance (between the bed and the wet-agglomerate) is not significant. Therefore, it is a reasonable assumption to consider that heat transfer is not affected by the local bed hydrodynamics.
2.4.2.1 Thermal models review

A review of the thermal models available in the literature for wet-agglomerate drying was presented in Chapter 1 – Section 1.3.2.1.

The model chosen for this research is the shrinking core model proposed by Sanchez Careaga (2013). This model was selected as it was a good balance between a realistic approach and a simplified model. Indeed, the shrinking core model considers only one phase of liquid to crack and assumes that the drying is only controlled by the "age" of the wet-agglomerate for given Fluid Coker conditions. It will be discussed in more detail in the following section.

2.4.2.2 Shrinking core model

This model was derived from Crank's (1975) equations on thermal diffusion through a sphere and presents a similar mathematical structure as the model for diffusion through ash layers presented by Levenspiel (1999).

The main assumptions of the models are:

- Single component model (all the cuts of liquid hydrocarbon behaves as a homogenous liquid)
- The liquid-to-solid ratio (L/S) decreases between $t = 0$ and $t = t_C$. With $t_C$ being the time required for full drying (when $(L/S) = 0$)
- The agglomerate diameter $D_{Agg}$ is assumed constant over time ($D_{Agg} = D_{Agg,0}$).

Figure 2-11 presents a simplified diagram of the shrinking core Model.
Figure 2-11. Simplified diagram of the shrinking core Model
The governing equation of the model, describing the decrease in the liquid-to-solid ratio over time is defined by Equation (2.8):

\[
\left(\frac{L/S}{(L/S)}_0\right) = \left(\frac{m_L(t)}{m_S(t)}\right) = \left(\frac{m_L(t)}{m_L,0}\right) \cdot \frac{m_S,0}{m_S(t)}
\]  \hspace{1cm} (2.8)

where:

- \((L/S)\): concentration of liquid in the solid (Liquid to dry Solid ratio)
- \((L/S)_0\): initial concentration of liquid in the solid (initial Liquid to dry Solid ratio)
- \(m_L(t)\) mass of liquid in the agglomerate at time \(t\) (kg)
- \(m_L,0\): mass of liquid in the agglomerate when formed at \(t = 0\) (kg)
- \(m_S(t)\) mass of solid in the agglomerate at time \(t\) (kg)
- \(m_S,0\): mass of solid in the agglomerate when formed at \(t = 0\) (kg)
- \(t\): time since agglomerate formation (s)

The decrease of the mass of liquid in the agglomerate over time is described by Equation (2.9):

\[
\frac{m_L(t)}{m_L,0} = \eta(t)^3
\]  \hspace{1cm} (2.9)

where:

- \(\eta\): normalized radial position of the reaction front. It is the ratio of the radial position of the reaction front \(r_{Reac}(t)\) to the agglomerate diameter \(D_{Agg}\)

The normalized radial position of the reaction front is calculated by solving Equation (2.10):

\[
(1 - \eta(t)).(1 + \eta(t) - 2.\eta(t)^2) = t/t_c
\]  \hspace{1cm} (2.10)

with:

- \(t_c\): time required for complete conversion (s). It is the total time required for a complete conversion of all the liquid trapped within the agglomerate, defined by Equation (2.11).
\[ t_c = \frac{R_{Agg,0}^2 (L/S)_0}{6. \gamma} = \frac{R_{Agg}^2 (L/S)_0}{6. \gamma} \]  

(2.11)

The parameter \( \gamma \) is constant with agglomerate size \( D_{Agg}(t) \) and the agglomerate liquid-to-solid ratio \( (L/S)(t) \). It is defined by Equation (2.12).

\[ \gamma = \frac{k_S (T_{bed} - T_{reaction})}{\rho_{S,bulk,0} \Delta H} \]  

(2.12)

with:

- \( \rho_{S,bulk,0} \): initial dry bulk density of coke in the agglomerate (kg·m\(^{-3}\)), as defined by Equation (2.13).

\[ \rho_{S,bulk,0} = \frac{m_{S,0}}{V_{Agg,0}} = \frac{m_{S,0}}{(4/3) \pi R_{Agg,0}^3} \]  

(2.13)

- \( k_S \): thermal conductivity of coke layers (W·m\(^{-1}\)·K\(^{-1}\)). It is used to consider the thermal transfer through the outer coke layers further away from the agglomerate center than the reaction front.

- \( T_{bed} \): bed temperature (K)

- \( T_{reaction} \): temperature at the reaction front (where the thermal cracking occurs, see Figure (2.11)) (K).

- \( \Delta H \): enthalpy changes when the liquid bitumen reacts (J·kg\(^{-1}\))

It should be noted that the temperature at the reaction front \( T_{reaction} \) is much lower than the bed temperature \( T_{bed} \) because of the significant heat that is required to convert liquid bitumen into cracked vapour: the liquid is at a much lower temperature than the average bed temperature (Gray et al., 2004; House et al., 2004; House, 2007).

The enthalpy \( \Delta H \) change when the liquid bitumen reacts and vaporizes at the reaction front temperature is computed using the assumption that the heat reaching the reaction front goes to react with the liquid only. In addition, due to the lack of better value, the enthalpy of vaporization and reaction used was measured at \( T_{bed} \) and not \( T_{reaction} \) (Sanchez Careaga, 2013). The enthalpy changes when the liquid bitumen reacts. \( \Delta H \) is expressed by Equation (2.14).
\[ \Delta H = \Delta H_{liq}(T_{bed}) - C_{p,L} \cdot (T_{bed} - T_{reaction}) + C_{p,G} \cdot (T_{bed} - T_{reaction}) \]  \hspace{1cm} (2.14) \\

with:

- \( \Delta H_{liq}(T) \): enthalpy of vaporization & reaction of liquid bitumen at temperature \( T \) (J·kg\(^{-1}\))
- \( C_{p,L} \): liquid bitumen heat capacity (J·kg\(^{-1}\)·K\(^{-1}\))
- \( C_{p,G} \): product vapour heat capacity (J·kg\(^{-1}\)·K\(^{-1}\))

The mass of solids in the agglomerate \( m_s \) can be calculated using Equation (2.15):

\[ m_s(t) = m_{s,0} + m_{s,new} = m_{s,0} + y_S \left( 1 - \frac{m_L(t)}{m_{L,0}} \right) \cdot m_{L,0} \]  \hspace{1cm} (2.15) \\

with:

- \( m_s(t) \): mass of solids in agglomerates (kg) at time \( t \)
- \( y_S \), the coke yield, is defined as the "mass ratio of the new coke formed over the initial mass of bitumen in the agglomerate" (Ali et al., 2010)

The initial mass of solids in the agglomerate \( m_{s,0} \), just after its production, can be calculated using Equation (2.16):

\[ m_s(t = 0) = m_{s,0} = m_{Agg,0} \frac{\rho_{Agg,0} \cdot \left[ (4/3) \cdot \pi \cdot R_{Agg,0}^3 \right]}{1 + \frac{m_{L,0}}{m_{s,0}}} \]  \hspace{1cm} (2.16) \\

with:

- \( m_{s,0} \): initial mass of solids in agglomerates (kg)
- \( \rho_{Agg,0} \): initial agglomerate density (kg·m\(^{-3}\))

Similarly, the initial mass of liquid in the agglomerate \( m_{L,0} \), just after its production, can be calculated using Equation (2.17):

\[ m_L(t = 0) = m_{L,0} = m_{Agg,0} - m_{s,0} = \rho_{Agg,0} \cdot \left[ (4/3) \cdot \pi \cdot R_{Agg,0}^3 \right] \cdot \left( 1 - \frac{1}{1 + \left( L/S \right)_0} \right) \]  \hspace{1cm} (2.17)
with:

- \( m_{s,0} \): initial mass of solids in agglomerates (kg)
- \( \rho_{\text{Agg},0} \): initial agglomerate density (kg·m\(^{-3}\))

The agglomerate voidage (dry basis) \( \varepsilon_{\text{Agg}} \) range is physically bounded between 0 (no agglomerate) and 1 (pure coke). Based on Masuda et al. (2006), this range can even be restricted between 0.3 and 0.5. Using the Shrinking core model assumption of constant agglomerate radius \( D_{\text{Agg}}(t) = D_{\text{Agg},0} \), the agglomerate voidage (dry basis) \( \varepsilon_{\text{Agg}} \) is calculated using Equation (2.18).

\[
\varepsilon_{\text{Agg}}(t) = \frac{V_{\text{Void}}(t)}{V_{\text{Agg}}(t)} = 1 - \frac{V_{S}(t)}{V_{\text{Agg}}(t)} = 1 - \frac{(m_{S}(t)/\rho_{S})}{(4/3) \pi R_{\text{Agg}}(t)^3} \\
= 1 - \frac{(m_{S}(t)/\rho_{S})}{(4/3) \pi R_{\text{Agg},0}^3}
\]

with:

- \( V_{\text{Void}}(t) \): volume of void (not solids) in agglomerate at time \( t \)
- \( V_{S}(t) \): volume of solids in agglomerate at time \( t \)

Using again the Shrinking Core model assumption of constant agglomerate radius \( D_{\text{Agg}}(t) = D_{\text{Agg},0} \), the agglomerate density \( \rho_{\text{Agg}} \) is calculated using Equation (2.19).

\[
\rho_{\text{Agg}}(t) = \frac{m_{S}(t) + m_{L}(t)}{V_{\text{Agg}}(t)} = m_{S}(t) \cdot \left[1 + \left( \frac{L}{S} \right)(t) \right] \\
= \frac{m_{S}(t) \cdot \left[1 + \left( \frac{L}{S} \right)(t) \right]}{(4/3) \pi R_{\text{Agg}}(t)^3}
\]

The last Shrinking Core Model parameter is the temperature at the reaction front (\( T_{\text{reaction}} \)). It is obtained by comparing and minimizing the standard deviation of the model data presented by House (2007).

All the relevant drying model parameters and their references in the literature are summarized in Table 2-2.
Table 2-2. Parameters used in thermal drying model

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Values</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td>Solids density $\rho_S$ (kg·m$^{-3}$)</td>
<td>1470</td>
<td>Mohagheghi Dar Ranji (2015)</td>
</tr>
<tr>
<td>Bed temperature $T_{Bed}$ (°C)</td>
<td>550</td>
<td>Sanchez Careaga (2013),</td>
</tr>
<tr>
<td></td>
<td>[530; 560]</td>
<td></td>
</tr>
<tr>
<td>Reaction front temperature $T_{reaction}$ (°C)</td>
<td>520</td>
<td>House (2007)</td>
</tr>
<tr>
<td>Thermal conductivity of coke $k_S$ (W·m$^{-1}$·K$^{-1}$)</td>
<td>1.0</td>
<td>House (2007)</td>
</tr>
<tr>
<td>Enthalpy of vaporization &amp; reaction of liquid bitumen $\Delta H_{liq}$ (J·kg$^{-1}$)</td>
<td>$1.15241 \times 10^6$ (at 550 °C)</td>
<td>Sanchez Careaga (2013)</td>
</tr>
<tr>
<td>Liquid bitumen heat capacity $C_{p,L}$ (J·kg$^{-1}$·K$^{-1}$)</td>
<td>2,272.1 (at 550 °C)</td>
<td>Sanchez Careaga (2013)</td>
</tr>
<tr>
<td>Product vapour heat capacity $C_{p,G}$ (J·kg$^{-1}$·K$^{-1}$)</td>
<td>1750 (at 550 °C)</td>
<td>(Bureau of Standards, 1929; Gas Processors Association, 2016)</td>
</tr>
<tr>
<td>Coke yield $y_S$</td>
<td>0.20</td>
<td>Ali et al. (2010), Sanchez Careaga (2013)</td>
</tr>
</tbody>
</table>

2.4.2.3 Impact of wet-agglomerate initial parameters on drying

This section presents a brief analysis of the wet-agglomerate initial parameters on the calculated drying kinetics.

Figure 2-12 presents the evolution of the liquid-to-solid ratio of a wet-agglomerate as a function of the dimensionless time spent in the bed (defined in section 2.2.2) for various initial agglomerates properties.

It is evident from Figure 2-12 that:
- Wet-agglomerates with a smaller initial diameter $D_{Agg,0}$ dry faster. This result is due to a faster reaction rate and is visible in Equation (2.11), where an increase of wet-agglomerate size increases the required time for complete drying.
- Wet-agglomerates with a smaller initial liquid-to-solid ratio $(L/S)_0$ dry faster, as they contain less liquid to evaporate.

**a)**

![Graph showing the effect of initial agglomerate size and liquid-to-solid ratio on drying time](image1)

- $\rho_{Agg,0} = 1100 \text{ kg}\cdot\text{m}^{-3}$, $(L/S)_0 = 0.50$
- $D_{Agg} = 0.5 \text{ cm}$
- $D_{Agg} = 1 \text{ cm}$
- $D_{Agg} = 1.5 \text{ cm}$
- $D_{Agg} = 2 \text{ cm}$

**b)**

![Graph showing the effect of liquid-to-solid ratio on drying time](image2)

- $\rho_{Agg,0} = 1100 \text{ kg}\cdot\text{m}^{-3}$
- $(L/S)_0 = 0.50$, $D_{Agg} = 1.5 \text{ cm}$
- $(L/S)_0 = 0.50$, $D_{Agg} = 1 \text{ cm}$
- $(L/S)_0 = 0.20$, $D_{Agg} = 1 \text{ cm}$
- $(L/S)_0 = 0.20$, $D_{Agg} = 1.5 \text{ cm}$
Wet-agglomerates with a smaller initial agglomerate density $\rho_{Agg,0}$, and therefore higher voidage $\varepsilon_{Agg,0}$, dry moderately faster. This is because they offer less resistance to heat transfer. As seen in Equation (2.11) and Equation (2.12), lighter agglomerates increase the value of the parameter $\gamma$, reducing the time required for complete drying.

2.4.3 Scaling summary

This section is a summary of the procedures for the agglomerate drying scaling between the experimental unit and the commercial Fluid Coker

First, for the wet-agglomerates time since production (or last re-wetting if considered), the dimensionless time ($t/\tau$) defined in Section 2.2.2 is used. Using Equation (2.5) allows for the computation of the "time in Fluid Coker" $t'$ (calculated from the recorded "time in the pilot" $t$) representative of the age of a wet-agglomerate in a commercial Fluid Coker.

Second, for the wet-agglomerates position, the scaling procedure presented in Section 2.2.1 is used. This implies a similar dimensionless radial location ($r/R$) and a similar
dimensionless radius to bed height ($r/h$) between the experimental unit and the commercial Fluid Coker.

Third, the initial properties of wet-agglomerates ($D_{Agg,0}$, $\rho_{Agg,0}$, $(L/S)_0$) and the properties relative to their thermal drying ($T_{Bed}$, $\rho_L$, $\mu_L$, ...) are defined for a realistic representation of a commercial Fluid Coker, following the recommendations of Section 2.2.3.

2.5 Wet-agglomerate destruction

In a commercial Fluid Coker, most agglomerates break up before reaching the stripper, as they interact with the bed, and most specifically, gas bubbles and their wakes (Weber, 2009)

Wet-agglomerates moving inside the bed are interacting with it. This interaction might compromise the physical integrity of the wet-agglomerate, leading to its breakage into smaller fragments. Agglomerates can be destroyed through two distinct mechanisms: erosion or fragmentation. As mentioned in Chapter 1 – Section 1.3.3, for the range of fluidization velocities used in Fluid Coker, the fragmentation destruction mode is dominant (Boyle et al., 2005; Weber, 2009). Therefore, fragmentation is the only destruction mode considered in the model.

The model checks if a wet-agglomerate of a given age (defined as the time since production, or since the last re-wetting) and located at a given position will break. To do so, it calculates and compares two parameters. First, the agglomerate yield strength, which is purely a function of the wet-agglomerate age. Second, the critical agglomerate yield strength, at the position considered in the bed and for a wet-agglomerate of a given age.

A flowchart summarizing the wet-agglomerate breakage component of the model is shown in Figure 2-13.
TRACER IN
REST OF BED
B) Test for possible shear breakage

Overview

Figure 2-13. Summarized flowchart of the shear breakage process in the model

2.5.1 Destruction test

At any time, the model performs a wet-agglomerate destruction test by checking if the local bed shear at the wet-agglomerate position is strong enough to break an agglomerate of a given age. This test is expressed as follow (Tardos et al., 1997; Weber, 2009):
\[
\begin{align*}
\tau_{agg}(t) &< \tau_{agg}^*([x, y, z], t), & \text{Breakage} \\
\tau_{agg}(t) &\geq \tau_{agg}^*([x, y, z], t), & \text{NO breakage}
\end{align*}
\] (2.20)

with:

- \(\tau_{agg}(t)\): agglomerate yield strength (or "yield stress") (Pa), at a given time \(t\)
- \(\tau_{agg}^*([x, y, z], t)\): critical agglomerate yield strength (Pa), at a given position \([x, y, z]\) and a given time \(t\)

Figure 2-14 summarizes the mechanism and outcomes of the wet-agglomerate destruction test performed.

**Figure 2-14. Wet-agglomerate destruction test**

It should be pointed out here that a similar approach was proposed by Shi et al. (2017) with a different set of equations. A major difference was an explicit calculation of the probability of encounter between a single wet-agglomerate and a single bubble. This probability is
implicitly calculated and averaged, for an overall bubble population, in the bed shear computation used in the model presented in this chapter.

An important model limit is related to the outcome of the wet-agglomerate after they break. It is assumed that the resulting fragments are small enough to dry very quickly. As a first approximation, this is computed by assuming an instantaneous drying of these fragments. This might not be the case in real systems, especially for the largest agglomerates, producing fragments of significant size.

2.5.1.1 Agglomerate yield strength

The agglomerate yield strength is calculated using the characteristic granule stress, $\tau_{\text{granule}}$, from the Herschel-Bulkley model (Tardos et al., 1997). As a first approximation, the apparent viscosity $k_{\text{HB}}$ is assumed negligible compared to the yield strength $\tau_{\text{Agg}}$. This is equivalent to considering the wet-agglomerates as a highly concentrated slurry of binder and particles (Tardos et al., 1997).

\[ \tau_{\text{granule}} = \tau_{\text{agg}} + k_{\text{HB}} \cdot \dot{\gamma}^{n_{\text{HB}}} \approx \tau_{\text{agg}} \quad (2.21) \]

with:

- $\tau_{\text{granule}}$: agglomerate characteristic granule stress (Pa)
- $k_{\text{HB}}$: apparent viscosity for the Herschel-Bulkley model (Pa·s)
- $\dot{\gamma}$: fluidized bed shear rate (s$^{-1}$)
- $n_{\text{HB}}$: flow index

The model uses the following assumptions for the agglomerate strength calculations. First, the coke is a fully wetted solid. Second, the wet-agglomerates are spherical. Therefore, the effect of the particle shape was not considered in the calculation of the agglomerate strength.

The equation used to calculate the agglomerate strength varies depending on its saturation and saturation state. The agglomerate saturation $S_{\text{Agg}}(t)$ is defined as the pore volume occupied by liquid divided by the total void volume of the agglomerate. This parameter varies with time as the wet-agglomerate dries. The values of $S_{\text{Agg}}$ are physically bounded
between 0 (no liquid in the wet-agglomerate) and 1 (all interstitial space occupied by liquid).

\[
S_{\text{agg}}(t) = \frac{\varepsilon_L}{1 - \varepsilon_S} = \frac{(L/S)(t) \cdot \rho_{agg}(t) \cdot \rho_S}{\rho_L \cdot (\rho_S + (L/S)(t) \cdot \rho_S - \rho_{agg}(t))}
\]  

(2.22)

**Figure 2-15.** Saturation states in liquid-solid agglomerates (Weber, 2009)

a) Pendular state, b) Funicular state, c) Capillary state

Once the saturation is calculated, the saturation state is known. Sherrington and Oliver (1980) defined the saturation states as the three modes of cohesion of wet-agglomerates. Driest agglomerates are in a pendular state, bounded only by discrete liquid bridges. Wettest agglomerates are in a capillary state, with the liquid filling most of the void, and the dominant cohesive force comes from capillary suction. The Funicular state is intermediate and transitional. The boundaries between the pendular/funicular state and the funicular/capillary state are denoted respectively, \( S_{\text{Agg},P} \) and \( S_{\text{Agg},C} \). The three states are presented in Figure 2-15.

For agglomerates in the pendular state \( (S_{\text{Agg}} < S_{\text{Agg},P}) \), the agglomerate strength is calculated using an empirical equation (Weber, 2009) developed based on theory of maximum static liquid bridge force (Seville et al., 2000) for an agglomerate voidage \( \varepsilon_{\text{Agg}} = 0.4 \)

\[
\tau_{\text{Agg}}(t) = \tau_{\text{Agg},P}(t) = 8.5 \times 10^{12} \ (1000 \ \mu_L)^{0.9} (S_{\text{Agg}}(t))^{0.35} (\pi d_{p,SM} \gamma_L)^2
\]  

(2.23)

with:

- \( \mu_L \): liquid bitumen viscosity (Pa·s)
- \( \rho_L \): liquid bitumen density (kg·m\(^{-3}\))
- \( \gamma_L \): liquid bitumen surface tension (N·m\(^{-1}\))
• \( d_{p,SM} \): Sauter-Mean diameter of coke particles (m)

To use the equation with agglomerates of different voidage, a correction based on the Rumpf (1961) theoretical equation for the pendular state was introduced. This is important as lowering the porosity of an agglomerate increases the strength of the granule (Iveson, 2001)

Since the empirical equation from Weber (2009) was obtained with agglomerates of voidage \( \varepsilon_{Agg} = 0.4 \), Equation (2.23) is modified as follow:

\[
\tau_{Agg}(t) = \tau_{Agg,P}(t) \\
= \left( \frac{1 - \varepsilon_{Agg}}{\varepsilon_{Agg}} \right) \cdot 8.5 \times 10^{12} \left( 1000 \, \mu_L \right)^{0.9} \left( S_{Agg}(t) \right)^{0.35} \left( \pi d_{p,SM} \gamma_L \right)^2
\]

\[
= \left( \frac{1 - \varepsilon_{Agg}}{\varepsilon_{Agg}} \right) \cdot 5.67 \times 10^{12} \left( 1000 \, \mu_L \right)^{0.9} \left( S_{Agg}(t) \right)^{0.35} \left( \pi d_{p,SM} \gamma_L \right)^2
\]

For agglomerates in the capillary state \((S_{Agg} > S_{Agg, C})\), the equation used for the agglomerate strength is based on capillary pressure \( P_C \) (Schubert, 1973; Schubert, 1975)

\[
\tau_{Agg}(t) = \tau_{Agg,C}(t) = S_{Agg}(t). P_C
\]

with:

• \( P_C \): capillary pressure (Pa). This parameter is hard to obtain, and it is usually calculated using an empirical equation. For well-wetted solids and agglomerate in drainage mode (such as drying) (Schubert, 1984):

\[
P_C = M \cdot \frac{1 - \varepsilon_{Agg}}{\varepsilon_{Agg}} \cdot \frac{\gamma_L}{d_{p,SM}}
\]

The empirical constant \( M \) was observed in the literature as varying within certain bounds, depending on the size range of solids particles considered.

- \( M \in [6; 8] \) for a narrow size range of particles (Pietsch et al., 1969; Schubert, 1973; Capes, 1980; Wynnyckyj, 1985)
- \( M \in [1.9; 14.5] \) for a wider size range of particles (Schubert, 1984; Kim and Sture, 2008). The larger variations are due to the fact that smaller particles fit in the
interstitial space between the larger particles modifying the strength of the bonds (Schaafsma et al., 1998; Iveson, 2001)

In this research, the solids used are fluid coke with a narrow range of particle size (see Appendix A: Fluid coke particle sized distribution). Therefore, the base value selected was M = 7. Since there was no additional criterion for further choice and no easy means to better characterize this coefficient, the sensitivity of the model results regarding this parameter will be tested in Section 2.9.1.3. In addition, for critical model results, the calculations will also be verified with other values of M in the range of 6-8.

For agglomerates in the funicular state (S_{Agg,P} < S_{Agg} < S_{Agg,C}), a transition equation was used (Schubert, 1973; Schubert, 1984):

\[
\tau_{Agg}(t) = \tau_{Agg,P}(t) \frac{S_{Agg,C} - S_{Agg}(t)}{S_{Agg,C} - S_{Agg,P}} + \tau_{Agg,C}(t) \frac{S_{Agg}(t) - S_{Agg,P}}{S_{Agg,C} - S_{Agg,P}}
\]  

(2.27)

with:

- \( S_{Agg,P} \): maximum agglomerate saturation in the pendular state. \( S_{Agg,P} \) was set equal to 25 % (Weber, 2009)
- \( S_{Agg,C} \): minimum agglomerate saturation in the capillary state. \( S_{Agg,C} \) was set equal to 80 % (Weber, 2009)

All the relevant breakage model parameters and their references in the literature are summarized in Table 2-3.

**Table 2-3. Parameters used for agglomerate yield strength**

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td>( \mu_L ) (Pa·s)</td>
<td>3.0 ( \times ) 10^{-3}</td>
<td>Weber (2009)</td>
</tr>
<tr>
<td></td>
<td>[1.0 ( \times ) 10^{-3}; 3.0 ( \times ) 10^{-3}]</td>
<td></td>
</tr>
<tr>
<td>( \rho_L ) (kg·m^{-3})</td>
<td>893</td>
<td>Argüelles-Vivas et al. (2012), Guan et al. (2013)</td>
</tr>
<tr>
<td>( \gamma_L ) (N·m^{-1})</td>
<td>2.2 ( \times ) 10^{-2}</td>
<td>Drelich and Miller (1994), Weber (2009)</td>
</tr>
<tr>
<td></td>
<td>[2.05 ( \times ) 10^{-2}; 2.7 ( \times ) 10^{-2}]</td>
<td></td>
</tr>
<tr>
<td>( d_{p,SM} ) (m)</td>
<td>140 ( \times ) 10^{-6}</td>
<td>MEASURED (Appendix A: Fluid coke particle sized distribution)</td>
</tr>
</tbody>
</table>
2.5.1.1.1 Impact of the wet-agglomerate initial parameters on agglomerate strength

This section presents a brief analysis of the effect of the wet-agglomerate initial parameters on the agglomerate strength.

Figure 2-16 presents the evolution of the strength of a wet-agglomerate as a function of its saturation for different initial agglomerate properties. The three saturation states are visible. The decrease in the pendular state is critical, as when an agglomerate reaches this state, it is likely to break quickly. As expected, lowering the porosity of an agglomerate increases the strength of the granule (Iveson, 2001).

![Figure 2-16. Evolution of agglomerate strength with agglomerate saturation, for different agglomerate properties](image)

\[
\begin{align*}
\rho_{\text{Agg},0} &= 1100 \text{ kg·m}^{-3}, (L/S)_0 = 0.10, \\
\epsilon_{\text{Agg},0} &= 0.32 \\
\rho_{\text{Agg},0} &= 1200 \text{ kg·m}^{-3}, (L/S)_0 = 0.31, \\
\epsilon_{\text{Agg},0} &= 0.37 \\
\rho_{\text{Agg},0} &= 1300 \text{ kg·m}^{-3}, (L/S)_0 = 0.60, \\
\epsilon_{\text{Agg},0} &= 0.45
\end{align*}
\]

Figure 2-17 shows the evolution of the strength of a wet-agglomerate as a function of its saturation for different values of the empirical coefficient M. The variations are limited to the capillary state and funicular state when \( S_{\text{Agg}} > 0.4 \).
Figure 2-17. Evolution of agglomerate strength with agglomerate saturation, for different empirical coefficient $M$ for the capillary pressure:

$(\rho_{\text{Agg,0}} = 1100 \text{ kg·m}^{-3}, (L/S)_0 = 0.31, \varepsilon_{\text{Agg,0}} = 0.37, D_{\text{Agg}} = [5.0; 15] \text{ mm, } \rho_s = 1470 \text{ kg·m}^{-3}, \rho_L = 893 \text{ kg·m}^{-3}, \mu_L = 3 \times 10^{-3} \text{ Pa·s, } d_{p,SM} = 140 \mu\text{m, } \gamma_L = 2.05 \times 10^{-2} \text{ N·m}^{-1})$

Figure 2-20

Figure 2-18. Evolution of agglomerate strength with agglomerate Liquid-to-Solid Ratio, for different initial properties (constant initial agglomerate density)
$(\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}, (L/S)_0 = 0.31, \varepsilon_{Agg,0} = 0.37, D_{Agg} = [5.0; 15] \text{ mm}, \rho_{S} = 1470 \text{ kg} \cdot \text{m}^{-3}, \\
\rho_{L} = 893 \text{ kg} \cdot \text{m}^{-3}, \mu_{L} = 3 \times 10^{-3} \text{ Pa} \cdot \text{s}, d_{p,SM} = 140 \mu\text{m}, \gamma_{L} = 2.05 \times 10^{-2} \text{ N} \cdot \text{m}^{-1})$}

Figure 2-18 presents the evolution of the strength of a wet-agglomerate as a function of its liquid-to-solid ratio for different initial agglomerate properties. As the agglomerate voidage $\varepsilon_{Agg}$ increases (as the agglomerate size increases), its strength decreases significantly: bigger agglomerates are more likely to break.

Figure 2-19 shows the evolution of the strength of a wet-agglomerate as a function of its liquid-to-solid ratio for different initial agglomerate densities. Minor differences are observable and only for the pendular state.

Figure 2-22

**Figure 2-19.** Evolution of agglomerate strength with agglomerate Liquid-to-Solid Ratio, for different initial properties (constant agglomerate voidage)

$(\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}, (L/S)_0 = 0.31, \varepsilon_{Agg,0} = 0.37, D_{Agg} = [5.0; 15] \text{ mm}, \rho_{S} = 1470 \text{ kg} \cdot \text{m}^{-3}, \\
\rho_{L} = 893 \text{ kg} \cdot \text{m}^{-3}, \mu_{L} = 3 \times 10^{-3} \text{ Pa} \cdot \text{s}, d_{p,SM} = 140 \mu\text{m}, \gamma_{L} = 2.05 \times 10^{-2} \text{ N} \cdot \text{m}^{-1})$}

Figure 2-20 presents the evolution of the strength of a wet-agglomerate as a function of its liquid-to-solid ratio for different values of the empirical coefficient M. The variations are limited in amplitude and occur only for (L/S) values greater than about 0.15.
Figure 2-20. Evolution of agglomerate strength with agglomerate Liquid-to-Solid Ratio, for different initial properties and with a constant initial agglomerate density

\[(\rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, \rho_S = 1470 \text{ kg} \cdot \text{m}^{-3}, \rho_L = 893 \text{ kg} \cdot \text{m}^{-3}, \mu_L = 3 \times 10^{-3} \text{ Pa} \cdot \text{s}, \]
\[d_{p,\text{SM}} = 140 \text{ \mu m}, \gamma_L = 2.05 \times 10^{-2} \text{ N} \cdot \text{m}^{-1}, M = 7)\]

Figure 2-21. Evolution of agglomerate strength with agglomerate Liquid-to-Solid Ratio, for different initial properties and with a constant agglomerate voidage

\[(\varepsilon_{\text{Agg,0}} = 0.476, D_{\text{Agg}} = [5.0; 15] \text{ mm}, \rho_S = 1470 \text{ kg} \cdot \text{m}^{-3}, \rho_L = 893 \text{ kg} \cdot \text{m}^{-3}, \mu_L = 3 \times 10^{-3} \text{ Pa} \cdot \text{s}, \]
\[d_{p,\text{SM}} = 140 \text{ \mu m}, \gamma_L = 2.05 \times 10^{-2} \text{ N} \cdot \text{m}^{-1}, M = 7)\]
Figure 2-22. Evolution of agglomerate strength with agglomerate Liquid-to-Solid Ratio, for different empirical coefficient M and with constant initial agglomerate properties

\( \rho_{Agg,0} = 1100 \; \text{kg} \cdot \text{m}^{-3}, \; (L/S)_0 = 0.31, \; \varepsilon_{Agg,0} = 0.37, \; D_{Agg} = [5.0; 15] \; \text{mm}, \; \rho_S = 1470 \; \text{kg} \cdot \text{m}^{-3}, \rho_L = 893 \; \text{kg} \cdot \text{m}^{-3}, \; \mu_L = 3 \times 10^{-3} \; \text{Pa} \cdot \text{s}, \; d_{p,SM} = 140 \; \mu\text{m}, \; \gamma_L = 2.05 \times 10^{-2} \; \text{N} \cdot \text{m}^{-1}) \)

Figure 2-21 presents the evolution of the strength of a wet-agglomerate as a function of its liquid-to-solid ratio for different initial properties (constant initial agglomerate density). Strong variations are visible, with smaller, dryer and more compact agglomerates significantly stronger.

Figure 2-22 presents the evolution of the strength of a wet-agglomerate as a function of its liquid-to-solid ratio for different initial properties (constant agglomerate voidage). The variations are minimal.

2.5.1.2 Critical agglomerate yield strength

The second component of the destruction test is the critical agglomerate yield strength (Tardos et al., 1997; Iveson et al., 2001; Liu et al., 2009; Hapgood and Litster, 2019). This parameter varies with local bed conditions \((d_b(x, y, z), u_b(x, y, z), \varepsilon_b(x, y, z))\) and with the agglomerate age \((\rho_{Agg}(t))\)
The critical agglomerate yield strength is computed using the critical Stokes deformation number $St_{\text{def}}^*$ (Tardos et al., 1997). This parameter is based on the following fundamental formula:

$$St_{\text{def}}^* = \frac{\text{Externally applied kinetic energy}}{\text{Energy required for granule deformation}} = \frac{m_s \cdot (R_{\text{Agg}} \cdot \dot{\gamma})^2}{2 \cdot V_s \cdot \tau_{\text{granule}}} \quad (2.28)$$

Experimental work demonstrated that a critical Stokes deformation number $St_{\text{def}}^* = 0.2$ was characteristic of wet-agglomerate breakage (Tardos et al., 1997; Liu et al., 2009). If $St_{\text{def}} > St_{\text{def}}^* = 0.2$, then the wet-agglomerate breaks.

By rearranging Equation (2.28), the model can compute the critical agglomerate yield strength $\tau_{\text{Agg}}^*$ as:

$$\tau_{\text{Agg}}^* (x, y, z) = \frac{\rho_s}{2 \cdot St_{\text{def}}} \left( R_{\text{Agg},0} \cdot \dot{\gamma}(x, y, z) \right)^2 \quad (2.29)$$

with:

- $\tau_{\text{Agg}}^* (x, y, z)$: critical agglomerate yield strength (Pa), at a given position $[x, y, z]$ and a given time $t$
- $St_{\text{def}}^*$: critical Stokes deformation number
- $\dot{\gamma}(x, y, z)$: average fluidized bed shear rate, at a given position $[x, y, z]$ ($s^{-1}$) (Ennis et al., 1991; Tardos et al., 1997)

It should be noted that bigger agglomerates are associated with larger critical agglomerate yield strength $\tau_{\text{Agg}}^*$. They are therefore more prone to breakage. In addition, the value of the critical agglomerate yield strength $\tau_{\text{Agg}}^*$ is strongly correlated with the intensity of the average fluidized bed shear rate $\dot{\gamma}$ at a given position $[x, y, z]$. More bed shear means that agglomerates are more likely to break.

The average fluidized bed shear rate $\dot{\gamma}$, at a given position $[x, y, z]$, is estimated using the two-phase inviscid flow theory (Clift and Grace, 1985). It considers a spherical inviscid dense-phase moving around a bubble and is given by the following equation:

$$\dot{\gamma}(x, y, z) = \frac{18 \cdot u_b(z)}{d_b \cdot \delta^2} \quad (2.30)$$
The dimensionless distance between two bubbles $\delta_A$ is defined as:

$$\delta_A = \frac{s_b(x, y, z)}{r_b} \quad (2.31)$$

with:

- $s_b$: bubble spacing (m)

The bubble velocity $u_b$ is calculated using a correlation from Hilligardt and Werther (1986).

$$u_b(z) = \varphi_b(z, R_{bed}(z)) \left( u_G(z) - u_{mf} \right) + 0.711 \theta_b(R_{bed}(z)) \sqrt{2. g \cdot r_b} \quad (2.32)$$

with:

- $\varphi_b(z/R_{bed}(z))$ & $\theta_b(R_{bed}(z))$: correction factors function of the ratio of the bubble height location to fluidized bed diameter $z/R_{bed}(z)$, the diameter of the fluidized bed $R_{bed}(z)$, and the Geldart (1973) powder classification of the fluidized particles (Hilligardt and Werther, 1986). Fluid coke belongs to group B powders.
- $u_G(z)$: superficial gas velocity, at height $z$
- $u_{mf}$: minimum fluidization velocity (m·s$^{-1}$)
- $g$: local gravitational field of Earth ($g = 9.80665$ m·s$^{-2}$)

It should be noted that correction factors $\varphi_b$ and $\theta_b$ used in the Hilligardt and Werther equation need to be scaled as follows:

- $\varphi_b$: similar geometric dimensionless profile, use experimental unit values $z_{EXP}$ and $R_{EXP}$
- $\theta_b$: use radius of commercial Coker, $R_{Coker}$

The cross-sectional superficial gas flux $F_{LG}$ is also assumed constant between injection banks. Therefore, there is a constant cross-sectional superficial gas velocity $u_G$ if the bed radius $R_{Bed}$ is constant.

Regarding the bubble diameter $d_b$ and bubble radius $r_b$ ($r_b = d_b/2$), a critical assumption made is that they are constant and equal to a specific value in the turbulent part of the bed. This is an approximation as the velocity change in commercial Fluid Coker makes the bed
go from a bubbling regime in its lower part (with a bubble size increasing with height) to a turbulent regime in its higher part (with an almost constant bubble size). This will be discussed in more details in Section 2.5.4. The bubble diameter $d_b$ used in the model is estimated for the conditions of operations of commercial Fluid Coker, using data found in the literature: $d_b \in [4.0 \times 10^{-2}; 7.5 \times 10^{-2}]$ m. The sources of these values are given in Table 2-4. The sensitivity of the bed shear to $d_b$ is discussed in Section 2.5.1.2.1. The sensitivity of the model results to $d_b$ is tested in Section 2.9.1.4.

The effect of bed temperature on bubble size is small. Sahu et al. (2018) found, for group B particles in a bed without recirculation, the bubble size remains almost constant with an increase in temperature ($T_{\text{bed}} \in [30; 600]$ °C).

The transition between the bubbling and the turbulent regime is discussed in Section 2.5.4. The location of this transition is of particular interest, both in the experimental unit and in commercial Fluid Coker.

Table 2-4. Some experimental measurements of bubble diameter $d_b$ in fluidized bed

<table>
<thead>
<tr>
<th>Experimental values</th>
<th>Conditions</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td>$d_b$ (m)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>[5.0 $\times$ 10^{-2}; 7.5 $\times$ 10^{-2}]</td>
<td><strong>Group A</strong>&lt;br&gt; FCC catalyst/Air&lt;br&gt; $d_{\text{p,SM}} = 64$ μm&lt;br&gt; $\rho_s \approx 1500$ kg·m$^{-3}$&lt;br&gt; $u_G \in [0.45; 1.1]$ m·s$^{-1}$&lt;br&gt; <em>Turbulent</em>&lt;br&gt; $D_{\text{bed}} = 0.2$ m, $H_{\text{bed}} = 0.85$ m&lt;br&gt; $T \approx 20$°C, $P \approx 1$ atm</td>
<td><em>Bi et al. (2000)</em></td>
</tr>
<tr>
<td>[4.0 $\times$ 10^{-2}; 7.5 $\times$ 10^{-2}]</td>
<td><strong>Group B</strong>&lt;br&gt; Glass Beads/Air&lt;br&gt; $d_{\text{p,SM}} = 139$ μm&lt;br&gt; $\rho_s = 2400$ kg·m$^{-3}$&lt;br&gt; $u_G = 0.1, 0.8$ &amp; 0.9 m·s$^{-1}$&lt;br&gt; <em>Turbulent</em>&lt;br&gt; $D_{\text{bed}} = 0.2$ m, $H_{\text{bed}} = 4.0$ m&lt;br&gt; $T \approx 20$°C, $P = 1$ atm</td>
<td><em>Gao et al. (2012), Wu et al. (2020)</em></td>
</tr>
<tr>
<td></td>
<td>Group B</td>
<td>(Chew and Hrenya, 2011; Chew and Cocco, 2021)</td>
</tr>
<tr>
<td>--------------------------------</td>
<td>------------------------</td>
<td>---------------------------------------------</td>
</tr>
<tr>
<td>[2.0 , 10^{-2}; , 4.5 , 10^{-2}]</td>
<td>Silica Sand/Air</td>
<td></td>
</tr>
<tr>
<td></td>
<td>( \bar{d}_{p,SM} = 375 , \mu m )</td>
<td></td>
</tr>
<tr>
<td></td>
<td>( \rho_S = 2650 , \text{kg} \cdot \text{m}^{-3} )</td>
<td></td>
</tr>
<tr>
<td></td>
<td>( u_G = 0.15 , \text{m} \cdot \text{s}^{-1} )</td>
<td></td>
</tr>
<tr>
<td></td>
<td><strong>Bubbling</strong></td>
<td></td>
</tr>
<tr>
<td></td>
<td>( D_{\text{bed}} = 0.19 , \text{m} ), ( m_S = 8 , \text{kg} )</td>
<td></td>
</tr>
<tr>
<td></td>
<td>( T \approx 20^\circ \text{C}, , P = 1 , \text{atm} )</td>
<td></td>
</tr>
</tbody>
</table>

The bubble spacing \( s_b \) is defined with the assumption of a homogeneous population of spherical bubbles (same diameter \( d_b \)) located in a bed region in the turbulent regime. Furthermore, these bubbles are assumed to be locally distributed in a uniform cube geometry of side \( s_b \). In addition, since the bubble voidage, \( \varepsilon_b \), cannot be measured directly, it is approximated from the bed voidage, \( \varepsilon_{\text{bed}} \). These combined assumptions give the following equation:

\[
s_b(x, y, z) = r_b \cdot \left( \frac{4}{3} \cdot \frac{\pi}{\varepsilon_{\text{bed}}(x, y, z) - \varepsilon_m} \right)^{\frac{1}{3}} \tag{2.33}
\]

with:

- \( \varepsilon_{\text{bed}}(x, y, z) \): local bed voidage
- \( \varepsilon_m \): emulsion phase voidage at minimum fluidization

By combining Equations (2.31) and (2.33), the model can obtain the following simplification:

\[
\delta_A = \frac{s_b(x, y, z)}{r_b} = \left( \frac{4}{3} \cdot \frac{\pi}{\varepsilon_{\text{bed}}(x, y, z) - \varepsilon_m} \right)^{\frac{1}{3}} \tag{2.34}
\]

All the relevant remaining parameters for the calculations of the critical agglomerate strength and their references in the literature are summarized in Table 2-5.
Table 2-5. Parameters used for critical agglomerate yield strength

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td>(u_{mf}) (m·s(^{-1}))</td>
<td>5.16 (10^{-3})</td>
<td><em>Calculated with Fluid Coker conditions</em></td>
</tr>
<tr>
<td></td>
<td></td>
<td><em>Wen and Yu (1966), Richardson and da S. Jerónimo (1979)</em></td>
</tr>
</tbody>
</table>

2.5.1.2.1 Effect of bubble diameter \(d_b\) on bed shear

This section briefly discusses the impact of the bubble diameter \(d_b\) on the bed shear.

Two competing effects are applied on the bed shear \(\dot{\gamma}\) when \(d_b\) increases:

- First, \(u_b\) increase with \(d_b\), increasing the bed shear, \(\dot{\gamma}\). As bubbles accelerate, they have higher kinetic energy. Therefore, when they impact an agglomerate, they create more shear.
- On the other hand, the bubble spacing, \(s_b\), increases, which can also be expressed as a reduction of the local bubble density. This leads to a decrease in the bed shear, \(\dot{\gamma}\). This phenomenon can be expressed as a lower frequency of impact between bubbles and agglomerates.

These two competing mechanisms match the ones observed with other models (Shi et al., 2017; Sun et al., 2020). To summarize: 1) the frequency of breakage (after an encounter) increases when the bubble diameter increases, 2) but the frequency of bubble/agglomerate encounters decreases when the bubble diameter increases (as, for a constant amount of gas, bigger bubbles mean fewer bubbles). Overall, with the equations used, the bed shear, \(\dot{\gamma}\), decreases when \(d_b\) increases, as presented in Figure 2-23.

Therefore, the critical agglomerate strength \(\tau_{Agg}^*\) also decreases when \(d_b\) increases, as seen in Figure 2-24. This decrease is moderate (factor of 2) for a typical range of bubble diameters in a turbulent bed ([4.0; 7.5] cm). The variations are stronger (factor of 4) when considering a range of bubble diameters in a bubbling bed ([2.0; 4.0] cm). These variations will not be accounted for in the model as a single bubble diameter is considered. The sensitivity of the model results to \(d_b\) is tested in Section 2.9.1.4.
Figure 2-23. Evolution of bed shear $\gamma$ & bubble velocity $u_b$, for different values of the bubble diameter $d_b$ & superficial cross-sectional gas velocity $u_G$

$(\varepsilon_{\text{bed}} = 0.60)$

Figure 2-24. Evolution of critical agglomerate strength $\tau_{\text{Agg}}^*$, for different values of the bubble diameter $d_b$ & superficial cross-sectional gas velocity $u_G$

$(\varepsilon_{\text{bed}} = 0.60, D_{\text{Agg}} = 1.0 \text{ cm}, \rho_S = 1470 \text{ kg} \cdot \text{m}^{-3}, St_{\text{def}}^* = 0.2)$
The effect of the cross-sectional superficial gas velocity, $u_G$, on the bed shear, $\dot{\gamma}$, and the critical agglomerate strength, $\tau_{Agg}^*$, are also presented in Figure 2-23 and in Figure 2-24. For the range of cross-sectional superficial gas velocity, $u_G$, used in the commercial Fluid Coker ([0.3; 0.9] m·s$^{-1}$), there are only minor variations of bed shear (around 20%) and moderate variations of critical agglomerate strength (factor of 2). These variations are considered in the model.

### 2.5.1.2.2 Local bed voidage

The local bed voidage $\varepsilon_{bed}(x, y, z)$ is required to compute the critical agglomerate yield strength.

A first assumption used to obtain it is the angular symmetry of the voidage profile:

$$\varepsilon_{bed}(x, y, z) = \varepsilon_{bed}(r, \theta, z) = \varepsilon_{bed}(r, z)$$

Using this assumption, the local bed voidage is measured experimentally in the pilot-scale unit, using pressure-based and radiation transmission based combined measurements.

$$\varepsilon_{bed}(x, y, z) = \varepsilon_{bed}(r, z) = \varepsilon_{bed, AVG}(z) \cdot \left( \frac{\varepsilon_{bed}(r, z)}{\varepsilon_{bed, AVG}(z)} \right)(r, z) \quad (2.35)$$

with:

- $\varepsilon_{bed, AVG}(z) = f(u_G)$ Axial average voidage (pressure-based measurements, as presented in Appendix K: Measurement of the bed voidage profile in the experimental unit)
- $\left( \frac{\varepsilon_{bed}(r,z)}{\varepsilon_{bed, AVG}(z)} \right)(r,z)$ Relative radial local voidage (pressure-based & radiation-based measurements combined as shown in Appendix K: Measurement of the bed voidage profile in the experimental unit) (Song et al., 2006)

The profile obtained is then scaled up to the commercial fluid Coker scale.
TRACER IN REST OF BED

B) Test for possible shear breakage

**Detailed**

- **AGGLOMERATE STRENGTH**
  - **CRITICAL AGGLOMERATE STRENGTH**
    - from local bed hydrodynamics
  - EXPERIMENTAL
    - Pressure-based 3D voidage profiles
    - Radiation-based 3D voidage profiles
  - **MODEL PARAMETER**
    - Granular temperature \( \theta \)
    - Restitution coefficient \( e \)
    - Solid volume fraction for uniform spheres \( \varepsilon_{S,max} \)
    - Empirical coefficient \( M \)
  - **COKER**
    - \( \rho_s, \rho_L, \rho_i \)
    - \( \varepsilon_{mf} \)

- **CALCULATE**
  - Agglomerate density \( \rho_{Agg}(t^*) \)
  - Agglomerate saturation \( S_{Agg}(t) \)
  - Pendular state
  - Capillary state
  - Capillary pressure \( P_c(t^*) \)
  - Minimum fluidization velocity \( U_{mf} \)
  - Bed voidage \( \varepsilon_{bed}(z) \)
  - Bubble velocity \( u_b(z) \)
  - Bubble separation \( s_b(r, z) \)
  - Bubble radius \( r_b(z) \)
  - Critical agglomerate strength \( \tau_{Agg}^*(r, z) \)
  - Critical Stoke number \( St_{def} \)
  - Critical bubble velocity \( u_{b,c} \)
  - Critical bed voidage \( \varepsilon_{bed,c} \)
  - Average bed voidage \( \varepsilon_{bed,AVG}(z) \)
  - Bed voidage profile \( \varepsilon_{bed}(z', z^2) \)
  - Bed voidage profile \( \varepsilon_{bed,AVG}(z') \)
  - Bed voidage profile \( \varepsilon_{bed,c}(z', z^2) \)
  - Bed voidage profile \( \varepsilon_{bed,AVG}(z') \)

- **CHECK**
  - Saturation state
  - Capillary state

- **OUTPUT**
  - Information about agglomerate destroyed

**Diagram:** Figure 2-25. Detailed flowchart of the shear breakage process in the model

**Assumptions:**
- Shrinkage core model = constant \( D_{Agg} \)
- Assumption of angular symmetry
- Assumption of constant \( \varepsilon_{mf} \)
- Assumption of constant \( U_{mf} \)
- Assumption of constant \( r_b(z) \)
- Assumption of constant \( \theta \)
- Assumption of constant \( e \)
2.5.2  Summary of shear breakage

A flowchart presenting a detailed overall view of the wet-agglomerate breakage component of the model is presented in Figure 2-25.

2.5.3  Shear maps

By combining the correlations and equations mentioned in the previous sections, the model can calculate the bed shear maps for a given unit geometry and operating conditions. Such maps are presented in Figure 2-26.

The bed shear increases with the superficial gas velocity. Therefore, due to the assumption of constant superficial gas flux, $F_{LG}$, between injection banks, the bed shear increases with the bed height, with clear separation at the location of injection banks. It will also increase with cross-sectional area restrictions, such as those created by inserting a ring baffle in the bed.

On the other hand, bed shear is also a function of the local bed voidage. As mentioned in the previous section, the local voidage was obtained through a combination of pressure-based and radiation-based measurements then fitted using an assumption of a unimodal core annulus (see Appendix K: Measurement of the bed voidage profile in the experimental unit). The comparison of measured bed voidage acquired on the experimental unit with and without a baffle showed two main differences: 1) the average bed voidage at a given height is lower in the zone above the baffle, 2) the radial voidage at a given height is flatter above the baffle. These two effects combined lead to a reduction of the bed shear above a baffle.

Therefore, when a baffle is added, the local increase of superficial gas velocity and the bed voidage have opposite effects.
2.5.4 Transition bubbling-to-turbulent

This section discusses the transition between the bubbling and turbulent regimes in the experimental unit and how it would scale up to a commercial Fluid Coker.

The bubbling/turbulent transition is measured in the experimental unit, using both the standard deviation of pressure measurements and 1D pseudo-voidage calculations from radiation transmission measurements. Both measurements are acquired in the same zone (of fully developed flow) and recorded simultaneously. Some correlations are also used for comparison.
Determining the bubbling/turbulent transition and its location in the experimental unit is essential for the shear breakage calculations in the model to support the assumption of constant bubble diameter, $d_b$. They also give some insight into the validity of the experimental measurements of voidage shape variation (presented in Appendix K: Measurement of the bed voidage profile in the experimental unit).

Figure 2-27 presents the measured bubbling/turbulent transition as a function of superficial cross-section fluidization velocity.

Table 2-6 summarizes the detected transition velocities from these same experiments.

Table 2-7 presents the range of transition velocities given by selected correlations for the pilot plant system used in this study in the experimental unit.

\[ a) \] Figure 2-27. Experimental detection of bubbling/turbulent transition as a function of superficial cross-section fluidization velocity
a) From the standard deviation of local pressure drop, b) Using a pseudo-voidage calculated from radiation transmission measurements

**Table 2-6. Transition bubbling/turbulent from experimental measurements**

<table>
<thead>
<tr>
<th>Transition bubbling/turbulent (m·s⁻¹)</th>
<th>Correlation values</th>
<th>Type of measurements</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td>No baffle:</td>
<td>[0.48; 0.60]</td>
<td>Standard deviation</td>
<td></td>
</tr>
<tr>
<td>Baffle:</td>
<td>[0.48; 0.55]</td>
<td>of local pressure</td>
<td>Botterill et al. (1982), Lucas et al. (1986)</td>
</tr>
<tr>
<td>No baffle:</td>
<td>[0.43; 0.60]</td>
<td>Linear Average of e</td>
<td>Bhowmick et al. (2015), Pant et al. (2017)</td>
</tr>
<tr>
<td>Baffle:</td>
<td>[0.48; 0.60]</td>
<td>bed with radiation</td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>transmission</td>
<td></td>
</tr>
</tbody>
</table>

**Table 2-7. Transition bubbling/turbulent from literature correlations**

<table>
<thead>
<tr>
<th>Transition bubbling/turbulent (m·s⁻¹)</th>
<th>Correlation values</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0.621</td>
<td>Seo and Park (1988)</td>
</tr>
<tr>
<td></td>
<td>0.643</td>
<td>Chehbouni et al. (1994), Chehbouni et al. (1995)</td>
</tr>
<tr>
<td></td>
<td>0.591</td>
<td>Bi et al. (2000)</td>
</tr>
<tr>
<td></td>
<td>0.645</td>
<td>Abba (2001)</td>
</tr>
</tbody>
</table>

The comparison of Tables 2-6 and 2-7 shows a relative agreement between measurements, both from pressure and radiation transmission, of the transition velocity and values given by literature correlations. This also agrees with the transition velocity experimentally measured during experiments aiming at eliminating residual slugging (presented in Appendix J: Reduction of residual slugging in the experimental unit).

It should also be noted that the addition of a baffle tends to increase the transition velocity.
Figure 2-28. Transition bubbling/turbulent (measured & from correlations) as a function of bed height, for a bed with/without baffle

a) No baffle, b) Baffle at Bank B
Figure 2-28 presents the measured height of the bubbling/turbulent transition in the experimental unit for a bed used with and without a baffle. The figure includes both the experimental values and the values from the literature correlations. It should be noted that the upper half of the bed is in a turbulent regime (> [0.6; 0.8] m). As seen in Figure 2-24, this region has the highest bed shear, and therefore this is where the agglomerates are the most likely to break. Therefore, the previous assumption of using a constant bubble diameter $d_b$ typical of a turbulent bed for bed shear calculations is reasonable as a first approximation.

Since a similar $u_G$ profile (with the same $u_{G,\text{Bottom}}$, $u_{G,\text{TOP}}$ and velocity transition profile) is used between the experimental unit and the commercial Fluid Coker, with the same solids, a similar transition velocity is expected. This was verified by using literature correlations: they gave similar transition velocities for both the experimental unit and commercial fluid Coker conditions.

### 2.5.5 Scaling summary

This section is a * summary of the procedures for the agglomerate breakage scaling between the experimental unit and the commercial Fluid Coker

First, for the wet-agglomerates time since production (or last re-wetting if considered), the dimensionless time ($t/\tau$) defined in Section 2.2.2 is used. Using Equation (2.5), this allows for the computation of the "time in Fluid Coker" $t'$ (calculated from the recorded "time in the pilot" $t$) representative of the age of a wet-agglomerate in a commercial Fluid Coker.

Second, for the wet-agglomerates position, the scaling procedure presented in Section 2.2.1 is used. This implies a similar dimensionless radial location ($r/R$) and a similar dimensionless radius to bed height ($r/h$) between the experimental unit and the commercial Fluid Coker.

Third, for the bed hydrodynamics, since a similar $u_G$ profile (same $u_{G,\text{Bottom}}$ & $u_{G,\text{TOP}}$) is used between the experimental unit and the commercial Fluid Coker, no scaling is required.
Regarding the other hydrodynamics parameters \((u_{mf}, d_b, \ldots)\), the properties relevant to commercial Fluid Coker are used, following the recommendations of Section 2.2.4.

Fourth, the initial properties of wet-agglomerates \((D_{Agg,0}, \rho_{Agg,0}, (L/S)_0)\) and the properties relative to their strength \((\rho_L, \mu_L, \ldots)\) are defined for a realistic representation of a commercial Fluid Coker, following the recommendations of Section 2.2.3.

### 2.6 Interaction of wet-agglomerates with the vapours in the gas phase

In Fluid Cokers, wet-agglomerates are in constant contact with vapours in the gas phase. There are two types of locations of vapour generation. First, fixed locations near the injection zones, where feed liquids are injected and quickly vaporize. Second, variable locations, at each active wet-agglomerate, where liquid trapped inside is slowly cracked and vaporized.

The local concentration of vapour can impact the rate at which the product liquid can evaporate from wet-agglomerates. This mechanism is not considered in the shrinking core model.

#### 2.6.1 Vapour saturation

As hydrocarbons are cracked and vaporized in a commercial Fluid Coker, they could locally accumulate and create new mass transfer limitations in the agglomerate drying mechanisms. This potential limitation is not included in the current model but could represent interesting future development.

A first step would be to only consider zones in the bed with local vapour saturation. It would require the measurement of at least four experimental parameters: 1) bubble-emulsion exchange rate, 2) radial mixing of bubble gas, 3) mixing of emulsion, 4) local bubble flux
2.7 Agglomerate trajectories: experimental measurements

As mentioned previously, the model requires an important input: agglomerate trajectories. Moreover, to ensure that the model produces relevant results, these agglomerates trajectories from production zones to stripper need to be accurate.

This research used experimental measurements to obtain agglomerates trajectories using a gas-solid system (the experimental setup is described in detail in Chapter 3). The method used Radioactive Particle Tracking (RPT) with a single radioactive tracer. In the model, two virtual "states" can be attributed to each position of the experimental tracer-agglomerate trajectories: 1) "wet" when the simulated wet-agglomerate carries liquid (after production or re-wetting, before complete drying or breakage), 2) "dry" when the simulated wet-agglomerate does not carry liquid (after complete drying or breakage, and before the trajectories enter a new production zone).

"Wet" simulated agglomerates are subject to drying and possible breakage. They can be re-wetted and can carry liquid to the stripper zone. The model ignores "Dry" simulated agglomerates until they become "Wet" again by going through a production zone.

Multiple trajectories are required to obtain a population of trajectories suitable to represent a realistic average behaviour of wet-agglomerates. Preliminary investigations showed that a population of several thousand or more was required. This number was usually achieved by runs lasting at least twelve hours.

There are several other possible methods to obtain agglomerates trajectories that would be suitable with the model presented:

- Using the same principle with other types of single tracer measurements (magnetic tracer, RFID, …)
- Using imaging method (Xray, ECT, …)
- Using CFD simulations of agglomerate trajectories. This would require the careful selection of the CFD model and equations used to ensure accurate predictions.
A flowchart presenting an overall view of the model used with agglomerate trajectories acquired with RPT measurements is shown in Figure 2-29.

**Figure 2-29.** Summary of liquid losses model, with agglomerate trajectories acquired with RPT measurements.

### 2.8 Example of results from model

This section presents some results from the model. Liquid losses predictions obtained with only some components of the model active are compared.
The experimental unit used is presented in Chapter 3. The five lateral injection banks were active. The nozzle penetration $L_N$ was equal to 0. One baffle with no flux-tubes was located at Bank B. The solids recirculation flowrate was set to $F_S = 0.45 \text{ kg} \cdot \text{s}^{-1}$ and the bed mass was $m_S = 58.9 \text{ kg}$.

A new parameter is introduced: the simplified initial agglomerate yield strength $B_{\text{Agg},0}$. This parameter combines the agglomerate yield strength $\tau_{\text{Agg}} = f(\varepsilon_{\text{Agg}}, S_{\text{Agg}})$ defined in Section 2.5.1.1 and the critical agglomerate yield strength $\tau_{\text{Agg}}^* = f(D_{\text{Agg}}^2)$ defined in Section 2.5.1.2 to obtain a simple parameter able to estimate the probability of an agglomerate produced to survive breakage. It is defined as follow:

$$B_{\text{Agg},0} = 1.0 \times 10^{-4} \left( \frac{1 - \varepsilon_{\text{Agg},0}}{\varepsilon_{\text{Agg},0}} \right) \frac{S_{\text{Agg},0}}{D_{\text{Agg}}^2} \quad (2.36)$$

$B_{\text{Agg},0}$ should only be used with a given solid, liquid, and bed operating conditions (T, P, ...). In addition, the effect of the agglomerate trajectories, and therefore the level of bed shear encountered, is not captured by $B_{\text{Agg},0}$.

### 2.8.1 Comparison of model results

The liquid losses predicted by the model are compared when: 1) Drying is the only component considered, 2) Drying & re-wetting are the only components considered, 3) Drying and the bed shear (breakage) are the only components considered, 4) Drying, re-wetting, and the bed shear (breakage) are all considered.

Figure 2-30 presents the fraction of wet-agglomerates formed (in the whole bed) that reach the stripper zone with liquid for various initial agglomerates properties.

Figure 2-31 shows the fraction of liquid carried by agglomerates towards the stripper (for the whole bed), for various initial agglomerates properties. It shows the effect of the initial liquid capture mentioned in Section 2.3.2 by showing separately: the fraction of liquid initially in wet-agglomerates reaching the stripper zone (sub-chart a)), the fraction of liquid injected captured by produced wet-agglomerates (sub-chart b)), and the fraction of liquid injected reaching the stripper zone after being carried by wet-agglomerates (sub-chart c)).
Figure 2-30. Fraction of wet-agglomerates formed (for the whole bed) reaching the stripper zone with liquid, for various initial agglomerates properties (Short Taper, Baffle @ B, No Flux-Tubes, $L_{\text{Nozzle}} = 0 \text{ cm}$, $L_{\text{jet}} = 9 \text{ cm}$, $u_G \in [0.3; 0.9] \text{ m}\cdot\text{s}^{-1}$, $F_S = 0.45 \text{ kg}\cdot\text{s}^{-1}$, $m_S = 58.9 \text{ kg}$)

In Figure 2-31-c), the comparison shows that, for small to medium agglomerates ($D_{\text{Agg}} < 1 \text{ cm}$, $(L/S)_0 < 0.31$), the use of a drying-only model would lead to an over-estimation of the predicted liquid losses. For medium to large agglomerates ($D_{\text{Agg}} > 1 \text{ cm}$, $(L/S)_0 > 0.31$), using a drying-only model would lead to an under-estimation of the predicted liquid losses. It should be noted that for the largest agglomerates tested ($D_{\text{Agg}} = 1.5 \text{ cm}$, $(L/S)_0 = 0.62$), the drying-only model under-estimation disappears, suggesting that for larger agglomerates, using a drying-only model would again overestimate the predicted liquid losses.
a) Fraction of liquid initially in wet-agglomerates reaching the stripper zone

\[ \rho_{Agg} = 1100 \, \text{kg} \cdot \text{m}^{-3} \]

Drying + re-wetting + bed shear

<table>
<thead>
<tr>
<th>( (L/S)_0 )</th>
<th>0.1</th>
<th>0.15</th>
<th>0.175</th>
<th>0.205</th>
<th>0.31</th>
<th>0.465</th>
<th>0.62</th>
</tr>
</thead>
<tbody>
<tr>
<td>( D_{Agg} ) (cm)</td>
<td>0.5</td>
<td>0.625</td>
<td>0.7</td>
<td>0.75</td>
<td>1</td>
<td>1.25</td>
<td>1.5</td>
</tr>
<tr>
<td>( t_c ) (s)</td>
<td>6.3</td>
<td>14.1</td>
<td>20.2</td>
<td>26.5</td>
<td>65.4</td>
<td>137.2</td>
<td>238.2</td>
</tr>
<tr>
<td>( Br_{Agg,0} )</td>
<td>2.981</td>
<td>2.194</td>
<td>1.808</td>
<td>1.611</td>
<td>0.906</td>
<td>0.534</td>
<td>0.334</td>
</tr>
</tbody>
</table>

b) Fraction of liquid injected captured by produced wet-agglomerates,

<table>
<thead>
<tr>
<th>( (L/S)_0 )</th>
<th>0.1</th>
<th>0.15</th>
<th>0.175</th>
<th>0.205</th>
<th>0.31</th>
<th>0.465</th>
<th>0.62</th>
</tr>
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<tr>
<td>( D_{Agg} ) (cm)</td>
<td>0.5</td>
<td>0.625</td>
<td>0.7</td>
<td>0.75</td>
<td>1</td>
<td>1.25</td>
<td>1.5</td>
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<tr>
<td>( t_c ) (s)</td>
<td>6.3</td>
<td>14.1</td>
<td>20.2</td>
<td>26.5</td>
<td>65.4</td>
<td>137.2</td>
<td>238.2</td>
</tr>
<tr>
<td>( Br_{Agg,0} )</td>
<td>2.981</td>
<td>2.194</td>
<td>1.808</td>
<td>1.611</td>
<td>0.906</td>
<td>0.534</td>
<td>0.334</td>
</tr>
</tbody>
</table>
c) Fraction of liquid carried by agglomerates towards the stripper (for the whole bed), for various initial agglomerates properties

(Short Taper, Baffle @ B, No Flux-Tubes, L\textsubscript{Nozzle} = 0 cm, L\textsubscript{Jet} = 9 cm,

\(u_G \in [0.3; 0.9] \text{ m} \cdot \text{s}^{-1}, F_S = 0.45 \text{ kg} \cdot \text{s}^{-1}, m_S = 58.9 \text{ kg}\))

\begin{itemize}
  \item [a)] Fraction of liquid initially in wet-agglomerates reaching the stripper zone,
  \item [b)] Fraction of liquid injected captured by produced wet-agglomerates,
  \item [c)] Fraction of liquid injected reaching the stripper zone after being carried by wet-agglomerates
\end{itemize}

In Figure 2-31-c), it is clear that adding the re-wetting component or the breakage due to bed shear (without the other) to the drying-only model significantly modifies the predicted liquid losses. It suggests a strong competing effect between these two components. While re-wetting affects any agglomerate size, the breakage due to bed shear is especially critical regarding larger agglomerates.

It is especially interesting to note that, when breakage due to bed shear is accounted for, the largest and wettest agglomerates (\(D_{\text{Agg,0}} = D_{\text{Agg}} = 1.5 \text{ cm}, (L/S)_0 = 0.62\)) are now longer the wet-agglomerates carrying the highest fraction of injected liquid to the stripper zone.
As described in Section 2.5.1.1.1, this is due to their lower strength. This makes these larger agglomerates more prone to breakage than smaller agglomerates, and therefore reduces their importance in the overall liquid losses. On the other hand, in drying-only or drying and re-wetting models, these larger and wetter agglomerates are always the ones carrying the highest fraction of injected liquid to the stripper zone.

Finally, in Figure 2-31-b), the effect of the agglomerate size on the fraction of liquid injected captured by produced wet-agglomerates is visible, lowering the liquid losses of the largest agglomerates.

2.8.2 Comparison of the importance of drying vs. the importance of breakage due to bed shear

This section briefly investigates the relative importance of the two mechanisms contributing to agglomerate removal: drying and breakage due to bed shear. With the thermal model used (Shrinking Core Model), drying is only controlled by the time required for full drying, $t_C$. The breakage from bed shear is controlled, for constant bed conditions, by the agglomerate strength, $\tau_{\text{Agg}}$. The simplified initial agglomerate strength, $\text{Br}_{\text{Agg},0}$ (defined at the beginning of Section 2.8), can be used instead.

The impact of these two critical variables, $t_C$ and $\text{Br}_{\text{Agg},0}$, on liquid losses predicted by the model results is analyzed.

2.8.2.1 Time for complete drying $t_C$

Figure 2-32 shows that at a constant $\text{Br}_{\text{Agg},0}$, when $t_C$ increases, more liquid reaches the stripper. This result is because a higher $t_C$ means a longer time required to dry and, therefore, more liquid carried to the stripper zone for the same travel time. In addition, wetter agglomerates resist better against breakage from bed shear.
Figure 2-32. Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, with drying, re-wetting and bed shear equations, for various values of time required for full drying $t_C$ (at a constant $Br_{Agg,0}$)

(Short Taper, Baffle @ B, No Flux-Tubes, $L_{Nozzle} = 0$ cm, $L_{jet} = 9$ cm,

$u_G \in [0.3; 0.9]$ m·s$^{-1}$, $F_S = 0.45$ kg·s$^{-1}$)

2.8.2.2 Simplified initial agglomerate yield strength $Br_{Agg,0}$

Figure 2-33 shows that at a constant $t_C$, when $Br_{Agg,0}$ increases, more liquid reaches the stripper. This result is because a higher $Br_{Agg,0}$ means a stronger agglomerate and therefore less breakage.
Figure 2-33. Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, with drying, re-wetting and bed shear equations, for various values of simplified initial agglomerate yield strength $B_{rAgg,0}$ (at a constant $t_C$)

(Short Taper, Baffle @ B, No Flux-Tubes, $L_{Nozzle}$ = 0 cm, $L_{jet}$ = 9 cm,

$u_G \in [0.3; 0.9] \text{ m·s}^{-1}$, $F_S = 0.45 \text{ kg·s}^{-1}$)

2.9 Impact of values uncertainties on the model parameters

This section presents a brief study of critical model parameters. Table 2-8 summarizes these relevant parameters.

| Table 2-8. Model parameters requiring tests, in each section of the model |
| --- | --- | --- |
| **Section of the model** | **Parameter** | **Range** |
| **Drying** | Feedstock and vapour properties: $\gamma_L, \rho_L, \mu_L, C_{pL}, \rho_G, C_{pG}, \Delta H_{liq}, ...$ | **See Table 2-2** & **Table 2-3** |
| | Bed temperature $T_{Bed}$ | [530; 560] |
| | | **550** |
| **Re-wetting** | Coefficient of re-wetting $C_{Rewet}$ | [1; 2] |
| | | **1** |
Shear Breakage

<table>
<thead>
<tr>
<th></th>
<th>Feedstock and vapour properties: $\gamma_L$, $\rho_L$, $\mu_L$, $C_p_L$, $\rho_G$, $C_p_G$, $\Delta H_{liq}$, ...</th>
<th>See Table 2-2 &amp; Table 2-3</th>
</tr>
</thead>
<tbody>
<tr>
<td>Bed temperature $T_{Bed}$</td>
<td>[530; 560]</td>
<td>550</td>
</tr>
<tr>
<td>Empirical coefficient $M$</td>
<td>[6; 8]</td>
<td>7</td>
</tr>
<tr>
<td>Bubble diameter $d_b$ (m)</td>
<td>$[4.0 \times 10^{-2}; 7.5 \times 10^{-2}]$</td>
<td>$4.0 \times 10^{-2}$</td>
</tr>
</tbody>
</table>

2.9.1.1 Feedstock and vapour properties & bed temperature $T_{Bed}$

Adjusting the feedstock will require adjusting the feedstock properties used in both the drying and shear breakage section of the model. Since hydrocarbon vapours are generated from the feedstock, vapour properties would also need to be updated.

Similarly, adjusting the bed temperature $T_{Bed}$ would modify both the feedstock and vapour properties used in both the drying and shear breakage section of the model.

2.9.1.2 Coefficient of re-wetting $C_{Rewet}$

As mentioned in Section 2.3.3, the constant value of the coefficient of re-wetting $C_{Rewet}$ used is conservative.

Indeed, the model considers that any wet-agglomerate re-entering a production zone is re-wetted, with the same coefficient of re-wetting. In a real system, not all the agglomerates are rewetted with the same coefficient of re-wetting. This ratio ranges from 1 to 2 based on experimental work by Pardo Reyes (2015). In addition, some agglomerates might end up breaking due to their interaction with the jet: instead of being re-wetted, they are destroyed. Therefore, using a constant value of $C_{Rewet} = 1$ is a safe side assumption that limits a possible overestimation of the re-wetting effect.

More experimental work is needed to obtain a more accurate depiction of the re-wetting mechanism.
2.9.1.3 Empirical coefficient M

This section shows the effect of empirical coefficient M, used in Equation (2.26), on liquid losses predicted by the model.

Figure 2-34 shows the fraction of liquid carried by agglomerates formed at the bank i towards the stripper, for various values of empirical coefficient M. The empirical coefficient M has only a moderate impact on predicted liquid losses. It creates small variations for M < 7 (down to - 10 %) and moderate for M > 7 (up to + 29 %).

![Diagram showing liquid transfer fractions](image)

**Figure 2-34.** Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, for various values of empirical coefficient M

(Short Taper, Baffle @ B, No Flux-Tubes, L_{Nozzle} = 0 cm, L_{jet} = 9 cm,
\[ u_G \in [0.3; 0.9] \text{ m}\cdot\text{s}^{-1}, F_S = 0.45 \text{ kg}\cdot\text{s}^{-1} \])

2.9.1.4 Bubble diameter \( d_b \)

This section shows the effect of bubble diameter \( d_b \) on liquid losses predicted by the model.

Figure 2-35 shows the fraction of liquid carried by agglomerates formed at Bank i towards the stripper zone for various values of bubble diameter \( d_b \). Three sizes of wet-agglomerates
are considered. It is observed that the bubble diameter, \( d_b \), can significantly impact model results. For each initial agglomerate properties, there is a critical bubble size related to strong variations in model results. This result is due to the shape of \( \tau_{\text{Agg}} = f(S_{\text{Agg}}) \) curves presented in Section 2.5.1.1.

For the range of bubble diameters ([2.0; 7.0] cm) found in the literature for conditions relevant to a commercial Fluid Coker:

- For small and dry agglomerates (\( D_{\text{Agg}} = 0.5 \) cm & \( (L/S)_0 = 0.10 \)): no effect (dried anyway)
- For medium agglomerates (\( D_{\text{Agg}} = 1.0 \) cm & \( (L/S)_0 = 0.31 \)): when \( d_b \) becomes smaller than 4-5 cm, the destruction rates increase significantly, and the liquid losses plummet.
- For big and wet-agglomerates (\( D_{\text{Agg}} = 1.5 \) cm & \( (L/S)_0 = 0.62 \)): when \( d_b \) becomes smaller than 4 cm, the destruction rates increase significantly, and the liquid losses plummet.

These values are used as a baseline in further analysis: an initial analysis is conducted, and the liquid losses are predicted for a \( d_b \) equal to 4 cm. This is a safe side assumption as it allows for a value suitable for both bubbling and turbulent bubble sizes found in the literature. Then, for promising results, a second analysis is conducted with a bubble diameter \( d_b \) equal to 5 cm. This second analysis allows checking against a possible underestimation of liquid losses for medium and large agglomerates.
**a)**

Liquid injected transferred to generated agglomerates

Liquid transferred from generated agglomerate to stripper

Liquid injected transferred to stripper (through agglomerates)

Typical bubble size in commercial Fluid Coker

\[ \rho_{Agg} = 1100 \text{ kg} \cdot \text{m}^{-3} \]

\[ (L/S)_0 = 0.10 \]

\[ D_{Agg} = 0.5 \text{ cm} \]

**b)**

Liquid injected transferred to generated agglomerates

Liquid transferred from generated agglomerate to stripper

Liquid injected transferred to stripper (through agglomerates)

Typical bubble size in commercial Fluid Coker

\[ \rho_{Agg} = 1100 \text{ kg} \cdot \text{m}^{-3} \]

\[ (L/S)_0 = 0.31 \]

\[ D_{Agg} = 1 \text{ cm} \]
Figure 2-35. Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, for various values of bubble diameter $d_b$

(Short Taper, Baffle @ B, No Flux-Tubes, $L_{Nozzle} = 0$ cm, $L_{jet} = 9$ cm,

$u_G \in [0.3; 0.9]$ m·s$^{-1}$, $F_s = 0.45$ kg·s$^{-1}$)

a) “Small” agglomerate, b) “Medium” agglomerate, c) “Big” agglomerate
2.10 References


Chapter 3

3 Experimental setup & methods

Experimental measurements are required in a newly designed scaled-down, gas-solid tapered cylindrical fluidized bed with solids recirculation to achieve the research objectives and predict product losses towards the bottom of commercial Fluid Cokers. This unit has lateral feed injections, grouped in banks of nozzles located around the periphery at several heights, to mimic the steam-bitumen bank injections of commercial Fluid Cokers. The experiments are conducted at room temperature (“cold” setup). The test material used is fluid coke provided by Syncrude Canada Ltd. The fluid coke used has a Sauter mean diameter of 140 µm, measured by a HELOS particle size analyzer. The measurements are available in Appendix A: Fluid coke particle sized distribution. This fluid coke has a particle density of 1470 kg·m⁻³ (Mohagheghi Dar Ranji, 2015). Fluid coke particles can be classified as type B particles, near the boundary between type A and B particles in the Geldart classification (Geldart, 1973; Song et al., 2006). The fluidization gas is compressed air at room temperature, coming from the Institute for Chemicals and Fuels from Alternative Resources (ICFAR) compressors. At room temperature (20 °C), the air used has a relative humidity lower than 5 % (Li et al., 2020). Lateral gas-liquid spray banks used in commercial Fluid Cokers are simulated with banks of pure gas injectors whose penetration into the bed can be adjusted. In addition, internals, such as ring baffles, can be added and removed from the unit.

Most experiments use a radioactive (⁴⁶Sc) tracer and twelve radiation sensors for Radioactive Particle Tracking (RPT) and radiation transmission experiments. Validation experiments use dyed coke. Additional experiments use liquid tracer injection, such as Varsol™, to study liquid & vapour carry-under and capacitance probes.
3.1 New tapered recirculating fluidized bed

A new experimental scaled-down, tapered, gas-solid cold fluidized bed, shown in Figure 3-1, was designed in collaboration with the industrial partners to simulate commercial Fluid Coker hydrodynamics according to these criteria:

- A fixed bed height to bottom diameter ratio of 6.33.
- A superficial gas velocity at the sparger level (in the lowest bed region) of 1 ft·s\(^{-1}\) (0.3048 m·s\(^{-1}\)).
- A superficial gas velocity at the bed surface of 3 ft·s\(^{-1}\) (0.914 m·s\(^{-1}\)). The additional gas flow is provided by air injection through lateral injection nozzles arranged in ring banks distributed over the bed height. These lateral air injections aim to simulate two phenomena resulting from bitumen-steam injections in commercial Fluid Cokers: the flow of atomization steam and the flow of vapours evolving from the thermal cracking of the injected bitumen (Song et al., 2006).

The sparger (item 1 in Figure 3-1) and stripper (item 2 in Figure 3-1) section of the unit are common to the two configurations. The sparger flow is controlled using an orifice plate (4 mm ID) equipped with a high-accuracy pressure gauge (0-145 psi, 0.1 psi resolution, McMaster-Carr) coupled to a pressure regulator (± 3 psi accuracy, McMaster-Carr), ensuring a stable gas flowrate. The top freeboard section (item 3 in Figure 3-1) reduces solids losses by increasing the cross-sectional area. The lateral injection system (item 4 in Figure 3-1), as described previously, has been represented in its totality in Figure 3-1 for the lower bank of injection (level A), then simplified for the other levels for the sake of clarity. At the top of the bed (item 5 in Figure 3-1), the gas outlet is connected to a cyclone to recover entrained particles returned to the bed through a dipleg. The cyclone exhaust is connected to a barrel equipped with a screen to ensure that radioactive particles cannot escape to the environment; this also allows for easy monitoring of solids losses from the unit.
Figure 3-1. Design drawing of the experimental unit
Fluid Coking™ is a process in which solids circulate between the reactor and the burner (Gray, 1994). Therefore, the new lab-scale tapered cold flow recirculating fluidized bed includes a standpipe and a riser to reproduce this solids recirculation. This recirculation is carried by pneumatic transport. The pneumatic gas flow is controlled using an orifice plate (7 mm ID) coupled to pressure regulators, ensuring a stable gas flowrate (item 6 in Figure 3-1). A pinch valve from EVR (Sudbury, ON), located at the bottom of the fluidized bed, controls the solids circulation flow. The solids flowrate is measured with a calibrated elbow flowmeter (item 7 in Figure 3-1) using the same method as Sanchez Careaga (2013). To prevent arching and mitigate pulsations in the solids flowrate, the small section of the bed located between the sparger and the pinch valve is fluidized by a complementary very low flowrate injection system (item 8 in Figure 3-1). The pneumatic riser discharges solids tangentially into the column freeboard, 1 m below the column top (item 9 in Figure 3-1). The tangential entry induces solids spinning, minimizing coke losses to the cyclone and ensuring uniform solids dispersion when falling back into the bed. Several taps (item 10 in Figure 3-1) are distributed over the bed height for pressure drop measurements. Finally, another tap is located just above the pinch valve (item 11 in Figure 3-1). This tap allows either solids screw sampling or capacitance measurements.

3.1.1 Tapered bed design

In commercial Fluid Cokers, the average superficial velocity of the rising gases and vapours increases from 0.3 m·s⁻¹ to 0.9 m·s⁻¹ due to the distribution of the steam-bitumen injection points over the bed height and the cumulative contribution of the vapours created by cracking. Consequently, many Fluid Cokers are tapered to mitigate this increase of the superficial velocity with elevation.

As discussed in Chapter 1 – Section 1.1.3, this tapered section requires special attention. It was experimentally observed that it creates a heterogenous bed structure, with a well-fluidized core and a less than well-fluidized annulus.
3.1.2 Gas bed injection

As shown in Figure 3-1, there are two primary sources of gas injection into the bed: a sparger, located near the bottom of the bed to fluidize the whole bed, and five banks of eight lateral injectors, used to simulate the gas and vapours associated with steam-bitumen injections in commercial Fluid Coker.

3.1.2.1 Sparger

![Design of experimental sparger (bottom view)](image)

**Figure 3-2.** Design of experimental sparger (bottom view)

The sparger supplies compressed air at the bottom of the bed to fluidize the bed inside the reactor (the axial position is visible in Figure 3-1). The sparger was initially designed according to the guidelines presented by Kunii and Levenspiel (1991) and consists of two loops (one internal and one external) to equalize the pressure within the sparger and reduce gas distribution issues (Figure 3-2). The internal loop was modified to use only seven 1.59 mm diameter holes per side (for a total of 28 holes), and the sparger injection was switched to a downward injection to minimize residual slugging for the range of fluidization velocities targeted (see Appendix J: Reduction of residual slugging in the
experimental unit). This modification was made to minimize residual slugging. The preliminary experiments conducted and the results obtained are presented in Appendix J: Reduction of residual slugging in the experimental unit. The sparger design is visible in Figure 3-2. The sparger flow is controlled using an orifice plate (4 mm ID) coupled to pressure regulators, ensuring a stable gas flowrate. In a typical run, the sparger is used to reach a cross-sectional fluidization velocity of 0.30 m·s⁻¹. This value is significantly higher than the minimum fluidization velocity calculated as 0.01 m·s⁻¹ (Wen and Yu, 1966; Richardson and da S. Jerónimo, 1979) and therefore eliminates the risk of an occurrence of defluidized zones within the bed, even when the cross-sectional superficial gas velocity decreases in the tapered section.

3.1.2.2 Adjustable lateral injectors

The experimental unit incorporates lateral air injections to simulate bitumen-steam injections and the vapours evolving from the bitumen thermal cracking within the commercial Fluid Coker. This section discusses the injector design and specificities.

3.1.2.2.1 Nozzle design

The nozzles used in the experimental unit only inject gas (air at room temperature). Therefore, to simulate the impact of steam-bitumen sprays on bed hydrodynamics, the parameters needed to represent the commercial Fluid Cokers are: the dimensionless jet penetration (ratio of jet penetration length to bed diameter, at the injection location) and cross-sectional average gas velocity above a given injection bank.

The type of nozzle most commonly used in commercial Fluid Coker to atomize bitumen droplets and distribute them into the fluidized bed is the GEN3 nozzle (Chan et al., 2015; McMillan et al., 2019), the successor of the “Terence E. Base” (TEB) nozzle (Base et al., 1999; Bennet et al., 2001). The nozzle diameter at the nozzle throat is 1.3 cm ID (Prociw et al., 2018). The range of nozzle penetrations and jet penetrations observed in a typical Fluid Coker are given in Table 3-1 (Sanchez Careaga et al., 2016).
In this study, each feed nozzle is simulated with a stainless-steel tube of constant inner diameter (ID) of 0.85 cm. The nozzle is shown in Figure 3-3.

![Figure 3-3. Nozzle used in the experimental unit](Picture by Yohann Cochet, 2021)

### 3.1.2.2.2 Jet penetration & radial nozzle tip position

In commercial Fluid Cokers, high-velocity gas-liquid jets transport gas, liquid, and entrained solids from the annulus to the central core. Interactions between these jets from different nozzles should generally be avoided (Wormsbecker et al., 2016). The experimental nozzle diameter was selected to achieve a relative jet penetration (defined as the ratio of the jet length by the bed diameter, at the injection location) similar between the experimental unit and the commercial Fluid Cokers (Wormsbecker, McMillan, et al., 2021). Literature correlations (Zenz, 1968; Merry, 1971; Yates et al., 1986; Roach, 1993; Benjelloun et al., 1995) were used to find an experimental nozzle diameter that would give a similar relative gas jet penetration in the experimental unit. The selected diameter was chosen by averaging the value obtained with the most relevant correlations for the conditions studied (Yates et al., 1986; Benjelloun et al., 1995). The experimental gas jet penetration was then verified by experimental measurements such as RPT (see Section 3.3) and E-probes (see Section 3.5). The range of jet penetration observed in the experimental unit is presented in Table 3-1. It should be noted that the range obtained matches the dimensionless jet penetration in commercial Fluid Coker.

Besides, in a commercial Fluid Coker, spray nozzles extend moderately beyond the Fluid Coker wall (Koeninger et al., 2017; Koeninger et al., 2018). Figure 3-4 shows a depiction
of a feed nozzle installed on the vessel wall and outlines the definition of nozzle penetration and jet penetration.

During commercial Fluid Coker operations, fouling occurs on the reactor wall, causing the diameter of the reactor to decrease throughout the run. The spray nozzle positions being kept constant means that the effective nozzle penetration (from the column inner wall to the nozzle tip) decreases over time. This evolution should not significantly impact the jet penetration (from the nozzle tip to the jet tip). Typical nozzle and jet penetrations in commercial Fluid Cokers are presented in Table 3-1.

![Diagram of nozzle penetration and jet penetration](image)

**Figure 3-4.** Nozzle penetration & jet penetration

The experimental unit is equipped with lateral nozzles with adjustable penetration. Therefore, different nozzle penetrations can be investigated to study the impact of this coke build-up near the spray nozzles. In addition, the experimental jet penetration can be adjusted by changing the gas flowrate using the pressure regulators and an orifice plate used for gas regulation in each nozzle. The jet and nozzle penetrations investigated in the experimental unit are presented in Table 3-1.
### Table 3-1. Comparison of commercial & experimental nozzle and jet penetrations

<table>
<thead>
<tr>
<th></th>
<th>Commercial Fluid Coker (Wormsbecker, McMillan, et al., 2021)</th>
<th>Experimental bed</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Top bed diameter</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$D_{\text{Bed,Top}}$ (m)</td>
<td><strong>Beginning of run</strong></td>
<td><strong>End of run</strong></td>
</tr>
<tr>
<td></td>
<td>11</td>
<td>0.25</td>
</tr>
<tr>
<td></td>
<td><strong>End of run</strong></td>
<td>9.2</td>
</tr>
<tr>
<td><strong>Nozzle penetration</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$L_N$ (m)</td>
<td><strong>Beginning of run</strong></td>
<td><strong>End of run</strong></td>
</tr>
<tr>
<td></td>
<td>0.9</td>
<td>0.03</td>
</tr>
<tr>
<td></td>
<td>0.02</td>
<td>0.01</td>
</tr>
<tr>
<td></td>
<td>0.01</td>
<td>0</td>
</tr>
<tr>
<td><strong>Jet penetration</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>(from nozzle tip)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$L_J$ (m)</td>
<td>2</td>
<td>0.03 – 0.05</td>
</tr>
<tr>
<td><strong>Nozzle penetration</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>over bed radius</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$L_N/R_{\text{Bed,Top}}$</td>
<td><strong>Beginning of run</strong></td>
<td><strong>End of run</strong></td>
</tr>
<tr>
<td></td>
<td>16 %</td>
<td>0%</td>
</tr>
<tr>
<td></td>
<td>16 %</td>
<td>8 %</td>
</tr>
<tr>
<td></td>
<td>0%</td>
<td>0%</td>
</tr>
<tr>
<td><strong>Jet penetration</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>(from nozzle tip)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>over bed radius</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$L_J/R_{\text{Bed,Top}}$</td>
<td>36 %</td>
<td>24 % - 40 %</td>
</tr>
<tr>
<td><strong>Jet penetration</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>(from wall)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>over bed radius</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$(L_N + L_J)/R_{\text{Bed,Top}}$</td>
<td><strong>Beginning of run</strong></td>
<td><strong>End of run</strong></td>
</tr>
<tr>
<td></td>
<td>53 %</td>
<td>44 %</td>
</tr>
<tr>
<td></td>
<td>40 % - 64%</td>
<td>24 % - 48 %</td>
</tr>
</tbody>
</table>

#### 3.1.2.2.3 Number & locations

A typical Fluid Coker can have 50 to 100 spray nozzles arranged in 5 to 8 banks at different heights, with more nozzles in the upper rings (Wormsbecker et al., 2016). The injectors in the experimental unit of this research are arranged in 5 horizontal banks at different heights. Each bank consists of 8 injectors distributed evenly around the perimeter of the fluidized bed. The injection banks and nozzles locations are represented in Figure 3-1. Each injector
flowrate is independently controlled using a sonic nozzle (1.3-2.7 mm ID) coupled to pressure regulators constructed following the recommendations by McCabe et al. (1993). This system allows for a stable gas flowrate independent of the downstream pressure.

Due to the combination of tapered shape and lateral injections, the cross-sectional superficial gas velocity varies over the bed height. These variations are presented in Figure 3-5. Due to this velocity variation, the regime transition correlations (Seo and Park, 1988; Bi et al., 2000; Abba, 2001) indicate that the bottom of the experimental unit is in a bubbling regime while the top of the unit is in a turbulent regime. This bubbling-turbulent transition was verified in this research by measurements of pressure and radiation transmission measurements, shown in Chapter 2 - Section 2.5.4.

![Figure 3-5. Diameter & cross-sectional gas velocity profiles](image)

### 3.1.3 Solids recirculation

A critical operating parameter of a commercial Fluid Coker is its solids recirculation flowrate between the reactor and the burner. With their significant volume, commercial Fluid Cokers recirculate solids with a solids flux in the stripper in the range of
With the experimental unit used in this research, this corresponds to a solids flowrate $F_s$ ranging from 0.50 to 0.65 kg·s$^{-1}$ and a mean solids residence time in the reactor of around 15 min. The much smaller volume of the experimental units leads to one of the following scenarios:

- Solids recirculation flowrate of 0.55 kg·s$^{-1}$, but a solids mean residence time of around 1 min
- Solids mean residence time of around 15 min, but a solids recirculation flowrate of around $10^{-2}$ kg·s$^{-1}$

This research favoured the first scenario: a realistic solids recirculation flux to achieve bed hydrodynamics similar to a commercial Fluid Coker. Overall, solids mixing patterns scale with the column diameter because of the bed core-annulus structure and do not scale with bubble diameter (Glicksman et al., 1994; Xing, 2020). The ratio of the mean solids residence time in a typical commercial scale by the mean solids residence time in the pilot-scale experimental unit is used to scale up the time distributions measured experimentally. The scaling strategy is discussed in more detail in Chapter 2 – Section 2.2 & Appendix I: Spatial and Temporal scaling of the model.

In the experimental unit, solids recirculation is achieved through pneumatic transport. The pneumatic gas flow is controlled using an orifice plate (7 mm ID) coupled to pressure regulators, ensuring a stable gas flowrate. The orifice plate was operated in the sonic regime; since the pressure upstream of the orifice plate was kept constant by a pressure regulator, the mass flowrate of conveying gas remained constant during each run, independently of any downstream pressure variations in the conveying line. Calibration of this orifice plate is provided in Appendix B: Calibration of the solids recirculation line: gas and solids flowrate calibration. A pinch valve from EVR (Sudbury, ON), located at the bottom of the fluidized bed, controls the solids circulation flow. A calibrated elbow flowmeter measures the solids flowrate. This system consists of a set of pressure taps located in the bottom elbow of the recirculation line (see item 6 in Figure 3-1) connected to a pressure transducer (Omega PX16X). The solids flowrate $F_s$ calibration curve can be obtained as a function of the measured voltage $U_{PT}$ and the gas mass flowrate in the
recirculation line. The mass flowrate calibration procedure used a bucket and stopwatch procedure and is presented in Appendix B: Calibration of the solids recirculation line: gas and solids flowrate calibration. A calibration curve (for a gas mass flowrate \( F_G \) of \( 3.57 \times 10^{-2} \text{ kg} \cdot \text{s}^{-1} \), corresponding to a superficial gas velocity \( u_G \) of about 9.4 m·s\(^{-1}\) in the recirculation line) is presented in Figure 3-6. For a constant gas flowrate \( F_G \) and constant gas properties, the Law of Energy Conservation lead to the following relationship (Kuphaldt, 2019) between the mass flowrate in the elbow and the pressure drop signal, in the form of the response voltage \( U_{PT} \) of the pressure transducer:

\[
F_S = A \sqrt{U_{PT}}
\]  

\[\text{(3.1)}\]

![Figure 3-6. Solid recirculation calibration curve](image)

\( (\text{Gas mass flowrate } F_G = 3.57 \times 10^{-2} \text{ kg} \cdot \text{s}^{-1}) \)

It needs to be pointed out there that higher solids flowrate \( F_S \) (above 0.50 kg·s\(^{-1}\)) and lower solids flowrate \( F_S \) (below 0.25 kg·s\(^{-1}\)) were achieved using different gas mass flowrates \( F_G \); the corresponding calibration curves are shown in Appendix B: Calibration of the solids recirculation line: gas and solids flowrate calibration.
3.1.4 Removable ring baffle(s)

(a)

Figure 3-7. Picture of baffles
a) Industrial Fluid Coker baffle (Kamienski et al., 2013), b) Experimental baffle, with and without flux-tubes (Picture by Yohann Cochet, 2021)

Local bed conditions can be modified by adding ring baffle(s). For example, in Fluid Cokers, baffles are used to reduce the bypassing of wet solids and agglomerates (Wyatt Jr. et al., 2013; Guío-Pérez et al., 2014). Internals, such as baffles, are also used in the petroleum and chemicals industries to improve fluidization by breaking and re-distributing the bubbles (Chitnis et al., 2013; Issangya et al., 2013; Kamienski et al., 2013). Baffles can
also modify local bubble distribution (Jahanmiri, 2017) and solids mixing patterns in a fluidized bed (Sanchez Careaga, 2013). This bubble flow modification was found to greatly improved the liquid spray distribution into the fluidized bed above the baffle (Jahanmiri, 2017; Li et al., 2018). An industrial baffle used in a commercial Fluid Coker is shown in Figure 3-7-a). The baffles used in the experimental unit (with and without flux-tubes) are shown in Figure 3-7-b).

3.1.4.1 Geometry

The ring baffles used in this research have a design following industrial specifications (Wyatt Jr. et al., 2011). Based on previous research, an angle of 45 degrees (Sanchez Careaga, 2013) was used. Schematics of the baffles designed are presented in Figure 3-8.

![Diagram of ring baffle](image)

**Figure 3-8.** Side and top schematic view of baffle used in the experimental unit

The baffles were built with cold-rolled steel (The University Machine Shop, Western University). Their sizes are summarized in Table 3-2.
Table 3-2. Size of baffles used in each experimental unit configuration.

<table>
<thead>
<tr>
<th>Baffle reference</th>
<th>Baf_ST-1</th>
<th>Baf_ST-2</th>
<th>Baf_ST-3</th>
<th>Baf_ST-4</th>
</tr>
</thead>
<tbody>
<tr>
<td>Injection bank reference</td>
<td>B</td>
<td>C</td>
<td>D</td>
<td>E</td>
</tr>
<tr>
<td>Baffle location</td>
<td>$H_{top}$ (cm)</td>
<td>39</td>
<td>54</td>
<td>69</td>
</tr>
<tr>
<td></td>
<td>$H_{bottom}$ (cm)</td>
<td>34.6</td>
<td>49.6</td>
<td>64.6</td>
</tr>
<tr>
<td>$L_{baffle} = H_{baffle}$ (cm)</td>
<td></td>
<td></td>
<td></td>
<td>4.4</td>
</tr>
<tr>
<td>$D_{baffle,top} = D_{bed}$ (cm)</td>
<td></td>
<td></td>
<td></td>
<td>24.6</td>
</tr>
<tr>
<td>$D_{baffle,bottom}$ (cm)</td>
<td></td>
<td></td>
<td></td>
<td>15.7</td>
</tr>
<tr>
<td>$D_{downcomers}$ (cm)</td>
<td></td>
<td></td>
<td></td>
<td>2.34</td>
</tr>
<tr>
<td>Downcomers pipe (reference)</td>
<td>3/4 (1.050 OD / .065 wall thickness)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$N_{downcomers}$</td>
<td></td>
<td></td>
<td></td>
<td>20</td>
</tr>
<tr>
<td>Angle $\Phi$ ($^\circ$)</td>
<td></td>
<td></td>
<td></td>
<td>18.0</td>
</tr>
</tbody>
</table>

3.1.4.2 Number & positions of baffle(s)

The optimal ring baffles locations within the bed were determined in a previous research (Jahanmiri, 2017). The baffle is set just below any of the banks of lateral injector locations. The lowest bank was not investigated. Possible baffle(s) locations within the reactor injection zone are presented in Figure 3-9. Both the case of the addition of a single baffle and multiple baffles will be investigated in Chapter 6.

This research investigated the impact of baffle position for different injection patterns. It also investigated the use of multiple baffles.
3.1.4.3 Downcomers (or flux-tubes)

Equipping the baffles with downcomers that allow downward solids flow and upward gas flow improves baffle performance of commercial Fluid Coker (Wyatt Jr. et al., 2011). The exact mechanisms behind this improvement are not yet clearly identified.
For the experimental baffle, industrial specifications were followed regarding the open area repartition between the downcomers and the center part of the baffle, as shown in Figure 3-10. A constant angle between the downcomers and a constant downcomer diameter was used for all the baffles installed in the experimental unit.

This research investigated the impact of adding downcomers on ring baffles to identify the underlying mechanisms which could explain the commercial observations.

### 3.2 General bed characterization

#### 3.2.1 Pressure based measurements

The pressure transducers used with the experimental units were silicon pressure sensors (Honeywell ASDX Series) connected to and controlled by a microcontroller board (Arduino Uno/Mega). The measurements were conducted using taps located over the bed height. Snubbers (McMaster-Carr) were installed at each tap to avoid solids leaks into the pressure sensors. With this system, the local pressure drop is measured with a sampling frequency of 33 Hz.

The pressure signal was recorded and analyzed to:

- measure the local standard deviation of the pressure drop in the bed, and consequently, detect the transition from bubbling regime to turbulent regime (Johnsson et al., 2000; Ellis et al., 2004).
- measure the local pressure drop in the bed, and consequently, find the average cross-sectional voidage at various bed heights (van Ommen and Mudde, 2008).
- monitor the evolution of bed mass over time. The solids entrainment losses at typical operating conditions (for a cross-sectional superficial gas velocity at bed surface $u_{G,\text{Top}}$ of 0.9 m·s⁻¹) were measured to be between 0.16 and 0.23 kg·hour⁻¹.

#### 3.2.2 Bed level & bed mass

The defluidized bed level is measured using a laser distance measure tool (Tacklife HD60). This device uses laser reflection to measure the distance between a light source and the
closest opaque surface in front of the tool. Therefore, the distance between the experimental unit lid and the defluidized bed surface can be measured. However, it should be noted that a slow defluidization (i.e. closing the gas valve gradually over 10-30 seconds) is required to ensure an even bed level.

Once the defluidized bed height is obtained, the defluidized bed volume can be estimated based on each immersed bed section dimension. Then, the bed mass can also be obtained using the minimum emulsion density. This minimum fluidization density $\rho_{mf}$ can be obtained from aerated bulk density measurements by draining a given mass of powder from a special hopper into a graduated cylinder (Abrahamsen and Geldart, 1980). It was measured to be between 940 and 970 kg·m$^{-3}$, depending on the proportion of fines in the fluid coke. A calibration curve for bed mass measurements using the laser distance measure tool is shown in Figure 3-11.

![Figure 3-11. Bed mass using laser distance measure tool](image)

As mentioned in the previous section, the bed mass evolution was tracked using the pressure transducer. It should be noted that fresh fluid coke was re-added to the bed daily.
This was necessary to compensate for solids entrainment losses and maintain the solids population balance.

3.3 Radioactive Particle Tracking (RPT) & tracer agglomerate

Used in fluidized beds, the Radioactive Particle Tracking (RPT) technique uses emitted gamma rays (γ-rays) radiation, from one or more radioactive tracer-particle(s), to obtain the source-detector distances from an array of sensors located externally to the bed. The matrices of these distances are then solved using various minimization methods to calculate the source position inside the bed (Chaouki et al., 1997). By accumulating the recorded data over a few hours to a couple of days, time-averaged 3D position and velocities profiles are obtained.

The main advantage of RPT is its non-intrusive nature: data can be obtained without disrupting the gas-solid flow inside the bed, especially near the bed core. The main limitation of RPT is related to the fact that the number of counts detected by any scintillation detector for a given tracer position located inside the vessel can vary because of the random nature of the decay of an unstable atom. Since radioactive decay is described by a Poisson distribution (Leo, 1994), the counts rate measured CR_i (defined as the number of counts measured C_i at detector i during the time interval Δt_M) can be described by Equation (3.2) (Holbert, 2002):

\[ CR_i = \frac{C_i}{\Delta t_M} \pm \sigma_{Rad} = \frac{C_i}{\Delta t_M} \pm \sqrt{\frac{C_i}{\Delta t_M}} \]  

(3.2)

For this reason, a random error in the location of the tracer-agglomerate is always expected. Figure 3-12 shows the variation of the error fraction in the measured counts rate \( \sigma_{Rad}/CR_i \), as a function of the radioactive source strength and sampling time. The ranges of values used in this research are highlighted in green. It should be noted that since the radiation strength decreases with the inverse square of the distance, the impact of distance on the error fraction in the measured counts rate \( \sigma_{Rad}/CR_i \) can be extrapolated from Figure 3-12-a).
Besides, varying bed density (i.e. phase holdup) also adds random noise in the recorded signal (Chaouki et al., 1997).

**Figure 3-12.** Fraction of error in counts rate measured $\sigma_{\text{Rad}/CR_i}$

a) As a function of the source strength, b) as a function of the sampling time $\Delta t_M$.
3.3.1 Radioactive tracer

The radioactive tracer is composed of two distinct parts: the core, a radionuclide source of the $\gamma$-ray emissions, and an encapsulation of this core used to adjust the size and density of the tracer to represent the target particles accurately.

3.3.1.1 Scandium-46 core

For this research, radioactive scandium ($^{46}$Sc) has been preferred to radioactive gold ($^{198}$Au) for solid RPT experiments because of its reasonable decay time (84 days compared to 2.6 days). This longer decay time allowed more experiments over a long time while presenting very low health and safety concerns as it decays into stable Ti$^{46}$. The tracer was radiated in the Material Test Reactor at McMaster University. The decay scheme of $^{46}$Sc is shown in Figure 3-13.

![Figure 3-13](image-url)  
**Figure 3-13.** Decay scheme of $^{46}$Sc (Helmer and Kuzmenko, 2004)
3.3.1.2 Encapsulation

The tracer needs to be prepared in a way that ensures similar aerodynamic properties as coke/bitumen agglomerates, such as shape, size, and density. Regardless of the method or tracer preparation and the materials used, all suffer from similar limitations: the material is not the same as the fluidized medium, and small particles in the same size range as the fluid coke particles (100-300 μm) cannot be used as they do not provide enough radiation.

![Figure 3-14. Range of tracer-agglomerates size and density studied in this research](image)

Previous researchers have employed various methods to produce representative tracer-agglomerates:

- Khanna et al. (2008), Godfroy (1997), and Sanchez Careaga (2013) mixed epoxy resin with radioactive metal powder (Au$^{198}$) in such proportions that the tracer had the same particle density as the fluidized particles. After drying and hardening, the resulting resin was cut and shaped into a tracer of the desired size. The advantage is that, with small powder particles, the tracer is essentially homogeneous.
• Moslemian et al. (1992) coated small radioactive metal parts (\(^{46}\text{Sc}\)) with epoxy and encapsulated this core in a polyurethane shell to match the particle diameter and density.

• Chaouki et al. (1997) described other attempts to introduce material irradiated to produce a radioactive tracer.

Figure 3-15. Radioactive tracer with encapsulation (Picture by Yohann Cochet, 2021)

a) Uncoated scandium core (before irradiation),

b) Epoxy coated scandium core (before & after irradiation),

c) Epoxy/Glass bubbles coated scandium core (after irradiation),

d) Encapsulated size & density adjusted tracer.

This research used a hybrid approach. The scandium core was initially coated with a fine layer of epoxy to limit fines production for safety reasons. The epoxy-coated scandium core was then irradiated in a nuclear reactor (Material Test Reactor, McMaster University, Canada) and partially transformed into radioactive scandium \(^{46}\text{Sc}\). A second coating was then applied, with a mixture of epoxy and hollow glass spheres (Mia 65 Glass Bubbles, Miapoxy). Finally, the double-coated core was encapsulated in a polypropylene cargo. This cargo had a relatively large size, between 0.65 and 1.45 cm diameter, but allowed for precise adjustment of the tracer density, between 900 and 1200 kg.m\(^{-3}\). Figure 3-14 shows the range of agglomerate sizes and densities investigated in this research. The range of sizes and densities covered allows for studying the effect of wet-agglomerate size and
density of liquid losses and comparing liquid losses due to wet-agglomeration vs. liquid losses carried by solids emulsion and micro-agglomeration. Figure 3-15 is a picture of the tracer, with the uncoated scandium core, the epoxy-coated scandium core, and the encapsulated size and density adjusted tracer.

### 3.3.2 Scintillation Detectors

Scintillation detectors consist of a scintillator, a material containing impurities that luminesce when excited by ionizing radiation, and a photomultiplier (PMT). A PMT which absorbs a photon re-emits the energy captured as a measurable electric pulse due to the photoelectric effect (Leo, 1994).

![Figure 3-16. Scintillation detectors with integrated PMT (Saint-Gobain Crystals, 2014)](image)

A scintillation detector is selected based on the type of scintillator needed. The two most important parameters are ionizing radiation detected by the scintillator and selecting the scintillator decay time. The decay time is critical to select a scintillation detector, as it defines the minimum sampling time for the system. The decay time is the delay between the incoming radiation absorption by the scintillator and its reemission as a photon of visible light. It can vary from a few nanoseconds to several hours, depending on the scintillator material (Leo, 1994). Therefore, based on the radionuclide selected, this research required a scintillation detector capable of detecting gamma rays ($\gamma$-ray). In
addition, a scintillator with minimum decay time was required to allow for accurate agglomerate tracking.

Sodium iodide doped with thallium NaI(Tl), is the most widely used scintillator material (Leo, 1994). Its decay time of only 0.23 µs makes it highly desirable for the objective of this research. The scintillation detectors used were the 2M2/2 model manufactured by Saint-Gobain Crystals. This NaI(Tl) scintillator is cylindrical, measures 2” diameter by 2” thickness, and has a 14-pin connector. An aluminum sleeve shields the scintillator to avoid disruption by external photons. The crystal contains 0.2% thallium (Tl) impurity as the activator. The scintillator is connected to built-in PMT (Marks, 2016). The scintillation detector is shown in Figure 3-16.

The ORTEC digiBase was coupled with the 2M2/2 scintillation detector to allow for data acquisition, transfer, and processing through CONNECTION-32 and MAESTRO®-32 programs (ORTEC, 2020). This base consisted of a preamplifier and a robust digital Multichannel Buffer (MCB). An ORTEC digiBase connected to a 2M2/2 scintillation detector is shown in Figure 3-17.

The scintillation detectors were located around the bed in three layers of four detectors separated by one-inch-thick cold rolled steel collimation plates (Figure 3-18). The scintillation detectors and the collimation plates were installed within a structure, allowing for their positioning at any chosen axial location. The collimation plates can also be moved radially relative to the unit. Their purpose was to reduce long-distance tracer detection, especially near the bottom and top of the bed, without disturbing measurement at short
tracer-detectors distances. By reducing the counts detection of detectors further away from the source, the weight of these detectors in CARPT coordinates calculations was reduced (Lin, 1981), allowing for more precise results.

Figure 3-18. RPT measurement system
3.3.3 Data transfer & tracer coordinates determination

Each digiBase was connected to a computer via a USB cable. These were the “client” computers of the data acquisition (DAQ) network (Lenovo ThinkPad). A “server” computer (Dell Inspiron) time stamps each DAQ event and synchronizes the “client” computers, allowing them to take a reading over the same time interval. This radioactive counts measurement over a given time interval is referred to as an “event” in this thesis. The network measured the detectors signal every 8 to 20 ms, significantly improving single computer DAQ systems. The time step value varies with the source radiation strength: a more potent source will lead to more detections in scintillation detectors, requiring a greater processing and transmission time between two measurement points. On both the “clients” and “server” computers, the data are processed using the LabWindows CVI (National Instruments, Austin, TX) platform and the C++ software developed by Sanchez Careaga (2013). The same research also demonstrated that adding a powered USB hub between the digiBase and the “client” computer sped up data transmission, reducing the required sampling time significantly (from 60 ms to 20-30 ms).

3.3.3.1 Computer-Aided Radioactive Particle Tracking (CARPT)

Several rendition techniques exist to determine the tracer coordinates inside the reactor at a given time using the combined radiation signals obtained from the scintillation detectors set. Computer Automated Radioactive Particle Tracking (CARPT) and Monte Carlo simulation are the two most common methods.

Due to the amount of data recorded per run in this research (1,000,000 – 10,000,000 events), the significant computing time required by Monte Carlo simulation is unsuitable (up to one second per event, using a Lenovo T410, with an Intel® Core™ i5-520M CPU @ 2.40GHz). On the other hand, the CARPT method was fast (down to 0.03 ms/event, using a Lenovo T410, with an Intel® Core™ i5-520M CPU @ 2.40GHz) at the cost of less accuracy. The following section will describe the primary calibration and calculations associated with CARPT. Chapter 4 will present the investigation and optimization of the CARPT method underwent during this research.
3.3.3.2 CARPT calibration

A calibration curve relating $\gamma$-rays counts to distance is established for each detector. The calibration data obtained are processed to produce a curve fit of the $\gamma$-rays raw data in an empty vessel. Various fitting methods are used to describe the different domains of distance versus $\gamma$-rays counts relationships. The overall bed can be described with a general polynomial/power 1D fitting. On the other hand, short range-distance is better described with fitting equations accounting for the tracer-detector solid angle. Finally, the most accurate results will require tedious in situ calibration, at the operating condition of interest, to properly account for the varying bed density (Chaouki et al., 1997).

The experimental unit size coupled with the large number of scintillation detectors used in this research limited the amount of calibration possible within a reasonable amount of time. Therefore, the calibration method was limited to the general polynomial/power 1D fitting for the overall bed. An example of a calibration curve is shown in Figure 3-19. The rest of

**Figure 3-19.** Typical CARPT calibration curves for one of the twelve detectors used in this research

- **Calibration 1)** $L_i = 0.0739 \, C_{Ni}^{0.5760}$
  $R^2 = 0.9797$
- **Calibration 2)** $L_i = 0.0833 \, C_{Ni}^{0.5602}$
  $R^2 = 0.9514$
- **Calibration 3)** $L_i = 0.0836 \, C_{Ni}^{0.5373}$
  $R^2 = 0.9516$
the CARPT calibration is presented in Appendix C: Standard RPT and CARPT calibration performed.

A significant limitation of this approach is that the calibration correlation ignores the angle between the tracer and a horizontal plane through the virtual detector center. This simplification contributes to the error since it aggregates variable values of wall absorption in the same 1D curve. Another significant limitation is that the calibration is conducted in an empty bed: the bed absorption occurring during measurements is not incorporated into the calibration curve. Besides, due to variable voidage, this bed absorption varies both with bed elevation and time.

3.3.3.3 Main CARPT equations

As mentioned in the previous section, this research used a general polynomial/power 1D calibration fitting for the overall bed. Several models were tested, and the best fit was obtained through a power law. To simplify calculations and remove the effect of the attenuation of the radionuclide other time, the counts at detector i is normalized by the sum of all counts detected by all detectors during the same time interval (Khanna et al., 2008; Sanchez Careaga, 2013), as in the following equation:

\[ L_i = A_i (CN_i)^{B_i} = A_i \left( \frac{C_i}{\sum_{i=1}^{N_D} C_i} \right)^{B_i} \]  \hspace{1cm} (3.3)

where:

- \( CN_i \): normalized counts detected at detector i
- \( C_i \): counts detected at detector i
- \( N_D \): total number of detectors
- \( L_i \): Euclidian distance between the tracer and the detector i
- \( A_i \) & \( B_i \): calibration coefficients of detector i

This normalization procedure and its implication on CARPT accuracies are discussed in detail in Chapter 4.
Once the tracer-detector distance $L_i$ is obtained for each detector, by defining an arbitrary reference frame, the CARPT formula can be written as shown in Equation (3.4):

$$L_i^2 = (x - x_i)^2 + (y - y_i)^2 + (z - z_i)^2$$

(3.4)

where:

- $L_i$: distance between the tracer and the detector $i$
- $[x_i, y_i, z_i]$: the coordinates of the position of detector $i$
- $[x, y, z]$: coordinates of the position of the tracer

As described by Lin et al. (1985), by applying a weighted linearized regression scheme combining the distances calculated and rearranging and rewriting the system of equations in matrix form, we get Equation (3.5):

$$\bar{A}_{N \times 4} \bar{X}_{4 \times 1} = \bar{B}_{N \times 1}$$

(3.5)

where:

$$\bar{A} = \begin{bmatrix} 1 & -2x_1 & -2y_1 & -2z_1 \\ 1 & \cdots & \cdots & \cdots \\ 1 & \cdots & \cdots & \cdots \\ 1 & -2x_N & -2y_N & -2z_N \end{bmatrix}, \quad \bar{X} = \begin{bmatrix} x \\ y \\ z \end{bmatrix}, \quad \bar{B} = \begin{bmatrix} L_1^2 - x_1^2 - y_1^2 - z_1^2 \\ \cdots \\ L_N^2 - x_N^2 - y_N^2 - z_N^2 \end{bmatrix}$$

- $L_i$: distance between the tracer and the detector $i$
- $[x_i, y_i, z_i]$: the coordinates of the position of detector $i$
- $[x, y, z]$: coordinates of the position of the tracer
- $\kappa$: parameter to minimize for convergence

This matrix corresponding to the system of equations cannot be inverted directly because the components in the left-hand side (LHS) vector are not all known with the same accuracy (Chaouki et al., 1997). Hence, more weight is given to the small rather than to the large tracer-detector distances. Lin et al. (1985) proposed the following weighting matrix $\bar{W}$, using the weighting exponent $\lambda$. The choice of this exponent value was discussed in previous research (Lin et al., 1985; Sun, 1985)
Finally, Lin et al. (1985) solved the system of equations using a standard matrix equation, as shown in Equation (3.6), with an optimal weighting exponent $\lambda = 3$.

$$\bar{X}_{4 \times 1} = \left( \bar{A}_{4 \times N}^T \bar{W}_{N \times N} \bar{W}_{N \times N}^T \bar{A}_{N \times 4} \right)^{-1} \bar{A}_{4 \times N}^T \bar{W}_{N \times N} \bar{W}_{N \times N} \bar{B}_{N \times 1}$$  (3.6)

As mentioned previously, Chapter 4 will present the investigation and optimization of the CARPT method underwent during this research. It includes signal denoising using pre- and post-processing recorded signals and absorption correction based on the fundamental radiation absorption equations.

### 3.3.4 RPT validation: dye sampling

Experiments were conducted with dyed coke to verify the radioactive tracer Residence Time Distribution (RTD). The objectives were to: 1) verify that light tracers aiming at simulating micro-agglomerates and emulsion phase (density around 900 kg·m$^{-3}$) have a similar residence time distribution as coke particles, 2) verify how heavy tracers aiming at simulating wet-agglomerates (density around 1200 kg·m$^{-3}$) travel through the bed. These experiments are discussed in detail in Chapter 4.

In solids RTD tests, the tracer must have physical properties similar to those of the particles that the RTD profile wants to represent. Indeed, particles of different sizes or densities may behave differently (Bader et al., 1988). Therefore, it is best to use identical particles and differentiate those used as a tracer to ensure realistic measurements. This is not achievable with RPT, but the use of dyed coke fulfills this objective. Therefore, dyed coke can be used to verify the representativity of RPT results.
Coloured food dyes were used. Several dyes of different colours were used consecutively to perform several tracer experiments with the same batch of bed particles. First, fluid coke
tracer particles were dyed then dried in a shaker reactor (Sanchez Careaga et al., 2020). Next, dyed coke particles were injected into the bed and recirculated. Finally, coke particles were sampled from the bed then mixed with water to recover the dyes. After filtration, the resulting aqueous solution was analyzed by a spectrophotometer to measure the concentration of each dye. The experimental setup for dye RTD measurements included: one pressurized injection vessel located in the recirculation line before the gas-solids mix zone and one screw sampler located above the pinch valve. The experimental setup is presented in Figure 3-20. More details about the dyed coke procedure are presented in Appendix D: Solids RTD validation experiments with dyed coke.

A critical difference with RPT measurements is that the sampled coke particles might have undergone more than one recirculation. Indeed, the sampled coke particles can be:

- Not dyed (most of the particles)
- Dyed
  - Underwent one-pass through the bed
  - Underwent more than one pass through the bed

Therefore, and contrary to the single tracer RPT method, the dye concentrations measured required the use of deconvolution methods to obtain representative results for solids residence time (Brereton, 1987; Levenspiel, 1999). On the other hand, RTD data obtained with RPT can be convoluted to compare with dyed coke results. A detailed explanation of the convolution/deconvolution procedure is provided in Appendix D: Solids RTD validation experiments with dyed coke.

### 3.4 Radiation transmission

Using the same equipment as for RPT, radiation transmission experiments can also be performed. For these measurements, the radioactive source ($^{46}$Sc in this research) is located outside the bed, on one side of the unit, and one (or more) radioactive sensor (NaI(Tl) scintillation detector in this study) is located on the other side of the unit. An example of a radiation transmission measurement setup is presented in Figure 3-21.
The radiation absorption measurement principle is based on the fundamental physics behind the decrease of radiation when the source distance increases. Two mechanisms explain this decrease: 3D dispersion of the radiation in the volume surrounding the source and medium absorption through medium traversed (Bartholomew and Casagrande, 1957).

The first mechanism explaining the decrease of radiation with distance is the 3D dispersion of the emitted radiation: the amount of radiation emitted at the source is spread over a sphere of increasing radius, leading to a quadratic decrease of radiation strength with distance. It can be expressed as the ratio of irradiance $I$ (at distance $r$ from source) to the
irradiance $I_0$ (at the closest possible distance from the detector). This ratio, also known as Inverse Square Law, is presented in Equation (3.7).

\[
\frac{I}{I_0} = \frac{L_{i,0}^2 \cdot \Phi_S}{L_i^2 \cdot \Phi_0}
\]  

(3.7)

with:

- $I$: radiation received at distance $L_i$ from the radioactive source ($J \cdot s^{-1} \cdot m^{-2}$)
- $I_0$: radiation received at a distance $L_{i,0}$ from the radioactive source ($J \cdot s^{-1} \cdot m^{-2}$)
- $L_i$: distance between source and detector $i$ (m)
- $L_{i,0}$: minimum distance between source and detector $i$ (m). $L_{i,0}$ larger than zero since both the source occupy a finite volume.
- $\Phi_S$: sub-incident flux, at $L_i$
- $\Phi_0$: incident flux, at $L_{i,0}$

The ratio of sub-incident flux at $L_i$ over the incident flux at $L_{i,0}$, $\Phi_S/\Phi_0$, comes from the second mechanism explaining the decrease of radiation with distance: radiation absorption through medium traversed. Within a given medium $j$, this ratio can be expressed by a generalized Beer-Lambert law, such as the one presented in Equation (3.8).

\[
\frac{\Phi_{S,j}}{\Phi_{0,j}} = \exp\left(-\alpha_j \cdot \rho_j \cdot L_j\right)
\]  

(3.8)

with:

- $\Phi_{S,j}$: sub-incident flux, after crossing the medium $j$
- $\Phi_0$: incident flux, before crossing the medium $j$
- $L_j$: absorbing medium $j$ thickness (m). It is the effective distance crossed through the medium $j$.
- $\rho_j$: absorbing medium $j$ density (kg·m$^{-3}$)
- $\alpha_j$: mass absorption coefficient of the medium $j$, for a given radiation energy spike ($m^2 \cdot kg^{-1}$)
It should be noted that each medium traversed requires its own Beer-Lambert absorption law: the overall absorption can be represented as the product of the absorptions through each medium i traversed. For instance, if radiation goes through air, coke and steel, the overall absorption is the product of the absorption in air, by the absorption in coke, by the absorption in steel.

The radiation absorption measurement takes advantage of radiation absorption through a medium (Equation (3.8)). When source and detector are kept immobile, the only non-random variations of radiation signals result from fluctuations in local bed density. A significant limitation is that the measuring volume is limited to the small cone between the source and the detector. Given its size, this cone can be approximated as a 1D snapshot of the bed. In a 3D system, this will only represent a partial measurement of the cross-sectional bed properties at a given height. On the other hand, pressure-based measurements mentioned in the previous section offer a complete 3D snapshot of the bed between the two measurement taps. The difference between these two methods is used in this research to

**Figure 3-22.** Bubbling/Turbulent detection through 1D pseudo voidage from radiation transmission ($H_{\text{bed}} = 115$ cm, $D_{\text{bed}} = 25$ cm – 2 replicates)
extract valuable information about radial voidage profiles at a given height. The method has been discussed in Chapter 2.

The radiation transmission signal was recorded and analyzed to:

- detect the transition from bubbling regime to turbulent regime. Figure 3-22 presents the bubbling/turbulent detection using 1D pseudo voidage from radiation transmission measurements.
- measure the 1D pseudo-voidage at various bed heights (Wormsbecker, Pugsley, van Ommen, et al., 2009; Li et al., 2019). By comparing this 1D pseudo-voidage with the voidage calculated at the same height through pressure measurements, the radial variation of voidage at this bed height can be obtained.

3.5 E-probes

The triboelectric effect (or “triboelectric charging”) is a type of electric charge transfer between certain materials after they come in physical contact. E-probes are sturdy intrusive probes using, at least partially, this triboelectric effect to get information about the bed hydrodynamics. Indeed, in the dense phase bed, the triboelectric current is generated by the collisions of particles induced by the bubble motion around the probe (Li et al., 2019).

This research used E-probes (Jahanmiri, 2017; Li et al., 2019) for targeted local validation measurements of internal bed structures, especially bubble profiles. It should be noted that the solids used in this research, fluid coke, made the use of E-probes more challenging than with silica sand. In addition, significant extra shielding and grounding were required to limit electrical noise. A schematic of the E-probe used is presented in Figure 3-23.

![Figure 3-23. Modified E-probe](image)

**Figure 3-23. Modified E-probe**
3.6 Vapour & liquid carry-under measurements

This research selected Varsol™, a tracer injected as a liquid and carried-under as either liquid or vaporized liquid. Two measurement tools are therefore used. First, a vapour concentration sensor is located in the recirculation line. Second, due to the pure air addition for pneumatic transport in the recirculation line, any Varsol™ carried-under as a liquid would instantaneously vaporize once in the recirculation line. Therefore, this first sensor would detect vapour and liquid carry-under indistinctly. Consequently, a second measurement tool is required to detect the injected flowrate for which liquid starts to be carried-under. Figure 3-24 presents this experimental system.

3.6.1 Tracer: Varsol™

Varsol™ was selected as a tracer for the vapour & liquid carry-under study as it could be injected as a liquid in the bed, then evaporates at room temperature (20 °C) at a relatively fast rate. Based on the manufacturer datasheet, its relative evaporation rate equals 14 % of the n-Butyl Acetate evaporation rate (Kaklamanos, 2017). This reference substance evaporate at a rate of 2.59 \(10^{-2}\) g·m\(^{-1}\)·s\(^{-1}\), with \(T = 20^\circ\text{C}\), \(U_G = 0.25\) m·s\(^{-1}\), \(L = 0.5\) m (Abbot, 2021).

Varsol™ also wets coke perfectly and therefore does not promote agglomeration (McDougall et al., 2005). The liquid density, vapour density, and vapour pressure of Varsol™ at room temperature are 789 \(\text{kg} \cdot \text{m}^{-3}\), 5.9 \(\text{kg} \cdot \text{m}^{-3}\), and 293 Pa, respectively (Kaklamanos, 2017). Finally, the Lower Explosive Limit (LEL) of Varsol™ is at 0.6 %. With a saturated vapour pressure of 2 mm Hg (20 °C), one can operate at room temperature well below the LEL, allowing for safe operation in the experimental unit.
Figure 3-24. Liquid & Vapor carry-under experimental set-up
3.6.2 Vapour sensor

MQ-135 is a gas sensor (Figure 3-25) based on a chemiresistor and designed to measure the concentration of gases such as CO, CO₂, NH₄, Acetone, Toluene, and Ethanol (Abbas et al., 2020). It was selected for the liquid & vapour carry-under measurements due to its small size (3 cm x 2 cm x 3 cm), high sensitivity to low concentration (down to 10 ppm), and relatively fast response time (steady-state reached in 6-7 minutes for up-step concentration measurements in a well-mixed volume). The MQ-135 is connected to and controlled by a microcontroller board (Arduino Uno/Mega).

![Figure 3-25. MQ-135 sensor (Robotshop, 2021)](image)

The principle of the MQ-135 sensor can be described as follows. One metal oxide chemiresistor of variable resistance $R_S$ comprises a micro Al₂O₃ ceramic tube, a Tin-Dioxide (SnO₂) sensitive layer, a measuring electrode, and a heater (Hanwei Electronics, 2012). The heater is necessary because metal oxide sensors require high temperatures (200 °C or higher) to operate: the activation energy must be overcome for the resistivity to change (Khanna et al., 2008). When detected gases interact with the chemiresistor, its variable resistance $R_S$ decreases. One amplification resistor of resistance $R_L$ is used to adjust the measured voltage. The electrical circuit is presented in Figure 3-26.

The MQ-135 sensor is powered with the input voltage $V_C$ (5V), and the microcontroller board measures the output voltage $V_{RL}$. The variable resistance of chemiresistor $R_S$ can be calculated, for a given amplification resistance $R_L$, using Equation (3.9):

$$R_S = \left( \frac{V_C \cdot R_L}{V_{RL}} \right) - R_L \quad (3.9)$$

Significant limitations of MQ-135 sensors are linked to their sensitivity to humidity and temperature (Kalra et al., 2016). The latter is especially important: it is related to the high operating temperature required, mentioned earlier, and means that the sensor cannot be
used reliably within a flowing gas. Indeed, the gas convection leads to unstable operating temperature and artificially increases the chemiresistor $R_S$ value.

Figure 3-26. MQ-135 circuit (modified from Hanwei Electronics (2012))

According to the manufacturer datasheet (Hanwei Electronics, 2012), the resistance $R_{S,0}$ is defined as the value of sensor resistance $R_S$ in clean air with 100 ppm of NH$_3$ ($T = 20 \degree$C, humidity = 65%). The ratio of variable resistor $R_S$ by $R_{S,0}$ is therefore constant and equal to 3.6 for clean air at these conditions. $R_S/R_{S,0}$ decreases with increasing gas tracer concentrations. Figure 3-27 shows the calibration curves for different gases. Figure 3-27-a) is extracted from the manufacturer datasheet. Figure 3-27-b) was obtained experimentally by injecting a small volume of liquid Varsof™ in a 1.18 L beaker. A small fan provided gas circulation within the beaker, and the signal was recorded until the liquid completely vaporizes (15 minutes or more).
Figure 3-27. MQ-135 calibration curves (T = 20 °C, Humidity = 65%)

a) Calibration from the manufacturer datasheet (Hanwei Electronics, 2012)

b) Experimental calibration with Varsol™

It can be noted that the Varsol™ calibration curve is no longer logarithmic for high concentrations. This variation from the expected trend could be due to the limitation of the chemiresistor itself (the range of concentration is tenfold higher than the one presented in the manufacturer datasheet) or to mass transfer limitation in the calibration volume. However, this limitation is not critical since the sensor is primarily used to detect the transition between no vapour detected and some vapour detected. Therefore, only small concentrations need to be measured accurately.

Finally, in logarithmic calibration zones, the target gas concentration (in ppm\text{mol}) can be expressed as in Equation (3.10):

\[
C_{ppm,\text{mol}} = A \left( \frac{R_S}{R_{S,0}} \right)^{-B}
\]  

(3.10)

A and B are calibration constants (for a given gas, temperature, and humidity).
3.6.3 Gas-liquid injector

Liquid Varsol™ is mixed with atomization air before injection into the bed and injected as an atomized liquid spray within the bed, similar to the steam atomized bitumen spraying in commercial Fluid Cokers. A syringe pump transports the liquid Varsol™ at a fixed flow rate (ranging from 0.5 mL·min⁻¹ to 7 mL·min⁻¹) from a storage tank to a gas-liquid mixer just before the bed injection. The combined mixer-injector design is presented in Figure 3-28. The liquid injection takes place in a single nozzle at a time (out of a total of 40 available, grouped in five banks of eight nozzles). The impact of the injection bank, the nozzle selected within the bank, the nozzle penetration, and the overall bed hydrodynamics (baffle addition, injection profile, …) are discussed in Chapter 2.

3.6.4 Capacitance probe

One significant advantage of using Varsol™ as a tracer is that its relative permittivity can be easily differentiated from the relative permittivity of dry coke particles (Hamidi, 2015; Mohagheghi Dar Ranji, 2015). Therefore, by locating a capacitance probe above the pinch valve leading to the recirculation line, any liquid Varsol™ carried-under can be detected. Coupled with the results from the chemiresistor, it allows for a qualitative differentiation between vapour & liquid carry-under. If the chemiresistor detects Varsol™ while the capacitance probe does not detect liquid, only Varsol™ in vapour form is carried-under. If the capacitance detects liquid, then at least some of the carry-under happens in the liquid
phase. More sophisticated analysis can also be used to build a calibration curve between Varsol™ concentration and capacitance measurement (Mohagheghi Dar Ranji, 2015), allowing for qualitative analysis of respective vapour & liquid carry-under.

An AC-based capacitance meter was used to measure the bed capacitance between electrodes. It used a sine-wave voltage as the excitation source to produce an AC input current and an amplifier to convert it into an AC voltage (Yang, 1996). The measurement electrodes are the section of the steel bed wall located near the pinch valve and a steel rod that is electrically insulated from the bed wall to block the conduction current and ensure that only the capacitance current passes through the electrodes. A microcontroller board recorded the capacitance signal (Arduino Uno/Mega). The capacitance measurement system is presented in Figure 3-29, and a simplified electrical schematic is presented in Figure 3-30.
Figure 3-30. Simplified electric scheme of capacitance measurement system

3.6.5 Summary of the vapour & liquid carry-under measurements

This section briefly summarizes the measurements methods used to investigate the vapour & liquid carry under in the experimental unit.

Table 3-3 presents each method with its principle, its measurement location, its advantages and limitations and gives relevant literature references.
Table 3-3. Summary of the vapour & liquid carry under measurements with Varso™

<table>
<thead>
<tr>
<th><strong>Measurement method</strong></th>
<th><strong>Chemiresistor</strong></th>
<th><strong>Capacitance probe</strong></th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Measured parameter</strong></td>
<td>Air composition (target gas concentration)</td>
<td>Medium capacitance (Liquid concentration)</td>
</tr>
<tr>
<td><strong>Measurement location</strong></td>
<td>In the dilute phase (recirculation line)</td>
<td>In the dense phase (drain pipe at the unit bottom)</td>
</tr>
</tbody>
</table>
| **Advantages** | - Sensitive to very low concentration | - Can measure the liquid & gas phase  
- Allows in-bed measurements |
| **Limitations** | - Can only measure in the gas phase  
- Requires zone of low flowing gas  
- Saturates for high concentrations | - Very low sensitivity to the gas phase  
- Electrostatic noise from the fluidized bed |
3.7 References


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Chapter 4

4 Improvement & validation of Radioactive Particle Tracking (RPT) measurements

Radioactive Particle Tracking (RPT) is a crucial measurement tool in this research. It is therefore critical to ensure accurate RPT results. This chapter investigates several methods to reduce both random and systematic RPT measurement errors. To achieve this, consistency criteria and independent experimental validation methods are selected to validate the RPT results. Several methods developed to enhance RPT data are reviewed and modified if necessary; these methods include the normalization of counts, radiation absorption correction, and time-series smoothing methods. Test criteria and validation experiments are then used to select the best RPT enhancement methods.

The experimental unit used in this research presents specific challenges related to the use of RPT.

1) It presents a relatively large measurement volume (diameter up to 0.25 m, height up to 1.4 m), resulting in larger distances between source and detectors when compared to typically published RPT studies (Khanna, 2008; Sanchez Careaga, 2013; Efhamia and Al-Dahhan, 2016; Kalo et al., 2019; Dam et al., 2021).

2) It uses gas velocity $u_G$ ranging from 0.3 m·s$^{-1}$ to 0.9 m·s$^{-1}$. Therefore, the tracer can be picked up and released by bubbles moving upward at a velocity of 2 m·s$^{-1}$. This results in tracer velocities and accelerations that can be high, requiring accurate monitoring of the tracer position every 10 ms or so. This short sampling time coupled with the random nature of $\gamma$-ray emissions from radioactive sources leads to increased error, as discussed in Chapter 3 - Section 3.3. To correct this, sophisticated data processing methods are required.
4.1 Validation of RPT results

Several consistency criteria are defined based on the physical limitations on tracer motion within a fluidized bed to validate RPT data. Critical experiments are also used to validate RPT with independent measurement methods.

4.1.1 Consistency of RPT results

A set of criteria was chosen to test and select the RPT enhancement methods. These criteria were selected based on known physical limitations that realistic solutions must respect. They include:

- A reduction of impossible tracer positions: the tracer cannot be outside the bed.
- A reduction of the amount of zones cells with an artificially low tracer presence: unless internals such as ring baffles are added, no zone of very low tracer presence should appear within the dense bed volume, except for the baffle(s) pockets (if baffle(s) is/are used).
- A reduction of impossible velocities: the tracer velocity cannot vary outside a range of reasonable values (\([-2.5; +2.5]\) m·s\(^{-1}\)) and should not fluctuate at a very high frequency (> 50 Hz).

4.1.1.1 Reduction of impossible positions

The radioactive tracer cannot be outside the vessel. The position of these walls is known precisely within the measurement volume. At any axial position, the radial boundaries are at the position of the tapered vessel wall. The axial boundaries are located respectively at the sparger level (\(z = 0\) m) and 20 cm above the bed surface (located, to be on the safe side, at \(z = 1.5\) m).

Therefore, a position measured through CARPT can either be within the wall boundaries (inside the bed) or outside the wall boundaries (outside the bed). For a dataset of \(N_E\) events with \(N_{Out}\) positions located outside the boundaries, the simple numerical criterion assessing the proportion of impossible positions is defined by the following equation:
\[ \%Pos_{imp} = \frac{N_{\text{out}}}{N_E} \]  \hspace{1cm} (4.1)

**Figure 4-1.** Criterion to reduce out of bed detected positions

(Short taper, \( u_{G,\text{Top}} = 0.9 \, \text{m/s} \), 5 injection banks, \( L_N = 0 \, \text{cm} \), 1 baffle below 2\(^{nd}\) injection, \\
\( F_S = 0.47 \, \text{kg/s}, m_{S,\text{Bed}} = 60.3 \, \text{kg}, D_{\text{Agg}} = 1.3 \, \text{cm}, \rho_{\text{Agg}} = 1100 \, \text{kg/m}^3, \Delta t_{\text{EXP}} = 43 \, \text{h} \))
The objective is to minimize \( \% \text{Pos}_{\text{Imp}} \). This minimization needs to be balanced by other criteria to avoid over-correction of the dataset (all positions artificially moved within vessel boundary). A visual example is presented in Figure 4-1.

\[ (\text{Detection per volume})_{\text{Normalized},i} = \frac{N_{\text{detection},i}}{\sum N_{\text{detection}}} \frac{1}{V_{\text{cell},i}} \]  

(4.2)

with:

- \( N_{\text{detection}} \): number of tracer detection in cell \( i \) during the experimental run
- \( \sum N_{\text{detection}} \): total number of tracer detection during the experimental run

**Figure 4-2.** \( z-r \) projection used for colormaps
• $V_{\text{cell},i}$: Volume of cell $i$

The colormap presented in Figure 4-1 is a 2D projection of the 3D tracer positions repartition detected for a whole recorded run. This assumes an angular symmetry of the system. This projection will be re-used extensively in this document and is presented in Figure 4-2.

4.1.1.2 Reduce the presence of bed zones with an artificially low tracer presence

The second criterion is used to avoid over-correction when reducing the number of impossible positions. This criterion is based on the principle that, in the dense bed, zones where the tracer would have locally a much lower probability of presence are very unlikely to appear. Using the same boundaries as the last criterion, the volume within the boundaries (inside the bed) $V_{\text{InBed}}$ is investigated. Each position of a 3D grid within the bed is included in a cell of 2.5 cm inter-node distance (cell volume of $1.056 \times 10^{-2}$ L). The tracer presence in these cells is investigated. This cell is considered to have an artificially low tracer presence if there are less than $[0.1 \cdot (\Sigma N_{\text{detection}})/N_{\text{Cell}}]$ tracer detections in a given cell. This threshold value corresponds to ten times fewer tracer detections than expected if the tracer positions are uniformly distributed in the bed volume. For 73 523 positions in the bed volume, and with 1 000 000 tracer positions recorded, it corresponds to less than one detection per cell.

The total volume of cells with an artificially low tracer presence in the bed, $V_{\text{LowTracerBed}}$, is obtained, and the numerical criterion assessing the proportion of artificially low tracer presence inside the bed is defined by:

$$\%\text{LowTracer}_{\text{In}} = \frac{V_{\text{LowTracerBed}}}{V_{\text{InBed}}}$$ \hspace{1cm} (4.3)

The addition of baffle(s) within the unit could create a gas-only baffle pocket with a very low probability of agglomerate presence. Therefore, the volume in-bed, $V_{\text{InBed}}$, is modified when one or more baffles are present to exclude baffle pockets.
The objective is to limit the increase in the relative value of $\% V_{\text{LowTracerIn}}$ when corrections are applied (the reference case being the dataset without corrections).

**Figure 4-3.** Criterion to reduce the presence of bed zones with artificially low tracer presence

a) Absence of zones with artificially low tracer presence, b) Presence of zones with artificially low tracer presence

(Short taper, $u_{G,\text{Top}} = 0.9 \text{ m}\cdot\text{s}^{-1}$, 5 injection banks, $L_N = 0 \text{ cm}$, 1 baffle below 2nd injection, $F_S = 0.47 \text{ kg} \cdot \text{s}^{-1}$, $m_{S,\text{Bed}} = 60.3 \text{ kg}$, $D_{\text{Agg}} = 1.3 \text{ cm}$, $\rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $\Delta t_{\text{EXP}} = 43 \text{ h}$)
4.1.1.3 Reduction of impossible velocities

The third criterion is based on the regularity of the tracer motion. When moving throughout the fluidized bed, the tracer cartesian coordinates should vary without excessive fluctuations, which would result in impossible velocities. Indeed, excessive fluctuations of coordinates can only be caused by the CARPT inaccuracies (due to the random fluctuations in the counts rate detected and systematic errors). The intensity of these fluctuations can be quantified using a moving standard deviation of Cartesian coordinates (with a window size of $\Delta t_{\text{MOV}}$). Therefore, the criterion is defined as the average of these moving standard deviations such as:

$$AVG_{\text{MovStd}} = \left[ \frac{\text{Average}(\text{Std}_{\text{Mov}}(x(t), \Delta t_{\text{MOV}}))}{\text{Average}(\text{Std}_{\text{Mov}}(y(t), \Delta t_{\text{MOV}}))} \right]$$

(4.4)

The objective is to minimize $AVG_{\text{MovStd}}$, but only to a certain extent: indeed, this criterion will promote static tracer behaviour if used alone. Therefore, it should be used only as a complementary criterion.

Due to a more extensive motion on the z-axis (between 0 and 1.5 m) compared to the x- & y-axis (between -0.15 and +0.15 m), the average moving standard deviation is more prominent for the z-component.

For the window size $\Delta t_{\text{MOV}}$, several values were tested, ranging from 15 ms (covering only two successive measured positions) to 5 seconds (time required to move 1.5 m, more than the bed height, at a velocity of 0.3 m·s$^{-1}$, the superficial gas velocity at the bottom of the bed). Some examples are presented in Figure 4-4. The objective was to obtain a window size $\Delta t_{\text{MOV}}$ allowing for a moving standard deviation MovStD which detect primarily high frequency fluctuations (due to signal error and which the signal processing is looking to get rid of) while giving a low weight to low-frequency variations (due to the regular tracer motion, which the signal processing should not modify significantly). Therefore, the resulting average $AVG_{\text{MovStd}}$ would become an efficient parameter to detect and eliminate high-frequency fluctuations.
Figure 4-4. Criterion to reduce unsteady trajectories (on x-coordinates)

a) x-coordinates, b) Moving standard deviation

(Short taper, \( u_{G,Top} = 0.9 \, \text{m} \cdot \text{s}^{-1} \), 5 injection banks, \( L_N = 0 \, \text{cm} \), 1 baffle below 2\(^{nd} \) injection, \( F_S = 0.47 \, \text{kg} \cdot \text{s}^{-1} \), \( m_{S,Bed} = 60.3 \, \text{kg} \), \( D_{Agg} = 1.3 \, \text{cm} \), \( \rho_{Agg} = 1100 \, \text{kg} \cdot \text{m}^{-3} \), \( \Delta t_{EXP} = 43 \, \text{h} \))

To select the value of \( \Delta t_{MOV} \), the value of the average moving standard deviation \( \text{AVG}_{\text{MovStd}} \) of two signals (from recorded RPT experiments) with different noise levels were compared. The objective was to test various time windows \( \Delta t_{MOV} \) to maximize the ratio of the average of the moving average of the high noise signal \( \text{AVG}_{\text{MovStd,1}} \) (signal 1) by the average of the moving average of the low noise signal \( \text{AVG}_{\text{MovStd,2}} \) (signal 2). As shown in Figure 4-5,
very short time windows $\Delta t_{\text{MOV}}$ (under 0.5 seconds) should be avoided as the time interval covered does not allow for a clear distinction of $\text{AVG}^\text{MovStd}$. Long time windows $\Delta t_{\text{MOV}}$ (above 3 seconds) become less efficient since the effect of tracer motion becomes bundled with the high-frequency fluctuations, making $\text{AVG}^\text{MovStd}$ less pertinent.

**Figure 4-5.** Example of the ratio of the average of moving standard average between a signal with high noise (signal 1) & a signal with low noise (signal 2), for several window time $\Delta t_{\text{MOV}}$

Therefore, the time window $\Delta t_{\text{MOV}}$ used for the moving standard deviation $\text{AVG}^\text{MovStd}$ was set to 2 seconds.

### 4.1.2 Comparison of RPT results with other measurement methods

This section presents several validation experiments conducted to verify the RPT results.

#### 4.1.2.1 Tracer time distribution

As introduced in Chapter 3 – Section 3.3.4, experiments were conducted with dyed coke to verify the radioactive tracer Residence Time Distribution (RTD). This verification is critical for the development of the model presented in Chapter 2.
Fluid coke was dyed using a mixture of water and coloured food dye. Dyed coke has the same properties as the original Fluid Coke. Dyed coke was then injected into the recirculation line as a pulse. Coke particles were sampled from the recirculation line in successive intervals of 5 s, using a screw sampler. Each time interval results in a batch of coke samples. Figure 3-20 presents the experimental setup. Based on the solids residence time in the vessel, averaging around two minutes, samples were acquired for 5 minutes (50 samples). The dyes from each coke batch were dissolved in water. After filtration, the resulting aqueous solution was analyzed by a spectrophotometer to measure the concentration of dye in each.

Figure 4-6 presents a sample of coke batch absorbance measured during a dyeing experiment. Three phases are visible: initially \((t \leq 22.5 \text{ s})\), the absorbance measured remains low and varies only marginally, then \((22.5 \text{ s} \leq t \leq 152.5 \text{ s})\) the absorbance increases with a reverse exponential speed, finally \((t \geq 152.5 \text{ s})\), the absorbance values remain constant and even starts to decrease slowly. It should be noted that samples obtained after a much longer time \((t \geq 2500 \text{ s})\) present a significantly lower absorbance value. This decrease can be explained by the dilution of dyed-coke particles injected \((1.5 \text{ kg})\) with the rest of the coke particles \((60 \text{ kg})\) in the bed, to a point where the dye concentration measured in the recirculation line at any time becomes very small. The total disappearance of dye detection suggests that, in the long run, the dye coating could be destroyed, either in the fluidized bed or due to the abrasion in the recirculated line. Once detached from coke particles, the dye becomes equivalent to fines and should be cleared quickly through the unit cyclone.

Using the spectrophotometer scans presented in Appendix D: Solids RTD validation experiments with dyed coke, an integration wavelength range is selected depending on the colour of the food dye used. Therefore, an average integrated absorbance can be obtained as a function of time (each batch of coke sampled during 5 s representing the average dyed-coke flow recirculated during this time interval).
Figure 4-6. Dyed-coke absorbance measurements
(Short taper, $u_{G,\text{Top}} = 0.9 \text{ m/s}$, 5 injection banks, $L_N = 0 \text{ cm}$, No baffle, $F_S = 0.50 \text{ kg/s}$, $m_{S,\text{Bed}} = 60.6 \text{ kg}$)

In the case of a single tracer method, such as the RPT, the tracer measurement allows direct measurement of the vessel RTD, excluding the recirculation loop. From a mathematical perspective, the measurement is performed in a close-close boundary system, allowing for direct residence time measurements. On the other hand, multiple-particle tracer methods, such as coke dye coating, do not allow this direct measurement. Indeed, when dyed coke is sampled from the bed, it is impossible to know if it had been recirculated zero, one, two or more times. The output tracer concentration measured is the convolution of the coke RTD for various numbers of recirculation passes.

Therefore, it is not possible to directly compare RTD results from RPT and coke dye coating. It either requires a deconvolution of measured dye concentration or a convolution of measured RPT Residence Time Distribution. The latter method was chosen for the sake of computational simplicity. The Residence Time Distribution of the vessel is convoluted
using the scheme presented in Appendix D: Solids RTD validation experiments with dyed coke. Various numbers of convolution pass $N_C$ are computed until the resulting profiles stop to vary with increasing $N_C$. The resulting convoluted RTDs are combined to obtain the overall average convoluted RTD.

### 4.2 Results obtained without corrections

In several previous research studies, the standard CARPT was used to obtain tracer trajectories, but with lower gas velocities, smaller vessel volume, or simpler geometries. Even with these configurations, non-negligible errors on the reconstructed tracer position were observed (Devanathan, 1991; Rammohan et al., 2003). Several works have been conducted to reduce this error (Roy et al., 2001; Degaleesan et al., 2002; Bhusarapu, 2005; Mosorov, 2013).

#### 4.2.1 Choice of RPT rendition method

As presented in Chapter 1 – Section 1.5.4, several RPT rendition techniques exist to determine the tracer coordinates inside the reactor at a given time using the combined radiation signals obtained from the set of scintillation detectors. The two most common methods are the Computer Automated Radioactive Particle Tracking (CARPT) and Monte Carlo simulation methods. The CARPT approach is relatively simple and fast but has only a moderate accuracy if applied without corrections. On the other hand, the Monte Carlo simulation method is much more complex, requires significantly more computing time, and usually provides more accurate results.

Since the datasets acquired during this research ranged from 1 000 000 to 20 000 000 events (an event is defined as the number of counts measured within a sampling time), fast and efficient data processing methods were required. A previous study showed that data processing is 300 times faster with the regular CARPT than with the Monte-Carlo simulation (Sanchez Careaga, 2013). For these reasons, the CARPT method was selected for this research.
For all values of computational time presented in this chapter, the data were acquired using a Lenovo T410, with an Intel® Core™ i5-520M CPU @ 2.40GHz.

4.2.2 CARPT implementation

The following section briefly describes the standard CARPT procedures, including the calibration and rendition phase.

4.2.2.1 Detector calibration

The calibration method used was discussed in Chapter 3 – Section 3.3.3.2. It was performed by placing the radioactive tracer at numerous known positions in the empty unit. It leads to the obtention of calibration curves relating γ-rays counts to distance established for each detector and the whole bed volume (Chaouki et al., 1997). These calibration curves were described by 1D power-law equations, such as presented in Equation (3.3). The detailed calibration procedure is presented in Appendix C: Standard RPT and CARPT calibration performed.

4.2.2.2 CARPT rendition method

The CARPT rendition method describes the set of equations used to convert the recorded time-series of counts measured at each detector into a time series of the 3D coordinates of the radioactive tracer.

4.2.2.2.1 Main CARPT equations

The standard CARPT method was described in Chapter 1 – Section 1.5.4.1. It was developed by Lin et al. (1985) as a weighted linearized regression scheme combining the distances calculated, rearranged in a matrix form. To reduce the importance of the random variations of the radioactive signal, it includes a built-in weighting function giving more importance to the detectors recording the strongest signal at a given time.

4.2.2.2.2 Normalization of counts

In this research, scandium radioactive tracers were chosen to have an initial theoretical strength of 200 μCi. However, due to the small mass of the tracer core (1.5-3 mg), it is
challenging to estimate the mass of scandium to irradiate accurately. Furthermore, since the irradiation time of the scandium core required to reach 200 μCi is directly calculated from its weight, these small mass inaccuracies lead to inaccuracies in initial tracer strength. Therefore, the method introduced by Khanna (2008) for the Monte-Carlo simulations and later expanded to CARPT by Sanchez Careaga (2013) was employed to normalize the counts:

\[ CN_i = \frac{C_i}{\sum_{i=1}^{N_D} C_i} \]  \hspace{1cm} (4.5)

where:

- \( CN_i \): normalized counts for detector i
- \( C_i \): counts for detector i
- \( N_D \): number of detectors.

The normalization of the tracer counts eliminates the need for accurate determination of tracer strength. The tracer can be assumed to be of constant strength throughout the experiments. When applied to 1D calibration curves of the CARPT method, this normalization introduces moderate errors when the tracer is not equidistant from the detectors. Although this additional error is small compared with the standard CARPT accuracies, Section 4.3.1.1.1 discusses how one can eliminate this kind of error when more precise rendition methods are applied, such as in this research.

### 4.2.3 Results from the standard CARPT method

This section presents results from the standard CARPT method. These results are investigated using the selection criteria described in Section 4.1.1 and the validation experiments described in Section 4.1.2.

#### 4.2.3.1 Consistency of RPT results

This section presents and discusses selection criteria results obtained with the RPT without corrections. The selection criteria were introduced in Section 4.1.1.
Table 4-1. Values of selection criteria when the iterative correction and the smoothing correction are not applied.

\( u_G = [0.3; 0.9] \text{ m·s}^{-1}, \) 5 injection banks \( (L_N = 0 \text{ cm}), \) 1 baffle below second injection bank, \( F_S = 0.47 \text{ kg·s}^{-1}, m_{S,Bed} = 60.3 \text{ kg}, \Delta t_{EXP} = 43 \text{ h} \)

<table>
<thead>
<tr>
<th>Parameters used</th>
<th>Original</th>
</tr>
</thead>
<tbody>
<tr>
<td>Proportion of positions detected out of bed</td>
<td>14.18%</td>
</tr>
<tr>
<td>Proportion of artificially low tracer presence inside the bed</td>
<td>0.250%</td>
</tr>
<tr>
<td>Average value of moving standard deviation on coordinates ( X_y_z(s) - \Delta t_{MOV} = 2 \text{ s} )</td>
<td>[0.0629; 0.0459; 0.1294]</td>
</tr>
<tr>
<td>Computational time required (ms/event)</td>
<td>0.0323</td>
</tr>
</tbody>
</table>

Table 4-1 presents the selection criteria results obtained with the RPT without corrections. The experimental conditions used were: \( u_{G,Top} = 0.9 \text{ m·s}^{-1}, \) five injection banks, \( L_N = 0 \text{ cm}, \) one baffle (no flux-tubes) located below the second injection bank, \( F_S = 0.47 \text{ kg·s}^{-1}, m_{S,Bed} = 60.3 \text{ kg}, D_{Agg} = 1.3 \text{ cm}, \rho_{Agg} = 1100 \text{ kg·m}^{-3}, \Delta t_{EXP} = 43 \text{ hours}. \) The average time interval between two measured radioactive counts was 11.7 ms. It should be noted that only the proportion of positions detected out of bed and the computational time required have an actual meaning as absolute values. The proportion of empty in-bed volume and the average value of moving standard deviation on coordinates are only helpful as a reference to study their relative variations when the correction(s) is/are applied.

The first two selection criteria, the proportion of positions detected out of bed, and the proportion of empty in-bed volume, can be better visualized using Figure 4-7.
Figure 4-7. Tracer positions repartition
(Short taper, $u_{G,\text{Top}} = 0.9 \text{ m/s}$, 5 injection banks, $L_N = 0 \text{ cm}$, 1 baffle below 2nd injection,
$F_S = 0.47 \text{ kg/s}$, $m_{S,\text{Bed}} = 60.3 \text{ kg}$, $D_{\text{Agg}} = 1.3 \text{ cm}$, $\rho_{\text{Agg}} = 1100 \text{ kg/m}^3$, $\Delta t_{\text{EXP}} = 43 \text{ h}$)

Figure 4-7 shows that the amount and repartition of impossible tracer positions (located outside the bed) in the RPT measurements without correction are clearly visible. This number of impossible locations is significant, which justifies the need for RPT corrections. It should be noted that most of the impossible tracer positions are located at heights between the lowest and highest lateral injection banks, a zone with strong radial motion and fast tracer motion. The amount and repartition of empty in-bed volume are also visible in Figure 4-7: no large empty zones within the bed are visible.

The third selection criterion, the average value of moving standard deviation on coordinates, can be better visualized using Figure 4-8.
$b$)

![Graph showing raw data with annotations and highlighted sections.]

- $Y$ (m) vs. $t$ (s)
- Highlighted sections indicating specific time intervals.
Figure 4-8. Example of cartesian coordinates obtained without CARPT correction
(Short taper, $u_{G, \text{Top}} = 0.9 \, \text{m/s}$, 5 injection banks, $L_N = 0 \, \text{cm}$, 1 baffle below 2\textsuperscript{nd} injection,
$F_S = 0.47 \, \text{kg/s}$, $m_{S, \text{Bed}} = 60.3 \, \text{kg}$, $D_{Agg} = 1.3 \, \text{cm}$, $\rho_{Agg} = 1100 \, \text{kg/m}^3$, $\Delta t_{\text{EXP}} = 43 \, \text{h}$)

a) $x$-coordinates, b) $y$-coordinates, c) $z$-coordinates

In Figure 4-8, the strong fluctuations of coordinates resulting from inaccuracies in radioactive counts rate detected are visible (for $t < 4 \, \text{s}$). These strong fluctuations are problematic when investigating velocities profiles or tracer flow patterns. Therefore, this is another argument in favour of the need for RPT correction(s).
4.2.3.2 Comparison of RPT results with other measurement methods

This section presents critical results obtained with the RPT without corrections and compares them with other independent methods, such as introduced in Section 4.1.2.

4.2.3.2.1 Tracer time distribution

A comparison was conducted with results obtained from dyed coke experiments to verify the radioactive tracer Residence Time Distribution (RTD) obtained with RPT. The details of this comparison are provided in Section 4.1.2.1 and Appendix D: Solids RTD validation experiments with dyed coke.

![Figure 4-9. Convoluted top to stripper RPT time distribution](image)

Figure 4-9 presents the convoluted Top-to-Stripper RPT time distributions for two range densities of agglomerates of tracer: “heavy (1100-1200 kg·m⁻³) and “light” (850-950 kg·m⁻³). The first category represents typical wet-agglomerates, while the second one represents solids in bed emulsion. The details of the convolution procedure are presented in Appendix D: Solids RTD validation experiments with dyed coke.
The results in Figure 4-9 show a moderate difference in Top-to-Stripper RPT time distributions between the “heavy” and “light” tracer: the heavy tracer travels faster through the bed, indicating that increasing the density of the tracer reduces its residence time.

Figure 4-10 presents a comparison between these convoluted Top-to-Stripper RPT time distributions and the measured absorbance of sampled coke (integrated over a selected wavelength range, see Appendix D: Solids RTD validation experiments with dyed coke) over time.

![Figure 4-10](image)

**Figure 4-10.** Comparison of dyed coke residence time profile and convoluted top-to-stripper RPT time distribution

(Short taper, \( u_{G,\text{Top}} = 0.9 \text{ m}\cdot\text{s}^{-1} \), 5 injection banks, \( L_N = 0 \text{ cm} \), \( F_S = 0.50 \text{ kg}\cdot\text{s}^{-1} \), \( D_{Agg} = 1.3 \text{ cm} \), \( \rho_{Agg} = [850; 1200] \text{ kg}\cdot\text{m}^{-3} \))

The results in Figure 4-10 indicate that the RPT measurements give realistic residence time distribution measurements, especially for coke particles in the bed emulsion. It also confirms that wet-agglomerates (and tracers used to simulate them) have a moderately shorter relative residence time distribution than bed emulsion. It should also be noted that
the scatter in the results observed with the dye method confirms that there is a random component to solids movement through the unit. The impact of the RPT correction on these results will be investigated in Section 4.4.3.1.

4.3 RPT enhancement methods

Radioactive Particle Tracking (RPT) is used to track the position of a radioactive tracer over time (Lin et al., 1985; Sanchez Careaga, 2013). The RPT method suffers from several sources of inaccuracies. First, the number of counts detected by any scintillation detector for a given tracer position located inside the vessel can vary because of the random nature of the decay of an unstable atom, described by a Poisson distribution (Leo, 1994; Holbert, 2002). A random error in the location of the tracer-agglomerate is always expected. As presented in Chapter 3 – Section 3.3.3, the error fraction in the measured counts rate $\sigma/CRM$ increases with decreasing radioactive source strength and decreasing sampling time. The RPT methods also suffer from inaccuracies related to the calibration method chosen. When using a simple 1D calibration curve, the source-detector angle is not considered, increasing error. Besides calibration being conducted in an empty vessel, the bed absorption is not considered. Finally, bed absorption varies with distance from source to detector, and bed elevation; it can also fluctuate with time due to fluctuations in local bed voidage resulting from rising gas bubbles.

Several methods are investigated to reduce RPT inaccuracies due to systematic error from calibration and random error.

4.3.1 Absorption correction

The standard CARPT does not account for variations in the adsorption of the $\gamma$-rays reaching each detector:

- As mentioned in Chapter 3, a significant limitation of the CARPT approach is that the calibration 1D correlations ignore the effect of the solid angle between the tracer and a horizontal plane through the detector’s virtual center. The angle at which the $\gamma$-rays cross the wall influences their travel length through the wall and, hence, the wall absorption of the radiation.
Another significant limitation is that the calibration is conducted in an empty bed: the bed absorption occurring during measurements is not incorporated into the calibration curve. Besides, due to variable voidage, this bed absorption varies not only with the distance from source to detector but also with bed elevation and with time.

This section describes a method applied to correct these effects to obtain more precise tracer positions. This method builds a 3D grid of re-calculated counts (for each detector) and performs a similarity search to determine a re-calculated tracer position in the grid. Two distinct steps are required in the method: 1) build a 3D grid of re-calculated counts accounting for absorption for each detector, 2) perform a similarity search in the constructed grids to match a measured count value with a re-calculated position.

Figure 4-11. Simplified flowchart of the absorption correction method

The first step, building a 3D grid of re-calculated counts accounting for absorption for each detector, can be performed through various methods of variable complexity, ranging from simple 1D calibration curves, to complex Monte-Carlo simulations, to black-box neural model. This research developed a new hybrid approach aiming to offer higher performances than 1D curves without the computational cost of Monte-Carlo simulations.
The second step, performing a similarity search in the constructed grids to match a measured count value with a re-calculated position, used an iterative approach developed in previous research (Mosorov, 2013). This iterative method was modified to achieve faster processing and higher accuracy.

A simplified flowchart of the absorption correction method is presented in Figure 4-11.

4.3.1.1 Create a 3D grid of re-calculated counts accounting for absorption

To re-calculate the tracer position while accounting for radiation absorption, the first step is to build a 3D grid of re-calculated counts accounting for each detector’s absorption. A new hybrid approach was developed to bridge the gap between the simplicity of the reverse CARPT calibration curves (based on a 1D calibration in an empty vessel) and the high complexity of the Monte Carlo simulation (where counts at a given position are calculated using an exact depiction of the various mechanisms involved in radiation emission, dispersion, absorption, and detection).

This hybrid approach is a 2-step process: 1) populate the grids with re-calculated counts obtained at any node within a 3D grid with the 1D calibration curves (created in an empty vessel), 2) correct these re-calculated counts based on the theoretical absorption coefficients calculated based on the media traversed by the radiation (thickness, densities) and the energy spikes emitted.

4.3.1.1.1 Re-calculated counts from reverse CARPT 1D calibration curves

To build a 3D grid of re-calculated counts accounting for absorption, each node within the grid (a node being defined as a point of the 3D grid at a given position \(X_j = [x_j; y_j; z_j]\)) needs to be associated with a re-calculated count value \(CC(X_j)\). Each detector \(i\) requires its own 3D grid of counts.

Chapter 3 – Section 3.3.3.2 shows the calibration 1D equations obtained in the empty vessel and used to find a distance source-detector were described (Equation (3.3)). To populate
the nodes of the 3D grid with calculated counts number \( CC_i(X_j) \), the simplest method is to use similar equations but reversed, transforming distances source-detector \( i \) into counts measured at the detector \( i \). A significant difference is the use of calibration curves with non-normalized counts. As mentioned in Section 3.3.3, the counts normalization was introduced by Khanna (2008) for Monte-Carlo simulations, and later extended to CARPT methods by Sanchez Careaga (2013), to avoid considering the source age in the RPT calculations. When used with CARPT, a drawback is that it introduces additional errors in the 1D calibration curves. Indeed, for a given detector \( i \), if the source is located at two positions \( j \) at the same distance \( L_{ij} \) from the detector, the counts values measured at each position are the same (excluding random errors). However, the normalized counts values can be different depending on the positions of the other detectors located around detector \( i \). The error introduced is due to a 3D normalization projected onto 1D calibration curves. This error is moderate and is more critical for detectors with neighbours unevenly arranged around.

Therefore, the reverse 1D CARPT calibration curves used to populate the grid can be described in Equation (4.6):

\[
CC_i(X_j) = A'_i \left(L_{ij}\right)^{B'i}
\]

where:

- \( CC_i(X_j) \): counts detected at detector \( i \), from node \( j \)
- \( X_j \): position of node \( j \), such as \( X_j = \begin{bmatrix} x_j \\ y_j \\ z_j \end{bmatrix} \)
- \( L_{ij} \): Euclidian distance between the detector \( i \) and the grid node \( j \)
- \( A'_i \) & \( B'_i \): calibration coefficients of detector \( i \)

An example of a calibration curve for reversed CARPT counts calculations is presented in Figure 4-12.
Each of the twelve detectors has its own set of reversed calibration equations. For detector $i$, the parameter $B'_i$ is constant over time, and the parameter $A'_i$ varies over time. The equations obtained can be compared to the theoretical 3D dispersion of radiation, using the Inverse Square Law described in Equation (3.7). The theoretical equations give normalized counts $CN_i$ proportional to the distance $L_i^{-2}$. The exponents obtained by calibration range within $[-1.4312; -2.1934]$. One of the leading causes for the variation between the theoretical exponent and the ones obtained through correlation is due to the fact that the calibration correlations ignore the effect of the angle between the tracer and a horizontal plane through the virtual center of the detector. As the calibration is conducted within the empty vessel, the 1D curves aggregate variable values of steel wall absorption (due to taps, variable intersection length when the angle varies, wall irregularities). Another important contributing factor is the variation between the crystal properties of each detector. Besides, the calibration is done in an empty vessel: the significant radiation attenuation due to the bed particles is not considered (only wall absorption considered).

**Figure 4-12.** Typical calibration curves for reversed CARPT counts re-calculations for one of the twelve detectors used in this research
Therefore, a correction procedure based on the theoretical absorption coefficients calculated based on the media traversed by the radiation (thickness, densities) and the energy spikes emitted.

These counts $CC_i(X_j)$ are a function of the age of the tracer, whose $\gamma$-ray emissions weaken with age. Therefore, an attenuation coefficient $\zeta$ is required to obtain the real counts value at the measurement time, a function of: the age of the tracer at the calibration time $t_C$, the original strength of the tracer used for calibration $C_{0,C}$, the age of the tracer at the measurement time $t_M$, and original strength of the tracer used for measurement $C_{0,M}$. All these parameters need to be quantified precisely to ensure an accurate correction. The correction can be expressed such as described in Equation (4.7):

$$CC_i(X_j, t_{MEAS}) = \zeta(C_{0,CAL}, C_{0,MEAS}, t_{CAL}, t_{MEAS}) \cdot CC_i(X_j, t_{CAL})$$

(4.7)

with:

- $CC_i(X_j, t_{MEAS})$: counts detected at detector $i$, from node $j$, at measurement time
- $CC_i(X_j, t_{CAL})$: counts detected at detector $i$, from node $j$, at calibration time
- $\zeta$: radiation attenuation coefficient over time
- $X_j$: position of node $j$, such as $X_j = \begin{bmatrix} x_j \\ y_j \\ z_j \end{bmatrix}$ (m)
- $C_{0,MEAS}$: original strength of the tracer used for measurement
- $C_{0,CAL}$: original strength of the tracer used for calibration
- $t_{MEAS}$: age of the tracer at the measurement time (s)
- $t_{CAL}$: age of the tracer at the calibration time (s)

Therefore, after populating the 3D grids with the counts $CC_i(X_j)$, they are normalized to remove the effect of the age of the tracer. The normalization performed at this stage does not introduce the additional errors mentioned in the previous paragraph since the normalization is applied on the 3D grids directly and not on the 1D calibration curves. The normalization equation is described in Equation (4.8):

$$CCN_i(X_j) = \frac{\zeta(C_{0,C}, C_{0,M}, t_C, t_M) \cdot CC_i(X_j)}{\sum_{i=1}^{ND} \zeta(C_{0,C}, C_{0,M}, t_C, t_M) \cdot CC_i(X_j)} = \frac{CC_i(X_j)}{\sum_{i=1}^{ND} CC_i(X_j)}$$

(4.8)
where:

- \( \text{CC}_{i}(X_j) \): normalized counts detected at detector \( i \), from node \( j \)
- \( \text{CC}_i(X_j) \): counts detected at detector \( i \), from node \( j \)
- \( N_D \): total number of detectors
- \( X_j \): position of node \( j \), such as \( X_j = [x_j, y_j, z_j] \)

This procedure allows for the creation of 3D grids of normalized counts, independent of the age of the tracer and which can be compared with measured normalized counts at any time \( t \).

### 4.3.1.1.2 Correction of re-calculated counts of the 3D grids

The objective was to use the theoretical radiation absorption to correct the re-calculated counts populating the 3D grids. This absorption depends on the distance within the various media between the radioactive source and the radiation sensor. In the experimental unit, these media include: air, steel wall and coke particles.

The ratio of sub-incident flux over the incident flux at \( L_{j,0} \), \( \Phi_{S,j}/\Phi_{0,j} \), due to radiation absorption through any medium traversed can be expressed, within the given medium \( j \), by a generalized Beer-Lambert law, such as the one presented in Equation (4.9).

\[
\frac{\Phi_{S,j}}{\Phi_{0,j}} = \exp\left(-\alpha_j \cdot \rho_j \cdot L_j\right)
\]  (4.9)

with:

- \( \Phi_{S,j} \): sub-incident flux, after crossing the medium \( j \)
- \( \Phi_{0,j} \): incident flux, before crossing the medium \( j \)
- \( L_j \): absorbing medium \( j \) thickness (m). It is the effective distance crossed through the medium \( j \).
- \( \rho_j \): absorbing medium \( j \) density (kg·m\(^{-3}\))
- \( \alpha_j \): mass absorption coefficient of the medium \( j \), for a given radiation energy spike (m\(^2\)·kg\(^{-1}\))
Radioactive scandium emits γ-ray at two energy spikes: 0.889 MeV and 1.173 MeV (Helmer and Kuzmenko, 2004). Therefore, for any medium $i$, there will be two mass absorption coefficients $\alpha_i$ and two absorption curves. To simplify, all calculations will be presented using results from an average mass absorption coefficient $\alpha_{AVG,j}$. The relevant values of the mass absorption coefficient are shown in Table 4-2, along with the densities $\rho_i$.

There are three zones of absorption in the experimental system shown in Figure 4-13: the fluidized bed between the source and the inner wall of the vessel, the vessel wall (steel), and the air gap between the outer vessel wall and the detector. Since the fluidized bed is composed of coke solids fluidized in air, three different media absorb radiations between any of the twelve detectors and any source position within the bed: air, steel walls, and coke.

Figure 4-13. Absorption model in the experimental unit
Table 4-2. Media densities & γ-ray mass attenuation coefficients

<table>
<thead>
<tr>
<th>ρj (kg·m⁻³)</th>
<th>αj (m²·kg⁻¹)</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Emission</td>
<td>Emission</td>
</tr>
<tr>
<td></td>
<td>peak 1</td>
<td>peak 2</td>
</tr>
</tbody>
</table>

Therefore, the overall absorption can be expressed as:

\[
\frac{\Phi_S}{\Phi_0} = \exp(-\alpha_{\text{AVG,bed}} \cdot \rho_{\text{bed}} \cdot L_{\text{bed}}) \cdot \exp(-\alpha_{\text{AVG,steel}} \cdot \rho_{\text{steel}} \cdot L_{\text{steel}}) \cdot \exp(-\alpha_{\text{AVG,air}} \cdot \rho_{\text{air}} \cdot L_{\text{air,out}}) = \exp(-\alpha_{\text{AVG,air}} \cdot \rho_{\text{air}} \cdot L_{\text{air, in}}) \cdot \exp(-\alpha_{\text{AVG,coke}} \cdot \rho_{\text{coke}} \cdot L_{\text{coke}}) \cdot \exp(-\alpha_{\text{AVG,air}} \cdot \rho_{\text{air}} \cdot L_{\text{air,out}})
\]

(4.10)

We can simplify Equation (4.10). First, for the range of distances considered in this research (up to 1.5m), the air absorption is neglectable, as shown in Figure 4-14.

Besides, the absorption distance in coke is a direct function of the distance in bed L_{\text{bed}} and the bed voidage ε_{\text{bed}}. Therefore, Equation (4.10) becomes:

\[
\frac{\Phi_S}{\Phi_0} = \exp(-\alpha_{\text{AVG,coke}} \cdot \rho_{\text{coke}} \cdot (1 - \epsilon_{\text{bed}}) \cdot L_{\text{bed}}) \cdot \exp(-\alpha_{\text{AVG,steel}} \cdot \rho_{\text{steel}} \cdot L_{\text{steel}})
\]

(4.11)

The absorption related to the steel walls is only a function of the thickness of steel traversed L_{\text{steel}}. The evolution of this absorption as a function of L_{\text{steel}} is presented in Figure 4-15. The absorption related to the bed itself is a function of the distance between the source and the wall L_{\text{bed}} and a function of the bed voidage ε_{\text{bed}}. The evolution of this absorption as a function of L_{\text{bed}} and ε_{\text{bed}} is presented in Figure 4-16.
Figure 4-14. Radiation absorption in air

Figure 4-15. Radiation absorption in steel wall as a function of wall effective thickness $L_{\text{steel}}$
It should be noted that since calibration equations were obtained in an empty bed, as described in Section 4.2.2.1, the steel wall absorption is included but averaged into a 1D calibration. Therefore, the steel wall absorption correction is relative and applied over the average steel wall absorption at a distance source-detector $L$.

Considering everything discussed above, if the bed voidage $\varepsilon_{\text{bed}}$ is constant, the coefficient correction for the re-calculated counts obtained with 1D calibration curves (in an empty vessel) and associated to sensor $i$, can be expressed as:

$$
\frac{CC_{\text{Grid},i}(x, y, z)}{CC_{\text{Cal1D},i}(x, y, z)} = 
\exp\left(-\alpha_{\text{AVG}} \cdot \rho_{\text{coke}} \cdot (1 - \varepsilon_{\text{bed}}) \cdot L_{\text{Bed},i}(x, y, z)\right) \cdot \frac{\exp\left(-\alpha_{\text{AVG}} \cdot \rho_{\text{Steel}} \cdot L_{\text{SteelWall},i}(x, y, z)\right)}{\exp\left(-\alpha_{\text{AVG}} \cdot \rho_{\text{Steel}} \cdot L_{\text{SteelWall,AVG},i}(L)\right)} = (4.12)
$$

$$
\exp\left(-\alpha_{\text{AVG}} \cdot \rho_{\text{coke}} \cdot (1 - \varepsilon_{\text{bed}}) \cdot L_{\text{Bed},i}(x, y, z)\right) \cdot C_{\text{SteelWall},i}(x, y, z)
$$

with:

$$
C_{\text{SteelWall},i}(x, y, z) = \frac{\exp\left(-\alpha_{\text{AVG}} \cdot \rho_{\text{Steel}} \cdot L_{\text{SteelWall},i}(x, y, z)\right)}{\exp\left(-\alpha_{\text{AVG}} \cdot \rho_{\text{Steel}} \cdot L_{\text{SteelWall,AVG},i}(L)\right)}
$$

For any position $(x, y, z)$, simple geometric relationships (such as Euclidean norms or intersections of line and cylinders) can be used to calculate $L_{\text{bed},i}$, $L_{\text{SteelWall},i}$ and $L_{\text{SteelWall,AVG},i}$ (and therefore the steel wall absorption correction $C_{\text{SteelWall},i}$). However, it is common knowledge that the bed voidage $\varepsilon_{\text{bed}}$ varies within a fluidized bed, both with space and time: integration is required to obtain accurate results.

Figure 4-17 represents a schematic of the geometry associated with bed absorption calculations.
Figure 4-16. Radiation absorption in bed as a function of
a) distance source-wall $L_{\text{bed}}$, b) bed voidage $\varepsilon_{\text{bed}}$
Figure 4-17. Geometry associated with bed absorption calculations

To avoid the calculations of complex 3D integrals varying with time, two approximations are selected:

- Use of the time-averaged bed voidage: $\varepsilon_{bed}(x, y, z, t) \sim \varepsilon_{bed}(x, y, z)$
- Variation of bed voidage within a 2D plan located at a height $z$, the bed voidage is assumed constant and equal to the average bed voidage at this height: $\varepsilon_{bed}(x, y, z) \sim \varepsilon_{bed}(z)$

With these assumptions, Equation (4.12) can be rewritten as:
\[
\frac{CC_{\text{Grid},i}}{CC_{\text{Cal1D},i}}(x,y,z) = \exp\left(-\alpha_{\text{AVG}} \cdot \rho_{\text{coke}} \cdot (1 - \varepsilon_{\text{bed}}(z)) \cdot L_{\text{Bed},i}(x,y,z) \right) \cdot C_{\text{SteelWall},i}(x,y,z) \quad (4.13)
\]

Using Figure 4-17, \(L_{\text{Bed},i}\) can be expressed as:

\[
L_{\text{Bed},i}(x,y,z) = \frac{H_{\text{Bed},i}(y)}{\sin(\alpha_{\text{Bed},i}(x,y,z))} \quad (4.14)
\]

The solid angle \(\alpha_{\text{bed}}\) between the ray-wall intersection and the source (node) position can be given as:

\[
\alpha_{\text{Bed},i}(x,y,z) = \arctan\left(\frac{H_{\text{Bed},i}(z)}{L'_{\text{Bed},i}(x,y)}\right) \quad (4.15)
\]

Besides, we can express the distance \(L'_{\text{Bed}}\) as:

\[
L'_{\text{Bed},i}(x,y) = \sqrt{L_{x,\text{Bed},i}(x)^2 + L_{y,\text{Bed},i}(y)^2} \quad (4.16)
\]

Therefore Equation (4.15) becomes:

\[
\alpha_{\text{Bed},i}(x,y,z) = \arctan\left(\frac{H_{\text{Bed},i}(z)}{\sqrt{L_{x,\text{Bed},i}(x)^2 + L_{y,\text{Bed},i}(y)^2}}\right) \quad (4.17)
\]

Therefore, we can integrate Equation (4.13) such as:

\[
\frac{CC_{\text{Grid},i}}{CC_{\text{Cal1D},i}}(x,y,z) = 
\left(\int_{z_{\text{cross}}}^{z_{\text{node}}} \exp\left(-\alpha \cdot \rho_{\text{p}} \cdot (1 - \varepsilon_{\text{bed}}(z)) \cdot L_{\text{Bed},i}(x,y,z) \right) dL_{\text{Bed},i}(x,y,z) \right) \cdot C_{\text{SteelWall},i}(x,y,z) \quad (4.18)
\]

with:

- \(L_{\text{Bed},i}(x,y,z) = \frac{z}{\sin(\alpha_{\text{Bed}})}\)
- \(dL_{\text{Bed},i}(x,y,z) = \frac{dz}{\sin(\alpha_{\text{Bed},i})}\)
- \(z = H_{\text{Bed},i} = |z_{\text{node}} - z_{\text{cross},i}|\)

Therefore, Equation (4.18) becomes:
\[
\frac{CC_{Grid,i}}{CC_{Cal1D,i}}(x, y, z) = \\
\left( \int_{z_{cross}}^{z_{node}} \exp\left(-\alpha \rho_p (1 - \varepsilon_{bed}(z)) \right) \frac{z}{\sin(\alpha_{Bed,i})} \frac{dz}{\sin(\alpha_{Bed,i})} \right) \cdot C_{SteelWall,i}(x, y, z) = \quad (4.19)
\]

Finally, the following 3D correction equation is defined:

\[
\frac{CC_{Grid,i}}{CC_{Cal1D,i}}(x, y, z) = \\
\left( \int_{z_{cross}}^{z_{node}} \exp\left(-\alpha \rho_p (1 - \varepsilon_{bed}(z)) \right) \frac{1}{\sin(\alpha_{Bed,i})} \frac{dz}{\sin(\alpha_{Bed,i})} \right) \cdot C_{SteelWall,i}(x, y, z) \quad (4.20)
\]

It should be noted that when the tracer is in the same horizontal plane as the detector \(i\) \(\alpha_{Bed,i} = 0\), then Equation (4.13) should be used instead of Equation (4.20).

For detector \(i\), the correction coefficient \(\frac{CC_{Grid,i}}{CC_{Cal1D,i}}(x, y, z)\) can be calculated for any node position \(X = \begin{pmatrix} x \\ y \\ z \end{pmatrix}\) within the 3D grid. Figure 4-18 shows the bed absorption correction, the relative wall absorption correction, and the whole absorption connection for one detector (Detector 6) and one plane \(y = 0\).

Figure 4-19 presents the absorption correction maps for various detectors within the same plane \(y = 0\). Figure 4-20 shows the absorption correction maps for one detector within different planes \(y\).

This new method allows for a fast generation of a dense grid. For instance, in this research, a 3D grid covering the whole bed \(z \in [0; 1.5] \text{ m}, r \in [-0.13; +0.13] \text{ m}\) and with an internode distance of 1 cm had 73 523 nodes within the bed volume. Once the 1D reversed-CARPT calibration curves had been obtained, all the 3D grids for the twelve detectors were generated entirely in 10-15 s.
Figure 4-18. Absorption correction coefficient (y = 0, Detector 6)

a) 3D plane & detectors position (top view), b) Bed absorption correction, c) Relative wall absorption correction, d) Whole absorption correction
Figure 4-19. Absorption correction coefficient (y = 0)

a) 3D plane & detectors position (top view), b) Whole absorption correction for detector 1, c) Whole absorption correction for detector 4, d) Whole absorption correction for detector 7, e) Whole absorption correction for detector 10
Figure 4-20. Absorption correction coefficient (Detector 6)

a) 3D planes & detector position (top view), b) Whole absorption correction in plane $y = -8.5$ cm, c) Whole absorption correction in plane $y = 0$ cm, d) Whole absorption correction in plane $y = +8.5$ cm
4.3.1.2 Obtain re-calculated positions from 3D grids of re-calculated counts

The iterative search algorithm developed by Mosorov (2013) was adapted to calculate the corrected tracer position. This algorithm was developed as an extension of the position reconstruction algorithms using a least-square approach search. A typical least square approach reconstructs the tracer position as the node \( j \) of the 3D grid, at the position \( X_j \), which minimizes \( \chi^2 \) such as (Larachi et al., 1994; Doucet et al., 2008):

\[
\chi^2 = \sum_{i=1}^{N_D} \left( \frac{CCR_i(X_j) - CR_i(X)}{\sigma_{Rad,i}^2} \right)^2
\]

where:

- \( CR_i(X) \): measured counts rate for detector \( i \)
- \( CCR_i(X_j) \): calculated counts rate for detector \( i \), at node \( j \) in a 3D grid
- \( X \): (unknown) tracer position
- \( X_j \): position of node \( j \) in a 3D grid
- \( N_D \): total number of detectors
- \( \sigma_{Rad,i} \): statistical uncertainty or the standard deviation, \( \sigma_{Rad,i} = \sqrt{CCR_i(X_j)} \)

This reconstruction algorithm presents some drawbacks: the resolution of the calibration grid limits its accuracy. Therefore, the fewer the number of points used in the 3D grid, the poorer the reconstruction accuracy. Besides, this number of points cannot be increased significantly due to the intensive computation power required to perform the minimization procedure on a considerable number of data points (Mosorov, 2013).

As mentioned before, a significant limitation in the determination of the tracer position is related to the random error in counts measurements: the statistical fluctuations of the number of photons emitted by a source, their dispersion, the sensor-photon interaction, and the finite measurement time interval can significantly influence counts measured, reducing the tracer reconstruction precision (Leo, 1994; Holbert, 2002; Mosorov, 2013). These fluctuations in radiation detection can be described as stochastic processes with a Poisson probability distribution and are not correlated with each other. Therefore, using a procedure
including an averaging operation of more than one calculated position will reduce their contribution to the counts measured (Mosorov, 2013).

4.3.1.2.1 Iterative search algorithm

The iterative algorithm can be described as follows (Mosorov, 2013).

The location of the tracer within the vessel, in the Cartesian coordinates, is $X = \begin{bmatrix} x \\ y \\ z \end{bmatrix}$. Any position at one of the $N_N$ nodes of the 3D grid created is located at $X_j (j \in [1; N_N])$.

The measured counts rate is $CR_i(X)$, for detector $i$, obtained when the tracer is at the unknown position $X$. In the same way, at any position at one of the $N_N$ node of the 3D grid, let us define the corrected counts rate $CCR_i(X_j)$, for detector $i$, obtained at 3D grid position $X_j (j \in [1; N_N])$.

For each measured event, the measured counts rate $CR(X)$ are processed through CARPT to back-calculate the position $X$. This initial guess of tracer position is later referred to as $X_0$.

The $N_N$ similarities $s_j (j \in [1; N_N])$ between measured counts rate $CR(X_{i,0})$ and corrected counts rate at each node of the 3D grid $CCR(X_{i,j})$, for detector $i$, are calculated such as:

$$s_j = \sum_{i=1}^{ND} \left| \ln (CCR(X_{i,j})) - \ln (CR(X_{i,0})) \right| \quad (4.22)$$

The vector $S = \begin{bmatrix} S_1 \\ S_2 \\ \cdots \\ S_{(N_N-1)} \\ S_{N_N} \end{bmatrix}$ is sorted in increasing order to obtain the sorted similarities vector $S_{SORT} = \begin{bmatrix} s_{sort,1} \\ s_{sort,2} \\ \cdots \\ s_{sort,(N_N-1)} \\ s_{sort,N_N} \end{bmatrix}$, with $s_{sort,1} \leq s_{sort,2} \leq \cdots \leq s_{sort,(N_N-1)} \leq s_{sort,N_N}$. 

For each iteration step $k$ ($k \geq 1$), the calculated tracer positions $X_k = \begin{bmatrix} x_k \\ y_k \\ z_k \end{bmatrix}$ is defined such as:

$$
x_k = \sum_{j=1}^{k} x(s_{sort,j}) \cdot w_j
$$
$$
y_k = \sum_{j=1}^{k} y(s_{sort,j}) \cdot w_j
$$
$$
z_k = \sum_{j=1}^{k} z(s_{sort,j}) \cdot w_j
$$

with:

- $x(s_{sort,j})$, $y(s_{sort,j})$ & $z(s_{sort,j})$: cartesian coordinates of the node with resulting in the $j^{th}$ sorted similarity.
- $w_j$: normalized weight ($\sum_{j=1}^{k} w_j = 1$) determining the contribution of sorted $j^{th}$ node position in the reconstructed tracer position, defined such as:

$$
w_j = \frac{s_{sort,1}}{\sum_{j=1}^{k} s_{sort,j}}
$$

This averaging step reduces the influence of the contribution of the statistical fluctuations through the averaging operation (Mosorov, 2013).

The calculated tracer positions $X_{k+1} = \begin{bmatrix} x_{k+1} \\ y_{k+1} \\ z_{k+1} \end{bmatrix}$ is calculated in the same way.

The iteration process is continued, cyclically increasing $k$, as long as the linear Euclidean distance $\|X_k \cdot X_{(k+1)}\|$ between $k^{th}$ calculated tracer position $X_k$ and $(k + 1)^{th}$ calculated tracer position $X_{(k-1)}$ remains larger than a chosen threshold value $\epsilon$.

$$
\|X_k \cdot X_{(k+1)}\| \leq \epsilon
$$
**TRUE:** Tracer at position $X_k$

**FALSE:** Iterate $k = k + 1$ & repeat

The selection of the threshold value $\epsilon$ is discussed in Section 4.3.1.3.1.1.

In the algorithm developed by Mosorov (2013), the map of calculated counts $\text{CCR}_j(X_j)$ was obtained using a Monte-Carlo simulation approach using the MCNP5 code (X-5 Monte Carlo Team, 2003; Mosorov and Abdullah, 2011) with 2 205 grid nodes. This limited number of nodes was due to the computational complexity inherent to the use of this method. In this research, the simplified approach is used, which will be described in Section 4.3.1.2.2.

It should be noted that Bhusarapu (2005) developed a relatively similar approach, the Cross-Correlation Based Search Algorithm. Nevertheless, this method still required to measure in-situ (i.e. at the operating conditions of interest) calibration data points (275 points in Bhusarapu (2005) work) as they are required to obtain an accurate first estimate of data position. It makes the Cross-Correlation Based Search Algorithm challenging to scale and limits the range of conditions that can be studied without requiring a re-calibration. On the other hand, the approach developed in this work aims to use empty-vessel calibration data.

### 4.3.1.2.2 Modifications of the iterative search algorithm

As mentioned earlier, the averaging step performed with Equation (4.23) reduces the influence of the contribution of the statistical fluctuations through the averaging operation. To take advantage of this feature, a minimum number of iterations $N_{\text{MIN}}$ is specified.

The selection of $N_{\text{MIN}}$ is discussed in Section 4.3.1.3.1.1.

A significant limitation of the algorithm is its computational complexity. To obtain a 3D grid fine enough, the space between the nodes needs to be small, requiring a high number of nodes $N_N$. Equation (4.22) requires creating one similarities vector of size $[N_N, 1]$ per event reconstructed. This vector is then sorted. Therefore, for a dataset including $N_E$ events, $N_E$ vectors of size $[N_N, 1]$ need to be created and sorted. In this research, for a 3D grid
covering the whole bed \((z \in [0; 1.5] \text{ m}, r \in [-0.13; +0.13] \text{ m})\) and with an internode distance of 1 cm, it results in a grid of 73,523 nodes within the bed volume.

A simple solution to speed up the calculation is to limit the number of nodes investigated to create the vector of similarities \(S_{\text{SORT}}\). Assuming that the initial position calculated with CARPT using 1D calibration curves gives a position relatively accurate, the iterative search can be limited to a number of in-bed nodes \(N_S\) \((N_S \leq N_N)\) within a volume centred around this initial position. All the equations described before can be used with \(N_S\) replacing \(N_N\).

The choice of \(N_S\) is critical since the computational time increases significantly when the number of search nodes \(N_S\) increases. On the other hand, the iterative search needs to include a big enough number of in-bed nodes \(N_S\) to converge properly. If the subset of nodes selected is too small, the algorithm will converge towards a local minimum rather than the actual reconstructed tracer position. Therefore, a minimum number of in-bed search nodes \(N_{S,\text{MIN}}\) is specified, such as \(27 \leq N_{S,\text{MIN}} \leq N_S \leq N_N\) (27 nodes corresponding to the volume of a search volume of +/- one node in each direction).

To select these \(N_S\) nodes, a first estimation of the tracer position is obtained using the regular CARPT procedure (as described in Chapter 3). Then, a 3D volume is defined around this position to form a rectangular prism of height \(H_S\), length \(L_S\), and depth \(D_S\), such as its volume is \(V_S = H_S \cdot L_S \cdot D_S\). To take into consideration the fact that the tracer position varies more along the z-axis than along the x-axis and y-axis, the height \(H_S\) is set as \(H_S = 2 \cdot L_S = 2 \cdot D_S\). When considering the first estimation of the position of the tracer, three situations can occur: 1) the first estimation of the tracer position is far away from the bed walls, a volume of \(N_S\) in-bed search nodes \((N_{S,\text{MIN}} \leq N_S)\) can be obtained within the rectangular prism of height \(H_S\), length \(L_S\) and depth \(D_S\), 2) the first estimation of the tracer position is close from the bed walls, the length of \(H_S\), \(L_S\) & \(D_S\) is increased (with \(H_S = 2 \cdot L_S = 2 \cdot D_S\)) until a number \(N_S \geq N_{S,\text{MIN}}\) of in-bed search-nodes can be obtained, 3) the first estimation of the tracer position is outside the bed walls, the length of \(H_S\), \(L_S\) & \(D_S\) is increased (with \(H_S = 2 \cdot L_S = 2 \cdot D_S\)) until a number \(N_S \geq N_{S,\text{MIN}}\) of in-bed search-nodes can be obtained. These different situations are illustrated in Figure 4-21.
The selection of $N_{S,MIN}$ is discussed in Section 4.3.1.3.1.2.

4.3.1.3 Selection of iterative model parameter & results from the selected model

The following section will discuss the selection of the model parameters, specifically, the minimum number of iterations to use $N_{It,MIN}$, the threshold value to stop the iteration $\varepsilon$, and the minimum number of nodes $N_{S,MIN}$ to include in the search volume.

Finally, examples of results obtained with the selected parameters will be presented.

4.3.1.3.1 Selection of model parameters

This section will discuss how the iterative model parameter has been selected. These parameters are:

- The minimum number of iterations per position $N_{It,MIN}$ and the convergence limit $\varepsilon$. They need to be selected, such as using enough re-calculated positions so that the averaging effect reduces signal noise, but keeping a reasonable number of iterations to limit required computing time and avoid introducing artificial error averaging over too many re-calculated positions (Mosorov, 2013).
• The minimum number of nodes to be used for the iterative search of a position \( N_{S,\text{MIN}} \). It needs to be selected to include enough nodes for the search to accurately estimate the actual tracer position. If too few nodes are included, the system might converge towards an incorrect local minimum instead of the correct position. On the other hand, the number of nodes needs to be limited to keep the required computational time within realistic limits.

### 4.3.1.3.1.1 Choice of \( N_{lt,\text{MIN}} \& \varepsilon \)

Figure 4-22 presents the variation of the distance of calculated positions in-between iterations, up to 25 iterations. The number of nodes used for the iterative search was set to an arbitrarily high value of \( N_{S,\text{MIN}} = 1500 \), to ensure a search in a volume large enough (verified in the next section). Figure 4-23 presents the variation in required computing time when \( N_{lt,\text{MIN}} \) increases.

![Variation in calculated positions in-between iterations](image)

**Figure 4-22.** Variation in calculated positions in-between iterations

(Short taper, \( u_{G,\text{Top}} = 0.9 \text{ m}\cdot\text{s}^{-1} \), 5 injection banks, \( L_N = 0 \text{ cm} \), 1 baffle below 2\text{nd} injection, \( F_S = 0.47 \text{ kg}\cdot\text{s}^{-1} \), \( m_{S,\text{Bed}} = 60.3 \text{ kg} \), \( D_{\text{Agg}} = 1.3 \text{ cm} \), \( \rho_{\text{Agg}} = 1100 \text{ kg}\cdot\text{m}^{-3} \), \( \Delta t_{\text{EXP}} = 43 \text{ h} \)
Figure 4-22 shows that the distance between two successive positions calculated initially decreases quickly (first 2-3 iterations) but rapidly reaches a mode of slow decrease, with variations under 5 mm. Therefore, given the size of the tracer (the radioactive core is around [1; 2] mm and the encapsulation around [80; 160] mm) and the inter-node distance used (1 cm), the chosen required convergence value was $\epsilon = 2.5$ mm.

Based on this choice of convergence value $\epsilon$, and using the convergence pattern visible in Figure 4-22, it was chosen to set a minimum required number of iteration $N_{I_{L,MIN}}$ of 5: any calculated position would be the weighted averaged of, at least, the five positions with the higher similarities with the original data. Using Figure 4-23, one can verify that this minimum required number of iterations $N_{I_{L,MIN}}$ has only a marginal impact on the increase of required computational time, which was desirable.

Figure 4-23. Evolution of the required computational time when $N_{I_{L,MIN}}$ increases (Short taper, $u_{G,Top} = 0.9$ m·s$^{-1}$, 5 injection banks, $L_{N} = 0$ cm, 1 baffle below 2$^{nd}$ injection, $F_{S} = 0.47$ kg·s$^{-1}$, $m_{S,Bed} = 60.3$ kg, $D_{Agg} = 1.3$ cm, $\rho_{Agg} = 1100$ kg·m$^{-3}$, $\Delta t_{EXP} = 43$ h)
4.3.1.3.1.2 Choice of $N_{S,\text{MIN}}$

The selection criteria used are the ones described in Section 4.1.1. It should be noted that the iterative methods reconstruct the tracer position using only nodes located within bed walls. Therefore, to reduce the proportion of positions detected outside bed limits, the first criterion is not relevant with this method. The criteria are:

- Limit the creation of zones of artificially low tracer presence within the bed.
- Reduce the average value of the moving standard deviation of the tracer position (reduce sharp unrealistic variations of positions).
- Limit required computing time.

![Graph showing the evolution of relative variation of proportion of artificially low tracer presence inside the bed as a function of $N_{S,\text{MIN}}$](image)

**Figure 4-24.** Evolution of the relative variation of the proportion of artificially low tracer presence inside the bed as a function of the minimum number of nodes used for iterative search $N_{S,\text{MIN}}$ ($N_{R,\text{MIN}} = 5$, $\varepsilon = 2.5 \text{ mm}$). The reference is the data originally recorded. (Short taper, $u_{G,\text{Top}} = 0.9 \text{ m} \cdot \text{s}^{-1}$, 5 injection banks, $L_N = 0 \text{ cm}$, 1 baffle below 2$^{\text{nd}}$ injection, $F_S = 0.47 \text{ kg} \cdot \text{s}^{-1}$, $m_{S,\text{Bed}} = 60.3 \text{ kg}$, $D_{\text{Agg}} = 1.3 \text{ cm}$, $\rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $\Delta t_{\text{EXP}} = 43 \text{ h}$)

Figure 4-24 presents the relative variation of the proportion of artificially low tracer presence inside the bed within the bed volume (as described in Section 4.1.1) when $N_{S,\text{MIN}}$
varies. The reference used is the fraction of voidage in the data originally recorded. Figure 4-25 presents the variation of the average value of moving standard deviation on the cartesian coordinates of the calculated positions when $N_{S,MIN}$ varies. The average was calculated for the whole dataset, and the window used for the moving standard deviation was $\Delta t_{MOV} = 2 \text{ s}$. Figure 4-26 represents the evolution of the required computational time when $N_{S,MIN}$ varies.

![Diagram](image_url)
Using Figure 4-24, the fraction of voidage within the bed volume reaches a local minimum for $N_{S,\text{MIN}} \in [100; 500]$. It then increases before converging towards a minimum for $N_{S,\text{MIN}} > 7500$. Using Figure 4-25, it is clear that the average value of moving standard deviation on the cartesian x- & y-coordinates are minimized for $N_{S,\text{MIN}} = 375$. The average value of moving standard deviation on the cartesian z-coordinates increases with $N_{S,\text{MIN}}$. Nevertheless, for small values of $N_{S,\text{MIN}}$, this increase is marginal. Finally, Figure 4-26 presents the power-law relation between the required computing time and the number $N_{S,\text{MIN}}$: it is clear that a small $N_{S,\text{MIN}}$ is critical to keep the computational time reasonable.
Therefore, a value of $N_{S,\text{MIN}} = 375$ nodes was selected. This allows for excellent results with the selected criteria while limiting the required computed time (110 times longer than the non-iterative process, which remains reasonable given the very fast computational time of CARPT).

### 4.3.1.3.2 Results from selected parameters

This section presents results obtained from the iterative methods. The experimental conditions used were: $u_{G,\text{Top}} = 0.9$ m·s$^{-1}$, 5 injection banks, $L_N = 0$ cm, 1 baffle below 2nd injection, $F_S = 0.47$ kg·s$^{-1}$, $m_{S,\text{Bed}} = 60.3$ kg, $D_{Agg} = 1.3$ cm, $\rho_{Agg} = 1100$ kg·m$^{-3}$, $\Delta t_{\text{EXP}} = 43$ h. The average time interval between two measured radioactive counts was 11.7 ms. The selected parameters were: the minimum number of nodes for iterative search $N_{S,\text{MIN}} = 375$, the minimum number of iterations per position calculation $N_{I,\text{MIN}} = 5$, and the convergence limit $\varepsilon = 2.5$ mm.
Figure 4-27 presents a sample of cartesian coordinates obtained with/without the iterative correction.

\[
\begin{align*}
N_{S,\text{MIN}} &= 375 \\
N_{I,\text{MIN}} &= 5 \\
\varepsilon &= 2.5 \text{ mm}
\end{align*}
\]
\( N_{S,\text{MIN}} = 375 \)
\( N_{N,\text{MIN}} = 5 \)
\( \varepsilon = 2.5 \text{ mm} \)
Figure 4-27. Example of cartesian coordinates obtained before/after iterative correction
(Short taper, $u_{G, \text{Top}} = 0.9 \text{ m} \cdot \text{s}^{-1}$, 5 injection banks, $L_N = 0 \text{ cm}$, 1 baffle below 2$^{\text{nd}}$ injection,
$F_S = 0.47 \text{ kg} \cdot \text{s}^{-1}$, $m_{S, \text{Bed}} = 60.3 \text{ kg}$, $D_{\text{Agg}} = 1.3 \text{ cm}$, $\rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $\Delta t_{\text{EXP}} = 43 \text{ h}$)

a) $x$-coordinates, b) $y$-coordinates, c) $z$-coordinates

Figure 4-27 shows the effect of the selected iterative procedure on the cartesian coordinates calculated. Two main observations can be done: 1) when the trajectory is stable (limited oscillations), the iterative procedure gives very similar results compared to the original CARPT procedure, 2) when the trajectory is not stable, the iterative procedure reduces the
intensity of the oscillations (as expected thanks to the built-in averaging procedure). Nonetheless, significant oscillations are still present after the iterative correction.

Table 4-3 summarizes the evolution of the selection criteria before/after the selected iterative correction. It should be noted that the required computational time is 43 times higher.

**Table 4-3.** Evolution of the selection criteria before/after the selected iterative correction.

<table>
<thead>
<tr>
<th>Parameters used</th>
<th>Original</th>
<th>Iterative</th>
</tr>
</thead>
<tbody>
<tr>
<td>/</td>
<td>/</td>
<td>N\textsubscript{S,MIN} = 375</td>
</tr>
<tr>
<td></td>
<td></td>
<td>N\textsubscript{N,MIN} = 5</td>
</tr>
<tr>
<td></td>
<td></td>
<td>( \varepsilon = 2.5 ) mm</td>
</tr>
<tr>
<td>Proportion of positions detected out of bed</td>
<td>14.18%</td>
<td>0%</td>
</tr>
<tr>
<td>Variation of the proportion of artificially low tracer presence inside the bed</td>
<td>0.25% (1)</td>
<td>0.46% (1.66)</td>
</tr>
<tr>
<td>Relative variation of the average value of moving standard ( \Delta t_{\text{MOV}} = 2 ) s</td>
<td>( \begin{bmatrix} x \ y \ z \end{bmatrix} ) - ( \begin{bmatrix} 1 \ 1 \ 1 \end{bmatrix} )</td>
<td>( \begin{bmatrix} 0.6423 \ 0.6719 \ 0.9979 \end{bmatrix} )</td>
</tr>
<tr>
<td>Computational time required (ms/event)</td>
<td>0.0323</td>
<td>1.3742</td>
</tr>
</tbody>
</table>

Figure 4-28 shows the effect of the iterative procedure on the position distribution within the bed. First, all the positions detected outside the bed are moved back within the bed. This is an explicit feature of the iterative model. Nonetheless, the internal repartition remains similar before/after the correction, confirming that the valid positions are not artificially modified.
Figure 4-28. Repartition of tracer positions detected

(Short taper, \(u_{G,\text{Top}} = 0.9 \text{ m} \cdot \text{s}^{-1}\), 5 injection banks, \(L_N = 0 \text{ cm}\), 1 baffle below 2\textsuperscript{nd} injection,
\(F_S = 0.47 \text{ kg} \cdot \text{s}^{-1}\), \(m_{S,\text{Bed}} = 60.3 \text{ kg}\), \(D_{\text{Agg}} = 1.3 \text{ cm}\), \(\rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3}\), \(\Delta t_{\text{EXP}} = 43 \text{ h}\))

a) Data originally recorded, b) Corrected data (using the iterative algorithm)

4.3.2 Signal smoothing

As discussed in Chapter 3 – Section 3.3, one major limitation of Radioactive Particle Tracking (RPT) is that the counts rate recorded by a scintillation detector for a given tracer position located inside the vessel can vary because of the random nature of the decay of an unstable atom (Leo, 1994; Holbert, 2002) (the counts rate is the number recorded by a detector for a sampling time interval, which was about 12 ms in these experiments). The variance \(\sigma\) of these counts rate can be estimated using a Poisson distribution to describe the radioactive decay (see Equation (3.2)). The coefficient of variation of the counts rate
(σ/CR_M) increases with decreasing source strength (see Figure 3-12-a)) and decreasing sampling time (see Figure 3-12-b)).

The data acquisition system used in this research was set up to acquire data as fast as possible (Sanchez Careaga, 2013) to achieve high temporal resolution. While this is desirable to obtain accurate solids flow patterns, this results in a higher random variation in the measured counts rate. This variation results in an error on the distance source-detector calculated with the calibration curves (see Chapter 3 – Section 3.3.3.2), which propagates to the coordinates determined through CARPT (Chaouki et al., 1997). This resulting noise in each coordinate is white noise, uncorrelated with each other (Degaleesan et al., 2002).

Filtering can be applied either to the radiation counts rate measured by each detector (“smoothing of counts rate”) or to the coordinates calculated (“smoothing of coordinates”). A simplified flowchart of the smoothing correction method is presented in Figure 4-29.

![Figure 4-29. Simplified flowchart of the smoothing correction method](image)

4.3.2.1 Smoothing of counts rate

A simple strategy explored to reduce the random variation in counts rate consists of combining several counts rate recorded at successive sampling times (see Chapter 3 – Section 3.3). Figure 4-30 shows how combining these sampling times can reduce the random error.

The smoothing of the counts rate was investigated to select the best method using the criteria described in Section 4.1.1.

The methods tested were:
• Average counts rate.
• Moving-average of counts rate.
• Average counts rate, with linear interpolation of removed counts rate.

**Figure 4.30.** Relative reduction in random variation (σ) of counts rate achieved by combining successive sampling time

These methods are presented in detail in Appendix F: Overview of the smoothing methods considers for the RPT enhancements: description and analysis.

Several parameters were tested for each type of smoothing, such as the number of points used to perform the average/moving average.

### 4.3.2.2 Smoothing of coordinates

Another approach is to smooth the calculated coordinates. The smoothing of coordinates was investigated to select the best method using the criteria described in Section 4.1.1.

The methods tested were:

• Moving-average of coordinates
• Exponential smoothing (NIST/SEMATECH, 2003)
- Simple (Holt linear method)
- Double (Holt-Winters & Brown methods)

- Wavelets (Degaleesan et al., 2002)
- Kernel smoothing (Hastie et al., 2008)
  - Pseudo-Gaussian
  - Nearest Neighbours
  - Weighted average (Epanechnikov & Tri-Cube methods)

These methods are presented in detail in Appendix F: Overview of the smoothing methods considers for the RPT enhancements: description and analysis.

Several parameters are tested for each type of smoothing, such as the number of points used to perform the moving average, the coefficient(s) used for exponential smoothing, the number of events used for kernel calculations, or the kernel radius to consider.

**4.3.2.3 Selection & combination of smoothing methods**

The best smoothing method combined averaging the counts rate with linear interpolation of removed counts rate, with a double exponential smoothing (Holt-Winters) of resulting coordinates, calculated with the standard CARPT method (and not with the absorption correction method described in Section 4.3.1). The smoothing parameters selected were an average over every two events and exponential smoothing parameters $\alpha$ and $\beta$ of 0.25. A detailed comparison of the performances of the smoothing methods is presented in Appendix G: Selection of the RPT enhancements processing methods: smoothing and iterative correction.

For this analysis, the experimental conditions used were: $u_{G,\text{Top}} = 0.9 \text{ m} \cdot \text{s}^{-1}$, 5 injection banks, $L_N = 0 \text{ cm}$, 1 baffle below 2nd injection bank, $F_S = 0.47 \text{ kg} \cdot \text{s}^{-1}$, $m_S = 60.3 \text{ kg}$, $D_{\text{Agg}} = 1.3 \text{ cm}$, $\rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $\Delta t_{\text{EXP}} = 43 \text{ hours}$. The average time interval between two measured radioactive counts was 11.7 ms.

Figure 4-31 presents a sample of cartesian coordinates obtained with/without the smoothing methods selected.
Figure 4-31. Example of cartesian coordinates obtained before/after iterative correction, smoothing correction. (Short taper, $u_{G,Top} = 0.9 \text{ m} \cdot \text{s}^{-1}$, 5 injection banks, $L_N = 0 \text{ cm}$, 1 baffle below 2nd injection, $F_S = 0.47 \text{ kg} \cdot \text{s}^{-1}$, $m_{S,Bed} = 60.3 \text{ kg}$, $D_{Agg} = 1.3 \text{ cm}$, $\rho_{Agg} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $\Delta t_{EXP} = 43 \text{ h}$)

Smoothing: Average with linear interpolation (every two points) of counts rate & Double Exponential Smoothing of coordinates (Holt-Winters, $\alpha = \beta = 0.25$).

a) x-coordinates, b) y-coordinates, c) z-coordinates
Figure 4-31 shows the effect of the selected smoothing procedure on the cartesian coordinates calculated. Two main observations can be made: 1) when the trajectory is stable (limited oscillations), the smoothing procedure gives very similar results compared to the original CARPT procedure, 2) when the trajectory is not stable, the smoothing procedure drastically reduces the intensity of the oscillations (achieving its objective).

Table 4-4 summarizes the evolution of the selection criteria before/after the selected smoothing correction.

**Table 4-4.** Evolution of the selection criteria before/after the selected smoothing correction.

<table>
<thead>
<tr>
<th>Parameters used</th>
<th>Original</th>
<th>Smoothing</th>
</tr>
</thead>
<tbody>
<tr>
<td>Proportion of positions detected out of bed</td>
<td>14.18%</td>
<td>7.97%</td>
</tr>
<tr>
<td>Variation of the proportion of artificially low tracer presence inside the bed</td>
<td>0.25% (1)</td>
<td>0.37% (1.46)</td>
</tr>
</tbody>
</table>
| Relative variation of the average value of moving standard deviation on coordinate \[
\begin{bmatrix}
  x \\
  y \\
  z
\end{bmatrix}
\](s) - \[\begin{bmatrix}
  x \\
  y \\
  z
\end{bmatrix}\] | \[
\begin{bmatrix}
  1 \\
  1 \\
  1
\end{bmatrix}
\] | \[
\begin{bmatrix}
  0.6879 \\
  0.6930 \\
  0.8832
\end{bmatrix}
\] |
| Computational time required (ms/event)   | 0.0323   | 0.0342                                                                     |

Figure 4-32 shows the effect of the smoothing procedure on the position distribution within the bed. First, the proportion of positions detected outside the bed is decreased. Nonetheless, the internal repartition remains similar before/after the correction, confirming that the valid positions are not artificially modified.
Smoothing methods results shows that the method could be a good complement to the iterative correction. Indeed, smoothing allows for a substantial reduction of coordinates fluctuations but falls short on reducing impossible positions. On the other hand, iterative correction eliminates impossible solutions but showed limited improvement regarding coordinates fluctuations. Both methods, when selected carefully, only result in a limited increase of artificially low tracer presence in the bed. Therefore, combining both methods is a promising solution to improve the RPT signal significantly.

**Figure 4-32.** Repartition of tracer positions detected

\(u_G = [0.3; 0.9] \text{ m} \cdot \text{s}^{-1}, \text{ 5 injection banks (} L_N = 0 \text{ cm), 1 baffle below second injection bank, } F_S = 0.47 \text{ kg} \cdot \text{s}^{-1}, m_{S,Bed} = 60.3 \text{ kg, } \Delta t_{\text{EXP}} = 43 \text{ h}\)

*Smoothing: Average with linear interpolation (every two points) of counts rate & Double Exponential Smoothing of coordinates (Holt-Winters, } \alpha = \beta = 0.25).*

a) Original RPT vs. b) RPT with smoothing selected method
4.4 Model results analysis

This section presents results obtained when processing data with both the absorption correction methods and the combined smoothing methods.

The selected combined smoothing methods are an averaged counts rate (with linear interpolation of removed counts rate) and double exponential smoothing (Holt-Winters) of the coordinates. The parameters used for the smoothing method are an average over every two events and exponential smoothing parameters $\alpha$ and $\beta$ of 0.25. The parameters used for the selected iterative method are the minimum number of nodes for iterative search $N_{S,\text{MIN}} = 375$, the minimum number of iterations per position calculation $N_{I,\text{MIN}} = 5$, and the convergence limit $\varepsilon = 2.5$ mm.

Figure 4-33. Simplified flowchart of the combined absorption correction method and smoothing correction method

For this analysis, the experimental conditions used were: $u_{G,\text{Top}} = 0.9$ m·s$^{-1}$, 5 injection banks, $L_N = 0$ cm, 1 baffle below 2nd injection bank, $F_S = 0.47$ kg·s$^{-1}$, $m_{S,\text{Bed}} = 60.3$ kg,
\[ D_{\text{Agg}} = 1.3 \text{ cm}, \quad \rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3}, \quad \Delta t_{\text{EXP}} = 43 \text{ hours}. \] The average time interval between two measured radioactive counts was 11.7 ms.

A simplified flowchart of the combined absorption correction method and smoothing correction method is presented in Figure 4-33.

4.4.1 Consistency of RPT results
b)

![Graph showing raw data and smoothed data over time](image)

- **Raw data**
- **Smoothed data**
- **Corrected data**

**Y (m)**

**t (s)**
Figure 4-34. Example of cartesian coordinates obtained before/after iterative correction, smoothing correction, and combined iterative & smoothing correction.

(Short taper, $u_{G,\text{Top}} = 0.9 \text{ m}\cdot\text{s}^{-1}$, 5 injection banks, $L_N = 0$ cm, 1 baffle below 2nd injection, $F_S = 0.47 \text{ kg}\cdot\text{s}^{-1}$, $m_{S,\text{Bed}} = 60.3 \text{ kg}$, $D_{Agg} = 1.3 \text{ cm}$, $\rho_{Agg} = 1100 \text{ kg}\cdot\text{m}^{-3}$, $\Delta t_{\text{EXP}} = 43 \text{ h}$)

*ITERATIVE*: $N_{S,\text{MIN}} = 375 / N_{r,\text{MIN}} = 5 / \varepsilon = 2.5 \text{ mm}$, *SMOOTHING*: Pre-processing: Average / Every 2 events (with interpolation) // Post-processing: Exponential Soothing / Double (Holt-Winters) / $\alpha = 0.25 / \beta = 0.25$

a) x-coordinates, b) y-coordinates, c) z-coordinates
Figure 4-34 presents a sample of cartesian coordinates obtained from the original dataset, from the data processed with the iterative correction, from the data processed with the smoothing correction, from the data processed with both the iterative and smoothing correction.

Figure 4-34 shows the effect of the combined iterative and smoothing procedure on the cartesian coordinates calculated. It can be observed that the coordinates fluctuations are minimized by combining the iterative procedure (thanks to its built-in averaging procedure) and the smoothing procedure. It should be noted that the z-component of coordinates presents the highest residual fluctuations. It is likely due to the higher position variation along the z-axis than the x-axis and y-axis.

Table 4-5 summarizes the evolution of the selection criteria when considering the original dataset, the data processed with the iterative correction, the data processed with the smoothing correction, the data processed with both the iterative and smoothing correction.

Figure 4-35 presents the z-r projection of the positions obtained for the whole run, from the original dataset, from the data processed with the iterative correction, from the data processed with the smoothing correction, from the data processed with both the iterative and smoothing correction. First, it should be noted that the internal repartition remains similar before/after the correction(s), with high concentration zones at the bottom of the taper and near the 4th injection bank. First, all the positions detected outside the bed are moved back within the bed. This is an explicit feature of the iterative model. It is also interesting to note that very few variations are visible in-between the case with iterative correction and the case with iterative and smoothing correction.

Figure 4-36 presents the z-r projection of the local z-velocity of the tracer obtained for the whole run, from the original dataset, from the data processed with the iterative correction, from the data processed with the smoothing correction, from the data processed with both the iterative and smoothing correction. Contrary to what was observed in the z-r projection of the tracer positions (Figure 4-35), both the iterative and the smoothing methods significantly alter the results.
Table 4-5. Evolution of the selection criteria when the iterative correction and/or the smoothing correction is/are applied or not.

<table>
<thead>
<tr>
<th>Parameters used</th>
<th>Original</th>
<th>Iterative</th>
<th>Smoothing</th>
<th>Iterative + smoothing</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>/</td>
<td>Iterative</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>N&lt;sub&gt;S,MIN&lt;/sub&gt; = 375</td>
<td>N&lt;sub&gt;It,MIN&lt;/sub&gt; = 5</td>
<td>N&lt;sub&gt;S,MIN&lt;/sub&gt; = 375</td>
</tr>
<tr>
<td></td>
<td></td>
<td>ε = 2.5 mm</td>
<td></td>
<td>ε = 2.5 mm</td>
</tr>
<tr>
<td>Proportion of positions detected out of bed</td>
<td>14.18%</td>
<td>0%</td>
<td>7.97%</td>
<td>0%</td>
</tr>
<tr>
<td>Variation of the proportion of artificially low tracer presence inside the bed</td>
<td>0.25% (1)</td>
<td>0.46% (1.66)</td>
<td>0.37% (1.46)</td>
<td>0.39% (1.54)</td>
</tr>
<tr>
<td>Relative variation of the average value of moving standard deviation on coordinate</td>
<td>0.6423</td>
<td>0.6719</td>
<td>0.6979</td>
<td>0.5509</td>
</tr>
<tr>
<td>Δt&lt;sub&gt;Mov&lt;/sub&gt; = 2 s</td>
<td>1.0979</td>
<td>0.8832</td>
<td>0.9449</td>
<td></td>
</tr>
<tr>
<td>Computational time required (ms/event)</td>
<td>0.0323</td>
<td>1.3742</td>
<td>0.0342</td>
<td>1.3965</td>
</tr>
</tbody>
</table>
Figure 4-35. Evolution of tracer positions repartition with different processing methods. (Short taper, \( u_{G,\text{Top}} = 0.9 \text{ m} \cdot \text{s}^{-1} \), 5 injection banks, \( L_N = 0 \text{ cm} \), 1 baffle below 2\textsuperscript{nd} injection,
The combination of both the iterative and smoothing methods leads to results significantly different from the use of any single method and gives the most realistic results:

1. At the bottom of the vessel ($z \in [0.1; 0.3]$ m), there are high positive (upward) $z$-velocities in the periphery and high negative (downward) $z$-velocities near the core. This pattern matches the geometry of the sparger (See Chapter 3 – Section 3.1.2.1) coupled with the downward solids recirculation in the experimental unit.

2. At the bottom of the taper level ($z = 0.32$ m), the high positive $z$-velocities shift towards the core. This corresponds to the solids flow pattern measured in tapered fluidized beds (Toyohara and Kawamura, 1990; Murthy et al., 2009).

3. Within the injection zone (above the first injection bank, $z = 0.56$ m), the lateral mixing induced by the lateral injection (nozzles tip are located at the wall for the results presented here) leads to a more homogenous $z$-velocity profile. Local maximum upward velocity is observed at mid-radius. Slightly negative velocities are observed near the core. This phenomenon is discussed in more detail in Section 4.4.2.

4. After the last injection bank ($5^{th}$ bank, $z = 1.16$ m), there is no more lateral mixing, and the pattern is reversed again, reversed with high negative $z$-velocities observed in the periphery. In contrast, high positive $z$-velocities are observed near the core.
<table>
<thead>
<tr>
<th>NO smoothing</th>
<th>Smoothing</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>z-velocity (m·s⁻¹)</strong></td>
<td><strong>z-velocity (m·s⁻¹)</strong></td>
</tr>
<tr>
<td><img src="image1.png" alt="Diagram without Iterative Correction" /></td>
<td><img src="image2.png" alt="Diagram with Iterative Correction" /></td>
</tr>
</tbody>
</table>

<table>
<thead>
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<th>Iterative Correction</th>
</tr>
</thead>
<tbody>
<tr>
<td><img src="image3.png" alt="Diagram without Iterative Correction" /></td>
<td><img src="image4.png" alt="Diagram with Iterative Correction" /></td>
</tr>
</tbody>
</table>
**Figure 4-36.** Evolution of tracer median z-velocity (high density) with different processing methods

(Short taper, \( u_{G,\text{Top}} = 0.9 \text{ m}\cdot\text{s}^{-1} \), 5 injection banks, \( L_N = 0 \text{ cm} \), 1 baffle below 2\textsuperscript{nd} injection, \( \dot{F}_S = 0.47 \text{ kg}\cdot\text{s}^{-1} \), \( m_{S,\text{Bed}} = 60.3 \text{ kg} \), \( D_{\text{Agg}} = 1.3 \text{ cm} \), \( \rho_{\text{Agg}} = 1100 \text{ kg}\cdot\text{m}^{-3} \), \( \Delta t_{\text{EXP}} = 43 \text{ h} \))

**ITERATIVE:** \( N_{S,\text{MIN}} = 375 / N_{R,\text{MIN}} = 5 / \varepsilon = 2.5 \text{ mm} \), **SMOOTHING:** *Pre-processing:*

*Average / Every 2 events (with interpolation) // Post-Processing: Exponential Soothing / Double (Holt-Winters) / \( \alpha = 0.25 / \beta = 0.25 \)*

Figure 4-37 presents the z-r projection of the local z-acceleration obtained for the whole run, from the original dataset, from the data processed with the iterative correction, from the data processed with the smoothing correction, from the data processed with both the iterative and smoothing correction. Similar to what was observed in the z-r projection of the tracer z-velocity (Figure 4-36), both the iterative and the smoothing methods significantly alter the results. The effect of the pseudo-ring sparger and the taper are again clearly visible. In addition, the effect of the baffle is discernible, with a zone of high solids acceleration located above its tip. This zone of high z-acceleration then converges towards the unit center.
<table>
<thead>
<tr>
<th>NO smoothing</th>
<th>Smoothing</th>
</tr>
</thead>
<tbody>
<tr>
<td><img src="image1.png" alt="Graph" /></td>
<td><img src="image2.png" alt="Graph" /></td>
</tr>
<tr>
<td><img src="image3.png" alt="Graph" /></td>
<td><img src="image4.png" alt="Graph" /></td>
</tr>
</tbody>
</table>

**Figure 4.37.** Evolution of tracer median $z$-acceleration (high density) with different processing methods (Short taper, $u_{G,Top} = 0.9 \text{ m}\cdot\text{s}^{-1}$, 5 injection banks, $L_N = 0 \text{ cm}$, 1 baffle)
below 2\textsuperscript{nd} injection, \( F_S = 0.47 \text{ kg}\cdot\text{s}^{-1} \), \( m_{S,\text{Bed}} = 60.3 \text{ kg}, D_{\text{Agg}} = 1.3 \text{ cm}, \rho_{\text{Agg}} = 1100 \text{ kg}\cdot\text{m}^{-3}, \Delta t_{\text{EXP}} = 43 \text{ h} \)

\textit{ITERATIVE:} \( N_{S,\text{MIN}} = 375 / N_{R,\text{MIN}} = 5 / \varepsilon = 2.5 \text{ mm}, \textit{SMOOTHING: Pre-processing: Average / Every 2 events (with interpolation) // Post-Processing: Exponential Soothing / Double (Holt-Winters) / } \alpha = 0.25 / \beta = 0.25 \)

### 4.4.2 Comparison of RPT results with other results from the literature

This section presents a comparison of some results experimentally obtained in this research, with and without corrections applied, with results from the literature.

The vertical tracer motion is mainly governed by its interaction with bubbles and their wakes, especially upward motion: the tracer goes up only when it is picked up by bubbles and their wakes. Therefore, the bubble distribution within the bed is important to understand vertical tracer motion.

Experiments performed in the literature with a comparable column and slightly different conditions (Song et al., 2004) suggested a core-annulus structure with bubbles flowing primarily at the center. However, a CFD simulation of these experiments (Li et al., 2012b; Li et al., 2012a) predicted a different profile. The bed voidage peaks halfway between the wall and the bed center, at the height just above the middle banks. It suggests that most bubbles do not concentrate at the center. This result could explain why the positive median tracer vertical velocity is not always located at the center.
Figure 4-38. Tracer median z-velocity vs. bed voidage from the literature
(Short taper, \( u_{G, \text{Top}} = 0.9 \text{ m} \cdot \text{s}^{-1} \), 5 injection banks, \( L_N = 2 \text{ cm} \), No baffle, \( F_S = 0.45 \text{ kg} \cdot \text{s}^{-1} \), \( FL_S = 14.3 \text{ kg} \cdot \text{s}^{-1} \cdot \text{m}^{-2} \), \( m_{S, \text{Bed}} = 59.9 \text{ kg} \), \( D_{Agg} = 1.3 \text{ cm} \), \( \rho_{Agg} = 1100 \text{ kg} \cdot \text{m}^{-3} \), \( \Delta t_{\text{EXP}} = 36 \text{ h} \))

ITERATIVE: \( N_{S,MIN} = 375 / N_{F,MIN} = 5 / \varepsilon = 2.5 \text{ mm} \), SMOOTHING: Pre-processing:
Average / Every 2 events (with interpolation) // Post-Processing: Exponential Soothing /

\[ Double \ (Holt-Winters) / \alpha = 0.25 / \beta = 0.25 \]

a) Tracer z-velocity (high density) with no correction or smoothing,

b) Tracer z-velocity (high density) with correction & smoothing,

c) Voidage distribution in reactor section (\( u_G \in [0.25; 0.74] \text{ m} \cdot \text{s}^{-1} \), \( FL_S = 18.6 \text{ kg} \cdot \text{s}^{-1} \cdot \text{m}^{-2} \))

(Song et al., 2004)

d) Contours of mean solid volume fraction of a CFD gas-solid only simulation

(FCC particles, \( u_G \in [0.25; 0.75] \text{ m} \cdot \text{s}^{-1} \), \( FL_S = 18.6 \text{ kg} \cdot \text{s}^{-1} \cdot \text{m}^{-2} \)) (Li et al., 2012b)

It should also be noted that the height at which a negative velocity is observed near the core corresponds to the end of the bubbling-to-turbulent regime transition observed in the experimental bed. Finally, these core zones in the upper part of the bed associated with negative (downward) velocities are also associated with positive (upward) accelerations. This is because some bubbles are slowing down the tracer motion, but not enough to pick it up enough times to get average upward motion.

4.4.3 Comparison of RPT results with other measurement methods

This section presents critical results obtained with the RPT and compares them with other independent methods, such as introduced in Section 4.1.2. These results include the RPT corrections selected in this chapter.

4.4.3.1 Tracer time distribution

The radioactive tracer Residence Time Distribution (RTD) obtained with RPT, with the corrections selected, are compared with results obtained from dyed coke experiments. The details of the comparison procedure are provided in Section 4.1.2.1 and Appendix D: Solids RTD validation experiments with dyed coke.
As in Figure 4-9, Figure 4-39 presents the convoluted Top-to-Stripper RPT time distributions for two range densities of agglomerates of tracer: “heavy (1100-1200 kg·m\(^{-3}\)) and “light” (850-950 kg·m\(^{-3}\)), with and without the corrections. As explained in Section 4.1.2.1, the RPT results were convoluted to simulate the results of the dye tracer experiments. The profiles obtained without and with the RPT corrections are presented for both tracers. The details of the convolution procedure are presented in Appendix D: Solids RTD validation experiments with dyed coke.

![Figure 4-39](image)

**Figure 4-39.** Top-to-stripper RPT convoluted time distribution (with and without correction)

(Short taper, \(u_{G,Top} = 0.9 \text{ m} \cdot \text{s}^{-1}\), 5 injection banks, \(L_N = 0 \text{ cm}\), No baffle, \(F_S = 0.50 \text{ kg} \cdot \text{s}^{-1}\), \(m_{S,\text{Bed}} = 60.6 \text{ kg}\), \(D_{\text{Agg}} = 1.3 \text{ cm}\), \(\rho_{\text{Agg}} = [850; 1200] \text{ kg} \cdot \text{m}^{-3}\), \(\Delta t_{\text{EXP}} = [36; 45] \text{ h}\))

The results in Figure 4-39 shows that the moderate difference in Top-to-Stripper RPT time distributions between the “heavy” and “light” tracer is maintained after the RPT corrections: the heavy tracer travels faster through the bed, indicating that wet-agglomerates have a shorter residence time than the overall solids population. The RPT
corrections effect modifies the Top-to-Stripper RPT time distributions in two major ways: the overall corrected profiles indicate shorter residence time, but the shortest time were artificial artifacts removed by the corrections (especially for the large tracers).

As in Figure 4-10, Figure 4-40 presents a comparison between these convoluted Top-to-Stripper RPT time distributions and the measured absorbance of sampled coke (integrated over a selected wavelength range, see Appendix D: Solids RTD validation experiments with dyed coke) over time.

**Figure 4-40.** Comparison of dyed coke residence time profile and convoluted top-to-stripper RPT time distribution (with and without correction)

(Short taper, \(u_{G,\text{Top}} = 0.9 \text{ m⋅s}^{-1}\), 5 injection banks, \(L_N = 0 \text{ cm}\), \(F_S = 0.50 \text{ kg⋅s}^{-1}\), \(D_{\text{Agg}} = 1.3 \text{ cm}\), \(\rho_{\text{Agg}} = [850; 1200] \text{ kg⋅m}^{-3}\))

Figure 4-40 shows that the RPT measurements give realistic residence time distribution measurements, especially for coke particles in the bed emulsion. It also confirms that wet-agglomerates (and tracers used to simulate them) have a moderately shorter relative
residence time distribution than bed emulsion. The corrections make the distinction between “heavy” and “light” tracer clearer.
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Chapter 5

5 Impact of injection redistribution and nozzle penetration on liquid losses

The model developed in Chapter 2 to predict the liquid carryover to the stripper section of a Fluid Coker is used in conjunction with experimental data obtained in a pilot-scale cold gas-solid recirculated fluidized bed. The RPT experimental data are processed with the method introduced in Chapter 4. In this chapter, it is used to study the impact of modifying the lateral injection pattern on the liquid losses to the bottom of a commercial Fluid Coker. Two types of modifications are investigated: first, the redistribution of the liquid feedstock between the various nozzle banks, and second, penetration of the spray nozzles into the fluidized bed.

The analysis is carried out using a “representative” agglomerate, with initial properties chosen based on the analysis conducted in Chapter 2. Then, the results are verified by investigating different initial agglomerate properties. In the same way, the main model parameters whose values are not known with precision are analyzed to ascertain the pertinence of the results.

Finally, a short section discusses the other sources of valuable liquid losses, from liquid trapped in the emulsion phase or gas trapped in the emulsion or bubble phase, and how they could be affected by the proposed changes. The first results of preliminary investigative experiments using Varsol™ are presented.

5.1 Lateral injection redistribution

In this section, the liquid losses predicted with the model are used to compare various redistribution patterns and select the redistribution offering the best improvement in liquid losses reduction. In addition, the model mechanisms are analyzed to understand the controlling parameters behind predicted improvements.
5.1.1 Lateral injection distribution in commercial Fluid Cokers

In a commercial Fluid Coker, lateral steam-bitumen spray banks are used to inject the feedstock into the bed of hot coke. Each of these banks comprises a ring of nozzles located around the bed circumference (see Chapter 1). Previous work demonstrated that modifying the lateral injection profile, i.e. injecting more bitumen in the higher parts of the bed than in the lower parts, could help reduce liquid losses to the bottom of the Fluid Coker (Wormsbecker, Wiens, et al., 2021).

While in some most commercial Fluid Cokers, each nozzle bank has the same number of nozzles, as in the experimental unit used in this research, in other cokers, the top lateral injection banks have more feed nozzles than the bottom banks (due to the increase of cross-sectional area from the tapered shape). Therefore, upper lateral injection banks tend to inject more steam-bitumen mixture than lower banks. However, when possible, the feed is typically introduced in a balanced approach with the same number of feed nozzles in operation on each ring. Therefore, the experimental unit used in this research was designed to have the same number of nozzles in each nozzle bank.

5.1.2 Lateral redistribution in the experimental unit

In the pilot-scale cold gas-solid recirculated unit, the lateral feedstock injections are simulated with banks of pure gas injectors. For a case using all the lateral injection banks, the gas feed (simulating the steam-bitumen injection and the resulting hydrocarbons vaporization) is typically introduced in a balanced approach to each bank (with the same gas flowrate for each bank).

In the experimental unit, a lateral redistribution is achieved by shutting down (or modifying) the gas flowrate injected in one (or more) lateral injection bank(s) and redistributing such as to keep a constant gas mass flowrate at the unit exhaust.

Unless noted otherwise, the redistribution is performed evenly between the remaining active lateral injection banks, proving the same extra gas flowrate for each bank still active.
5.1.3 Model results for a “representative” agglomerate, typical model parameters and standard experimental conditions

In a commercial Fluid Coker, the agglomerate population includes agglomerates with a wide range of size, density, and wetness (see Table 2-1). On the other hand, as explained in Chapter 2, the model developed in this work only considers agglomerates with a single set of initial properties. For the sake of clarity, as a first step, the model results are presented for an average “representative” agglomerate defined by the following initial properties: \( \rho_{\text{Agg},0} = 1100 \text{ kg m}^{-3} \), \( D_{\text{Agg},0} = 1 \text{ cm} \), \( (L/S)_0 = 0.31 \). Then, in Section 5.3, the model results will be verified for other values of initial agglomerates properties.

![Figure 5-1. Simplified schematics of the experimental setup](image)

Similarly, the model uses several parameters to predict the liquid losses at the bottom of a Fluid Coker. Some of these parameters are only known with moderate accuracy, or they
can vary within a specific range of values given in the literature. The most critical of these parameters are: the coefficient of re-wetting $C_{Rewet}$, the empirical coefficient $M$ and the bubble diameter $d_b$. As a first step, these parameters will be set as follows: $C_{Rewet} = 1$, $M = 7$, $d_b = 4.1 \text{ cm}$. Then, in Section 5.4, the model results will be verified for other values of these model parameters.

The experimental conditions used for this section of the results were selected based on the principles described in Chapter 3 – Section 3.1. A simplified schematic of the experimental system is presented in Figure 5-1. Additional experimental parameters include a recirculation flowrate $F_S \in [0.42; 0.48] \text{ kg} \cdot \text{s}^{-1}$. The target value was set to 0.45 kg·s$^{-1}$. In addition, the bed mass was varied: $m_S \in [58; 61.5] \text{ kg}$. The target value was 60 kg. The variability of these two parameters is accounted for in all results thanks to the experimental solid residence time $\tau_{EXP}$ (as defined in Appendix I: Spatial and Temporal scaling of the model). In addition, all the lateral injection nozzles were installed, such as having a penetration $L_N = 0 \text{ cm} \ (0\% \ of \ radius)$. The impact of different nozzle penetrations (up to 24% of the radius) is presented in Section 5.2. Finally, the experimental tracer-agglomerate used has the following properties: $D_{Tracer} \in [0.6; 1.4] \text{ cm}$, $\rho_{Tracer} \in [900; 1150] \text{ kg} \cdot \text{m}^{-3}$ (see Figure 3-14). The first results presented use a tracer-agglomerate with the following properties: $D_{Tracer} = 0.9 \text{ cm}$, $\rho_{Tracer} = 1100 \text{ kg} \cdot \text{m}^{-3}$. In Section 5.3, the model results will be verified with measurements with other values of tracer-agglomerates properties.

### 5.1.3.1 Liquid losses predicted by the model

Figure 5-2 presents the liquid losses predicted by the model in a commercial Fluid Coker. These liquid losses correspond to the fraction (on a mass basis) of feedstock liquid injected into the reactor that reached the top of the stripper zone by being carried within wet-agglomerates. Thus, these liquid losses do not include any feedstock liquid carried down as liquid in the emulsion or as vapour in the emulsion.

The results in Figure 5-2 show that, for the “representative” agglomerate considered, in the case where the 5 lateral injection banks of the experimental unit are considered (later referred to as the “base case”), liquid losses are around 0.1 % of the feedstock injected.
However, it should be noted that this value can vary for other agglomerates or other model parameters, as presented in Section 5.3 & Section 5.4.

![Graph showing fraction of injected liquid reaching the bottom of the Fluid Coker](image)

**Figure 5-2.** Fraction of injected liquid reaching the bottom of the Fluid Coker

\[ (\rho_{\text{Agg},0} = 1100 \text{ kg}\cdot\text{m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31) \]

The results also show that a significant reduction in liquid losses is only achieved when the lowest bank (Bank A) is turned off with the injection profile redistributed. A factor of reduction of 0.21 is achieved (for the “representative” agglomerate, with the model parameters used). When other banks are turned off with the injection profile redistributed, no significant variations are observed.

A statistical analysis is conducted to verify the significance of the results presented, using a test of hypotheses relative to the difference of means (Spiegel and Stephens, 2017: 277). The procedure involves a test against a two-tailed Student distribution with a 99% confidence interval and is described in Appendix N: Statistical analysis of the model results. The results are presented in Table 5-1.

The analysis confirms that only turning off and redistributing the lowest bank (Bank A) has a statistically significant impact. The other redistributions do not lead to significant modification of the predicted liquid losses.
Table 5-1. Statistical analysis (Differences of means) – Effect of the redistribution for the “representative” agglomerate

<table>
<thead>
<tr>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td>5 Banks (A off)</td>
<td>4 Banks (B off)</td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
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</tr>
<tr>
<td><strong>Average</strong></td>
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<tr>
<td><strong>Standard Deviation</strong></td>
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</tr>
<tr>
<td>$\sigma$</td>
<td>2.02 $10^{-4}$</td>
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<tr>
<td>$t_{\text{STUDENT}}$ (two-tailed, 99%)</td>
<td>3.25</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>TRUE</td>
</tr>
</tbody>
</table>

5.1.3.2 Impact of redistribution on mechanisms included in the model

This section investigates how lateral injection redistribution impacts the various mechanisms bundled in the model (as presented in Chapter 2). It should be noted that these mechanisms are interconnected: for instance, a longer drying time will promote more breakage. The objective of presenting these mechanisms one after the other is not to imply that they are independent, but to isolate controlling mechanisms.

5.1.3.2.1 Wetting at formation

In addition to considering only a population of agglomerates with only one single set of initial properties, another significant limitation of the model is related to agglomerate formation. The model can only directly calculate the mass fraction of liquid transferred by wet-agglomerates from their formation to their travel to the bottom of the Fluid Coker (if they do not dry or break up). The model calculates how much of the liquid injected is trapped in the agglomerate formed using an empirical correlation Chapter 2 – Section 2.3.2.2) varying with two parameters: the initial agglomerate diameter $D_{\text{Agg,0}}$ and the cross-sectional superficial velocity below the formation zone $u_{G,Below}$ (Li, 2016). Figure 5-3 shows both the fraction of liquid in the initial agglomerates reaching the bottom
of the Fluid Coker and the fraction of injected liquid reaching the bottom of the Fluid Coker.

**Figure 5-3.** Fraction of injected liquid reaching the bottom of the Fluid Coker

Respectively relative from liquid injected or liquid initially trapped in agglomerates

\[(\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31)\]

In the model, the contribution of each bank towards the liquid losses for the overall bed is equal to the ratio of the mass flowrate of feedstock injected in the bank divided by the total mass flowrates injected in all active banks. With the gas-solid experimental unit used in this research, this feedstock mass flowrate was replaced by the gas mass flowrate injected in each bank.

5.1.3.2.2 Production-to-stripper time & drying

Once an agglomerate is formed and leaves the formation zone (at the tip of a lateral injection jet), it starts to dry. As presented in Chapter 2 – Section 2.4, the drying model used in the model is the Shrinking Core Model. Therefore, for two identical agglomerates, the drying level is only controlled by the production-to-stripper time (as defined in Chapter 2 – Section 2.4.1.1).
Figure 5-4 compares the cumulative distribution of the production-to-stripper time of the base case with the cumulative distribution of the production-to-stripper time of the configuration where one bank was turned off with the injection pattern redistributed.

Regarding the production-to-stripper time distribution figures, it should be noted that they are made dimensionless using the average solids residence time in the bed (as explained in Chapter 2 – Section 2.2 & Appendix I: Spatial and Temporal scaling of the model). These residence times are, by definition, much longer than the production-to-stripper times since the fluidized bed is not a perfect CSTR for the agglomerates. This difference explains why almost all the agglomerates go through the bed in less than $t/\tau_{\text{EXP}} = 1$.

\textbf{a)} 5 banks vs. 4 banks (A off)
b) 5 banks vs. 4 banks (B off)

![Graph for 5 banks vs. 4 banks (B off)]

Cumulative distribution of production-to-stripper time

$\frac{t}{\tau_{EXP}}$

5 banks

4 banks (B off)

c) 5 banks vs. 4 banks (C off)

![Graph for 5 banks vs. 4 banks (C off)]

Cumulative distribution of production-to-stripper time

$\frac{t}{\tau_{EXP}}$

5 banks

4 banks (C off)
**d) 5 banks vs. 4 banks (D off)**

![Cumulative distribution graph](image)

**e) 5 banks vs. 4 banks (E off)**

![Cumulative distribution graph](image)

**Figure 5-4.** Cumulative distribution of the production-to-stripper time of tracer-agglomerates for the whole bed ($D_{Tracer} = 0.9 \text{ cm}$, $\rho_{Tracer} = 1100 \text{ kg} \cdot \text{m}^{-3}$, with replicates)

- a) 5 banks vs. 4 banks (A off), b) 5 banks vs. 4 banks (B off),
- c) 5 banks vs. 4 banks (C off), d) 5 banks vs. 4 banks (D off),
- e) 5 banks vs. 4 banks (E off)
Figure 5-4 shows that the only modification significantly lengthening production-to-stripper time is turning off and redistributing the lowest bank (Bank A). The dimensionless $t/\tau_{\text{EXP}}$ required for the 50% of the agglomerates to travel from the production zone to the stripper zone increases from 0.134 to 0.155 (Factor of 1.16).

It should be noted that when considering only the fastest travelling agglomerates, the variation is much more important: the dimensionless $t/\tau_{\text{EXP}}$ required for the 25% fastest agglomerates to travel from the production zone to the stripper zone increases from 0.048 to 0.080 (Factor of 1.68).

![Diagram of cumulative distribution](image)

**Figure 5-5.** Cumulative distribution of the production-to-stripper time of tracer-agglomerates with the contribution of each bank

(5 Banks distribution, $D_{\text{Tracer}} = 0.9$ cm, $\rho_{\text{Tracer}} = 1100$ kg·m$^{-3}$)

Figure 5-4 also shows that turning off and redistributing another bank does not significantly increase the production-to-stripper time compared to the base case.

These longer production-to-stripper times detected when Bank A is turned off and redistributed can be attributed to two main parameters: 1) a longer minimum distance between the production zones and the stripper zone, 2) a modified hydrodynamics slowing down agglomerates travelling down.
Figure 5-6. Cumulative distribution of the production-to-stripper time of tracer-agglomerates, contribution of one bank, with 5 Banks distribution vs. 4 Banks distribution (A off) ($D_{\text{Tracer}} = 0.9$ cm, $\rho_{\text{Tracer}} = 1100$ kg·m$^{-3}$)

a) Bank B, b) Bank C, c) Bank D, d) Bank E

The first parameter is straightforward: the longer the distance between the production zone and the stripper, the more time it requires to travel from one to the other. This effect is visible in Figure 5-5, presenting the cumulative distribution of the production-to-stripper time of tracer-agglomerates with the contribution of each bank. The further the bank is
from the stripper, the longer the production-to-stripper is. The lowest bank (Bank A) is especially significant when considering the fast-traveling agglomerates (fastest 10% of the overall distribution).

**Figure 5-7.** Vertical median tracer $z$-velocity as a function of the injection profile

a) 5 injection banks, b) 4 injection banks (Bank A off)

($D_{\text{Tracer}} = 0.9$ cm, $\rho_{\text{Tracer}} = 1100$ kg·m$^{-3}$, $L_{IN} = 0$ cm)

The second parameter is more complex to verify: does the modified bed hydrodynamics created when one bank is turned off and redistributed slow down or accelerate the travel of agglomerates to the stripper zone? Figure 5-6 indicates the former. For the banks B, C, D & E, the 4 banks common between the 5 banks distribution and the 4 banks distribution (A off), the production-to-stripper time indicates slower travel towards the stripper (except for the 2% fastest agglomerates generated at bank B). This result seems to indicate more favourable hydrodynamics once Bank A is off and redistributed. Figure 5-7
gives some insight into the reason behind this more favourable agglomerate hydrodynamics.

**Figure 5-8.** Average liquid content carried by agglomerates to the bottom of the Fluid Coker

\[
\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, \quad D_{\text{Agg},0} = 1.0 \text{ cm}, \quad (L/S)_0 = 0.31
\]

By shutting down and redistributing Bank A, the agglomerate hydrodynamics in the lower part of the bed (between 0.3 m and 0.7 m high) are modified, and a stronger positive vertical velocity appears. With the 5 banks distribution, the median upward vertical tracer velocity zone extends only up to 0.55 m (the location of Bank A) and with moderate positive velocities in the range [0; 0.2] m·s⁻¹ when above 0.45 m high. On the other hand, with A off, the median upward vertical tracer velocity zone extends up to 0.70 m (the location of Bank B, first active bank) and with stronger positive velocities in the range [0.2; 0.6] m·s⁻¹. This modification is because bubbles and their wakes carry up agglomerates. However, the presence of lateral injections disrupts the vertical motions of these bubbles, limiting how efficiently they can carry up agglomerates.

Figure 5-8 presents the average liquid content carried by agglomerates to the stripper zone. This parameter decreases with: 1) longer drying (longer production-to-stripper time), 2)
more agglomerate destruction, 3) less re-wetting. These effects can be complementary or opposed. The results presented in Figure 5-8 follow the same trend as the results given by the study of the production-to-stripper time (Figure 5-5): only turning off and redistributing Bank A significantly reduces the liquid losses. It suggests that the three mechanisms aforementioned are complementary, or at least that the drying (controlled by the production-to-stripper time) is not neglectable.

5.1.3.2.3 Bed shear & breakage

In addition to drying, the other mechanism limiting the fraction of wet-agglomerates reaching the bottom of the Fluid Coker is the breakage due to bed shear (see Chapter 2 – Section 2.5). This mechanism is connected to three parameters: agglomerate saturation, local bed shear, and agglomerate trajectories in the bed. It should be noted that the agglomerate saturation is a direct function of the agglomerate wetness. Therefore, drying impacts the breakage mechanism: dryer agglomerates break more easily. This relation is important as it highlights how interconnected the mechanisms of the model are.

The local bed hydrodynamics controls the local bed shear. Therefore, it varies with the injection profile, as shown in Figure 5-9.

As presented in Chapter 2 – Section 2.5.1.2, bed shear increases with the cross-sectional superficial gas velocity. Therefore, in the experimental unit, the bed shear increases with the bed height, with clear separation at the location of injection banks. The reality of the bed shear near the lateral injection jets is likely more complex than what is portrayed here, suggesting a possible path for future model refinement.

In addition, bed shear is a function of the local bed voidage. As described in Chapter 2 – Section 2.5.1.2.2, this local voidage was obtained through a combination of pressure-based and radiation-based measurements then fitted using an unimodal core annulus assumption (see Appendix K: Measurement of the bed voidage profile in the experimental unit). This assumption implied that, for a given height, the local voidage would be maximum at the center of the unit, which translated in the bed shear following the same pattern: stronger values in the core, with a maximum at the unit center.
Finally, it should be noticed that redistributing the lateral injection pattern has only a limited impact on the bed shear: the overall range of values remains similar, and the top and bottom sections of the bed have kept the same bed shear values. Thus, with the
equations used, the redistribution only modifies the location and intensity of the bed shear transition in the medium section of the bed.

Figure 5-10. Fraction of agglomerates reaching the bottom of the Fluid Coker with liquid

\( \rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, \ D_{\text{Agg,0}} = 1.0 \text{ cm}, \ (L/S)_0 = 0.31 \)

Figure 5-10 presents the fraction of agglomerates produced reaching the stripper with liquid. Figure 5-11 presents the average bed shear encountered by agglomerates carrying liquid. Similar to previous results, only turning off and redistributing Bank A leads to significant improvements compared to the base case: the number of agglomerates reaching the bottom of the Fluid Coker reduced by a factor of 0.53 (- 47%). The other variations from the base case led to neutral to negative outcomes.

This decrease in the fraction of agglomerates produced reaching the stripper with liquid can be partially explained by an increased average bed shear encountered by the wet-agglomerate. This increase is moderate when Bank A is turned off and redirected (Factor of 1.03) and even smaller for the other variations from the base case (Factor of 1.01-1.02). As mentioned previously, drying is another parameter contributing to this decrease in the fraction of agglomerates produced reaching the stripper with liquid.
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Figure 5-11. Average bed shear encountered by the wet-agglomerate

\[(\rho_{\text{Agg},0} = 1100 \text{ kg}\cdot\text{m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31)\]

It should be noted that despite giving valuable information, the average bed shear encountered by agglomerates carrying liquid is also impacted by the drying. Indeed, wetter agglomerates can reach the bottom of the bed before being destroyed more often than drier agglomerates. Since the lower part of the bed is the region with the lowest bed shear, it mechanically reduces the average bed shear encountered, intrinsically correlating the fraction of agglomerates produced reaching the stripper with liquid with the average bed shear encountered by agglomerates carrying liquid. When the first one increases, the second one tends to decrease.

5.1.3.2.4 Re-wetting

As presented in Chapter 2 – Section 2.3.2 & Section 2.8, re-wetting plays a significant role in determining the amount of liquid carried under to the bottom of the Fluid Coker. Indeed, after a re-wetting, the liquid-to-solid ratio (L/S) of the agglomerate is increased. Thus, this re-wetting has three main effects, interacting with the drying and breakage mechanisms: 1) more time is required for complete drying, 2) the agglomerate saturation (and therefore strength) is increased, 3) if an agglomerate reaches the stripper zone, it will carry more liquid.
An agglomerate can be re-wetted more than once. The location of the last re-wetting of agglomerates carrying liquid to the stripper zone is crucial as each re-wetting “resets” the agglomerate properties. The previous re-wetttings are not inconsequent, nonetheless, as they can be described as “relays” necessary for the agglomerate to overcome long travel times (leading to drying) or traverse zones of high bed shear with a larger strength.

The lowest location of re-wetting can almost be “pseudo-formation” zones: agglomerates with their last re-wetting at Bank A are similar to the agglomerates produced at Bank A if they reach the stripper zone. This result indicates a clear interest in focusing on the bank responsible for most of the last re-wetting of agglomerates carrying liquid to the stripper zone to maximize the impact of any modification.

Figure 5-12 presents the fraction of agglomerates re-wetted and the location of the last agglomerate re-wetting bank. It should be noted that the overwhelming majority of agglomerates carrying liquid to the bottom of the Fluid Coker have been re-wetted (> 70%) for any configuration (as already discussed in Chapter 2 – Section 2.8). In addition, turning
off and redistributing any bank reduces re-wetting. This result is expected as shutting down one bank reduces the number of available production zones by $\frac{1}{5}$th.

The lowest active bank is associated with most of last re-wetting of agglomerates carrying liquid to the stripper zone for all the cases tested. For instance, Bank A, when active, contributes to at least 50% of the last re-wetting (60% if the 5 banks are used). When A is off, Bank B (the lowest active bank) becomes dominant and contributes to 50% of the last re-wetting. On the other hand, the top two banks have a minor impact, which is consistent with the fact that few agglomerates travel quickly from the upper part of the bed to the stripper zone.

5.1.3.2.5 Summary of the observations

The analysis of the impact of a lateral injection redistribution on the mechanisms of the model can be summarized as follows:

- Production-to-stripper time & drying: turning off and redistributing the lowest bank (Bank A) is the only variation from the base case significantly lengthening the production-to-stripper time (and therefore allowing for more drying), by a factor of 1.16 (50% fastest agglomerates). This result is due to a combination of a longer minimum distance between the production zone and the bottom Fluid Coker and more favourable agglomerate hydrodynamics. Indeed, in the lower part of the bed, bubbles and their wakes can carry up agglomerates higher without being disturbed by lateral flow.

- Bed shear & breakage: turning off and redistributing the lowest bank (Bank A) is the only variation from the base case leading to a significant decrease in the number of agglomerates reaching the stripper zone with liquid, by a factor of 0.53. This result is due to a higher average bed shear encountered by agglomerates carrying liquid (from the modified bed hydrodynamics and the modified agglomerates trajectories) and a prolonged drying (leading to weaker agglomerates).

- Re-wetting: this mechanism is essential to compute the predicted liquid losses, especially the lowest bank of re-wetting. The lowest active bank act as a “reset” for
the liquid carried by agglomerates. Any lateral injection redistribution moderately reduces re-wetting, but the impact is limited as long as Bank A is still active.

5.2 Impact of nozzle penetration

This section will investigate the impact of nozzle penetration on the model predictions about the liquid carried under to the bottom of the Fluid Coker.

5.2.1 Variation of nozzles penetration in commercial Fluid Coker

In a commercial Fluid Coker, nozzles penetration decreases during a run (see Chapter 3 – Table 3-1). This decrease is due to a build-up of coke on the walls, which reduces the nozzle penetration over time by reducing the internal diameter of the Fluid Coker. At the beginning of a run, the typical nozzle penetration is around 16% of the radius (Wormsbecker, McMillan, et al., 2021). In the worst case scenario, just before a maintenance shutdown, nozzles could have an almost null effective penetration (0% of the radius).

5.2.2 Impact of variation of nozzles penetration for a “representative” agglomerate

As presented in Chapter 3 – Section 3.1.2.2, in this research the experimental unit uses nozzles whose penetration can be adjusted. Several nozzle positions were tested through two types of configurations. First, a configuration with all nozzles using the same penetration. Four penetrations were tested: 0 cm (0 % of radius), 1 cm (8 % of radius), 2 cm (16 % of radius), and 3 cm (24 % of radius). Then, a second configuration with different nozzles using different penetrations: 0 and 2 cm every other bank, with banks A, C & E using a 2 cm penetration (16 % of radius) and banks B & D using a 0 cm penetration (0 % of radius).

Similar to the lateral redistribution effect analysis, this analysis of the effect of nozzle penetration is carried out using a “representative” agglomerate and typical model parameters. Later, all results will be verified for other initial agglomerate properties (Section 5.3) and other model parameters (Section 5.4).
5.2.2.1 Similar penetration for all nozzles

The liquid losses prediction with various penetration (same penetration for all nozzles) are presented in Figure 5-13.

![Figure 5-13](image)

**Figure 5-13.** Fraction of injected liquid reaching the bottom of the Fluid Coker.

Impact of nozzle penetration (same penetration for all nozzles)

\[(\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31)\]

The significance of the variations is analyzed using the same statistical analysis as in Section 5.1.3.1 (Difference of means, presented in detail in Appendix N: Statistical analysis of the model results). The results are shown in Table 5-2 and Table 5-3.

Using Figure 5-13, Table 5-2 and Table 5-3, two patterns emerge, depending on the redistribution profile.

First, when 5 banks are used, an increased nozzle penetration leads to a small increase in liquid losses. Going from a penetration of 0 cm (0 % of bed radius) to 2 cm (16 % of bed radius) modifies the liquid losses by a factor of 1.22. The statistical analysis indicates that these variations are not significant.
Table 5-2. Statistical analysis (Differences of means) – Effect of nozzle penetration for the “representative” agglomerate

Constant penetration, Base case (5 banks)

<table>
<thead>
<tr>
<th></th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>5 Banks</td>
</tr>
<tr>
<td></td>
<td>0 cm</td>
<td>1 cm</td>
</tr>
<tr>
<td></td>
<td>2 cm</td>
<td>3 cm</td>
</tr>
<tr>
<td>Number of samples</td>
<td>5</td>
<td>2</td>
</tr>
<tr>
<td></td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
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<td>1.23 \times 10^{-3}</td>
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<td>1.41 \times 10^{-3}</td>
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<td>1.84 \times 10^{-4}</td>
<td>4.70 \times 10^{-5}</td>
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<td>σ</td>
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<td></td>
<td>2.71 \times 10^{-4}</td>
<td></td>
</tr>
<tr>
<td></td>
<td>t</td>
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</tr>
<tr>
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</tr>
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</tr>
<tr>
<td>Statistically different?</td>
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<td>FALSE</td>
</tr>
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</table>

Table 5-3. Statistical analysis (Differences of means) – Effect of nozzle penetration for the “representative” agglomerate

Constant penetration, 4 banks (Bank A off & redistributed)

<table>
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<th>Comparison</th>
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</thead>
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<td></td>
<td>4 Banks</td>
<td>4 Banks</td>
</tr>
<tr>
<td></td>
<td>(A off)</td>
<td>(A off)</td>
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<tr>
<td></td>
<td>0 cm</td>
<td>1 cm</td>
</tr>
<tr>
<td></td>
<td>1 cm</td>
<td>2 cm</td>
</tr>
<tr>
<td></td>
<td>3 cm</td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
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<tr>
<td></td>
<td>3</td>
<td>2</td>
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<tr>
<td>Average</td>
<td>1.59 \times 10^{-4}</td>
<td>1.78 \times 10^{-4}</td>
</tr>
<tr>
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<td>7.48 \times 10^{-5}</td>
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<td>Standard Deviation</td>
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<td>6.32 \times 10^{-7}</td>
</tr>
<tr>
<td></td>
<td>4.15 \times 10^{-6}</td>
<td>4.88 \times 10^{-6}</td>
</tr>
<tr>
<td>σ</td>
<td>3.17 \times 10^{-5}</td>
<td>2.95 \times 10^{-5}</td>
</tr>
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<td></td>
<td>3.19 \times 10^{-5}</td>
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</tr>
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</tr>
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<td></td>
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<td>3.50</td>
</tr>
<tr>
<td></td>
<td>3.71</td>
<td></td>
</tr>
<tr>
<td>Statistically different?</td>
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<td>TRUE</td>
</tr>
</tbody>
</table>

On the other hand, when 4 banks are used with A off and its flow redistributed, an increased nozzle penetration leads to a small decrease in liquid losses. Going from a penetration of 0 cm (0 % of bed radius) to 2 cm (16 % of bed radius) modifies the liquid losses by a
factor of 0.57. The statistical analysis indicates significant variations for an increased nozzle penetration greater or equal to 2 cm.

As mentioned above, a decreasing nozzle penetration corresponds to the temporal evolution of the effective nozzle penetration in commercial Fluid Cokers. Therefore, these results suggest that as the run progresses: if 5 banks are used, the liquid losses do not vary significantly, and if 4 banks are used (A off and redistributed), the liquid losses moderately increase.

The mechanisms behind these results will be investigated in Section 5.2.4.

5.2.2.2 Variable nozzles penetration

The liquid losses predictions with an uneven penetration pattern (0 and 2 cm every other bank) are compared to the liquid losses predicted for a constant nozzle penetration of 0 cm (0 % of the bed radius) and 2 cm (16 % of the bed radius). The results are presented in Figure 5-14.

Figure 5-14. Fraction of injected liquid reaching the bottom of the Fluid Coker. Impact of nozzle penetration (same penetration for all nozzles vs. variable penetration) $\left( \rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg,0}} = 1.0 \text{ cm}, (L/S)_0 = 0.31 \right)$
The significance of the variations is analyzed using the same statistical analysis (difference of means) as in Section 5.1.3.1. The results are presented in Table 5-4.

Using Figure 5-14 and Table 5-4, it appears that the configuration with different nozzles using different penetrations (0 and 2 cm every other bank) is not desirable. Indeed, when used, the liquid losses predicted become worse by a factor of 3.2 to 14.9.

While this result is indicative of a configuration that should be avoided, it is nonetheless interesting as it demonstrates that nozzle configurations with uneven penetrations can significantly modify liquid losses. Therefore, other nozzle configurations should be investigated to identify if some of them could help reduce liquid losses.

Again, the mechanisms behind these results will be investigated in Section 5.2.4.

Table 5-4. Statistical analysis (Differences of means) – Effect of nozzle penetration

<table>
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<tr>
<th>Variable penetration</th>
<th>5 Banks</th>
<th>4 Banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td><strong>0 cm</strong></td>
<td>1.16 $10^{-3}$</td>
<td>3.57 $10^{-3}$</td>
</tr>
<tr>
<td><strong>0-2 cm</strong></td>
<td>2.69 $10^{-4}$</td>
<td>1.04 $10^{-3}$</td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
<td>5</td>
<td>2</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>$1.16 10^{-3}$</td>
<td>$3.57 10^{-3}$</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>$2.69 10^{-4}$</td>
<td>$1.04 10^{-3}$</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{STUDENT}$ (two-tailed, 99%)</td>
<td>4.03</td>
<td>3.71</td>
</tr>
</tbody>
</table>
5.2.3 Verification of the impact of nozzle penetration for various lateral redistribution (case of similar penetration for all nozzles)

Figure 5-15. Fraction of injected liquid reaching the bottom of the Fluid Coker. Impact of nozzle penetration (same distance for all nozzles) for different configurations 

\( \rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31 \)

Figure 5-15 shows that the liquid losses variations due to nozzle penetration varying between 0 % and 16 % of the bed radius (same penetration for all nozzles) are negligible compared to liquid losses variations due to the lateral injection redistributions. The statistical analysis shows similar results as in Section 5.1.3.1 and is presented in Appendix N: Statistical analysis of the model results.

5.2.4 Impact of nozzle penetration on mechanisms included in the model

Similar to Section 5.1.3.2, this section investigates how the variation of the nozzle penetration impacts the various mechanisms bundled in the model (as presented in Chapter 2). Again, it should be noted that these mechanisms are interconnected.
Based on the previous results, the analysis was conducted for two lateral injection distributions: the base case and the case with Bank A off and redistributed. In addition, for the configuration with the same penetration for all nozzles, only the case with a 2 cm (16% of bed radius) and 0 cm (0% of bed radius) penetration are considered. These cases were chosen as they match the nozzle penetration respectively at the beginning and the end of typical runs of a commercial Fluid Coker.

5.2.4.1 Wetting at formation

Since the empirical equation used to calculate the fraction of liquid injected that is trapped into an agglomerate of a given size only considered the cross-sectional superficial gas velocity (Li, 2016), this component of the model is not impacted by changing the nozzle penetration length.

A possible improvement of the model might be to switch to an equation considering the local velocity.

5.2.4.2 Production-to-stripper time & drying

Figure 5-16 compares the cumulative distribution of the production-to-stripper time of configuration with various nozzle penetrations. Two lateral redistribution patterns are presented, the base case (5 banks) and the case with 4 banks (A off and redistributed).

Figure 5-16 clearly shows that for both redistribution patterns, an increased nozzle penetration (constant penetration for all nozzles) leads to a longer production-to-stripper time. This is because production zones closer to the wall are not in a region as rich in bubbles as production zones closer to the core of the bed. The variation is more pronounced for the 4 banks case (with A off and redistributed): the travel time is increased by a factor of 2.94 versus a factor of 1.94 for the base case (50% fastest agglomerates, with nozzles penetration going from 0 to 2 cm penetration).
Figure 5-16. Cumulative distribution of the production-to-stripper time of tracer-agglomerates ($D_{\text{Tracer}} = 1.1$ cm, $\rho_{\text{Tracer}} = 1100$ kg·m$^{-3}$, with replicates)

a) 5 Banks, b) 4 Banks (A off)

For the configuration with different nozzles using different penetrations (0 and 2 cm every other bank), agglomerates undergo a strong bypassing, visible for both redistribution...
patterns. The production-to-stripper time distribution is skewed towards a shorter time, suggesting the apparition of a strong downward agglomerate corridor.

Contrary to lateral injection redistribution, the only parameter that can explain these results is a modification of the bed hydrodynamics. Figure 5-17 gives some insight into the reason behind these modifications of the agglomerate hydrodynamics.

Figure 5-17-b) shows how moving the nozzle towards the center (same penetration for all nozzles) helps to concentrate gas at the center, explaining the longer production-to-stripper time distribution observed. Figure 5-17-c) is not as clear but gives some indications about the apparition of stronger downward channels than the base case, on the periphery of the unit between elevations of 0.5 m and 0.8 m). It might also be possible that the agglomerates

---

**Figure 5-17.** Vertical median tracer velocity (5 injection banks) as a function of the nozzle penetration

a) $L_{IN} = 0$ cm, b) $L_{IN} = 2$ cm, c) A, C & E: $L_{IN} = 2$ cm / B & D: $L_{IN} = 0$ cm

($D_{Tracer} = 0.9$ cm, $\rho_{Tracer} = 1100$ kg·m$^{-3}$)
that are moving down quickly have a complex trajectory that cannot easily be captures by this representation.

5.2.4.3 Bed shear & breakage

Figure 5-18 presents the fraction of agglomerates produced reaching the stripper with liquid. Figure 5-19 shows the average bed shear encountered by agglomerates carrying liquid.

Figure 5-18 clearly shows that for both redistribution patterns, an increased nozzle penetration (constant penetration for all nozzles) leads to a decreased fraction of agglomerates produced reaching the stripper with liquid. The decrease is higher for the 4 banks case (A off and redistributed): factor of 0.42 versus factor of 0.66 for the base case (with nozzles penetration going from 0 to 2 cm penetration). This is because the production zones are closer to the bed core, a zone with a higher bed shear (results in Figure 5-19 tend to confirm this observation). Thus, the agglomerates are more likely to encounter high shear before having the opportunity to move down to zones of low bed shear. Once again, this effect combined with the drying over a longer period of time, weakens the agglomerates.

\( a \)
**Figure 5-18.** Fraction of agglomerates reaching the bottom of the Fluid Coker with liquid

a) Constant nozzle penetration, b) Variable nozzle penetration

\((\rho_{\text{Agg},0} = 1100 \text{ kg}\cdot\text{m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31)\)
Figure 5-19. Fraction of agglomerates reaching the bottom of the Fluid Coker with liquid

\[ (\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31) \]

For the second nozzle configuration, with different nozzles using different penetrations (0 and 2 cm every other bank), a significant increase of the fraction of agglomerates produced reaching the stripper with liquid is observed. Figure 5-19 indicates that the average bed shear encountered by wet-agglomerates is higher than the one for the case of a constant nozzle penetration of 0 cm. This suggests that the increase of the fraction of agglomerates produced reaching the stripper with liquid is solely due to the shorter drying time (bypassing) of the fastest agglomerates.

5.2.4.4 Re-wetting

Figure 5-20 presents the fraction of agglomerates re-wetted and the location of the last agglomerate re-wetting bank for the first nozzle configuration, with the same penetration for all nozzles. Figure 5-21 presents the fraction of agglomerates re-wetted and the location of the last agglomerate re-wetting bank for the second nozzle configuration, with different nozzles using different penetrations (0 and 2 cm every other bank).
Figure 5-20. Location of last agglomerate re-wetting bank, impact of nozzle penetration
(Constant nozzle penetration)

a) 5 Banks, b) 4 Banks (A off)

\( \rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, \ D_{\text{Agg},0} = 1.0 \text{ cm}, \ (L/S)_0 = 0.31 \)
Figure 5-21. Location of last agglomerate re-wetting bank, impact of nozzle penetration
(Variable nozzle penetration)
a) 5 Banks, b) 4 Banks (A off)
\[ \rho_{\text{Agg},0} = 1100 \, \text{kg} \cdot \text{m}^{-3}, \quad D_{\text{Agg},0} = 1.0 \, \text{cm}, \quad (L/S)_0 = 0.31 \]

Figure 5-20 shows minor variations of the fraction of agglomerates re-wetted when the nozzle penetration increases (same penetration for all nozzles). For a penetration of 2 cm
(16% of the radius), the base case presents virtually no reduction (factor of 1.02). In contrast, the case with 4 banks (A off and redistributed) shows a small decrease of the fraction of agglomerates re-wetted (factor of 0.95). Regarding the and location of the last agglomerate re-wetting bank, it should be noted that the proportion of re-wetting at the lowest active bank decreases significantly when the nozzle penetration increases (same penetration for all nozzles).

Figure 5-21 shows minor variations of the fraction of agglomerates re-wetted with the second nozzle configuration, with different nozzles using different penetrations (0 & 2 cm every other bank). Regarding the location of the last agglomerate re-wetting bank, the results are difficult to interpret.

5.2.4.5 Summary of the observations

The analysis of the impact of the nozzle penetration on the mechanisms of the model (for two redistribution patterns, the base case, with 5 banks, and the case with 4 banks, with the Bank A off and its flow redistributed) can be summarized as follow:

- Agglomerate travel time from the production sites to the stripper & drying:
  o For the configuration with a constant penetration length for all nozzles, a longer nozzle penetration leads to longer production-to-stripper time distribution (and therefore longer drying) for both redistribution patterns. The production-to-stripper time increase is more pronounced for the 4 banks case (A off and redistributed case)
  o The second nozzle configuration, with different nozzles using different penetrations (0 and 2 cm every other bank), creates more bypassing for both redistribution patterns.

- Bed shear & breakage:
  o For the configuration with a constant penetration length for all nozzles, a longer nozzle penetration leads to more breakage for both redistribution patterns. This result is due to a higher average bed shear encountered by agglomerates carrying liquid (due to the modified bed hydrodynamics and
the modified agglomerates trajectories) and a prolonged drying (leading to weaker agglomerates). The breakage increase is slightly more pronounced for the 4 banks case (with A off and redistributed) than for the base case (5 banks).

- The second nozzle configuration, with different nozzles using penetrations of 0 and 2 cm every other bank, leads to less breakage. This result is mainly controlled by the shortened drying of bypassing agglomerates, which leads to much stronger agglomerates.

- Re-wetting:
  - For the configuration with a constant penetration length for all nozzles, a longer nozzle penetration, the overall re-wetting decreases slightly. The re-wetting decrease is slightly more pronounced for the 4 banks (A off and redistributed case) than for the base case. For both redistribution cases, less re-wetting tends to occur in the lowest active bank.
  - The nozzle configuration, with different nozzles using different penetrations (0 and 2 cm every other bank), does not show meaningful variation in the overall re-wetting. Similarly, the location of the last re-wetting of agglomerates carrying liquid to the stripper zone is not significantly modified.

The base case (5 banks) and the 4 banks case (A off and redistributed) show similar underlying mechanisms when the nozzle penetration is increased (constant penetration length for all nozzles). The difference in predicted liquid losses between these two cases is due to the amplitude of variation of these mechanisms and the impact of the re-wetting at Bank A. Removing the lowest bank (Bank A) makes the system more sensitive to the change related to an increased nozzle penetration, especially the increase in the formation-to-stripper time distribution. Therefore, it results in a small decrease in liquid losses for the case with 4 banks (A off and redistributed) when no significant variations are visible for the base case (5 banks).
5.3 Effect of agglomerate properties

In commercial Fluid Coker, there is a wide range of agglomerates formed and moving through the bed at any time (see Chapter 2). So far, the model was used by considering only a “representative” agglomerate with a single size, density, and initial liquid content. This section investigates the possible impact of agglomerate properties on the results predicted by the model.

It should be noted that two categories of agglomerate properties need to be investigated. The first category to consider is related to the experimental tracer-agglomerate properties such as diameter and density. This tracer-agglomerate is studied from the point of view of production-to-stripper time distribution. It is assumed that two tracers with similar production-to-stripper time distribution behave similarly. Then there are the modelled wet-agglomerates, which follow the trajectories of the experimental tracer-agglomerate but with a variable amount of liquid assigned by the model. These modelled agglomerates are characterized by: the initial diameter, the initial density, and the initial liquid-to-solid ratio (wetness).

5.3.1.1 Overview of the effect of initial agglomerate properties on liquid losses

![Graph showing the effect of initial agglomerate properties on liquid losses]
**Figure 5-22.** Fraction of injected liquid reaching the bottom of the Fluid Coker, effect of lateral redistribution patterns and initial agglomerate parameters

$\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, L_{\text{IN}} = 0 \text{ cm}$

Figure 5-22 shows liquid losses predicted when the lateral redistribution pattern is modified for several initial diameters and initial liquid-to-solid ratios. The significance of the variations is analyzed using the same statistical analysis (Difference of means) as in Section 5.1.3.1. The results are presented in Table 5-5.

It is visible in Figure 5-22 that the largest and wettest agglomerates ($D_{\text{Agg},0} = D_{\text{Agg}} = 1.5 \text{ cm}, (L/S)_0 = 0.62$) are not the wet-agglomerates carrying the highest fraction of injected liquid to the stripper zone. As discussed in Chapter 2 – Section 2.8.1, this is due to the inclusion of the breakage due to bed shear in the model calculations. Due to their lower strength, these larger agglomerates are more prone to breakage than smaller agglomerates, reducing their importance in the overall liquid losses. On the other hand, in a model not considering breakage due to bed shear these larger and wetter agglomerates would be the ones carrying the most liquid to the stripper (due to their slower drying and higher initial liquid content).

Figure 5-22 and Table 5-5 show that the relative variation of predicted liquid losses observed with the “representative” agglomerate is the same for any diameter and liquid-to-solid ratio tested. More configurations are presented in Appendix M: Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters, with their statistical analysis in Appendix N: Statistical analysis of the model results.

Figure 5-23 shows the liquid losses predicted when the nozzle penetration (same penetration for all nozzles) is modified, for several combinations of initial diameter and initial liquid-to-solid ratio. Figure 5-24 shows the liquid losses predicted with a variable nozzle penetration (0 and 2 cm every other bank). The significance of the variations is analyzed using the same statistical analysis (Difference of means) as in Section 5.1.3.1. The results are presented in Table 5-6 Table 5-7.
Table 5-5. Statistical analysis (Differences of means) – Effect of redistribution and initial agglomerate parameters

<table>
<thead>
<tr>
<th>ρ_{Agg,0} = 1100 \text{ kg·m}^{-3}</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>5</td>
<td>6</td>
</tr>
<tr>
<td>t_{STUDENT} (two-tailed, 99%)</td>
<td>3.25</td>
<td>4.03</td>
</tr>
</tbody>
</table>

| D_{Agg,0} = 1.0 \text{ cm, (L/S)}_{0} = 0.31 | Average | 1.16 \times 10^{-3} | 1.59 \times 10^{-4} | 9.80 \times 10^{-5} | 1.04 \times 10^{-4} | 8.43 \times 10^{-4} | 1.03 \times 10^{-3} |
| Standard Deviation                | 2.69 \times 10^{-4} | 3.17 \times 10^{-5} | 3.47 \times 10^{-6} | 2.21 \times 10^{-4} | 9.03 \times 10^{-5} | 4.91 \times 10^{-4} |
| σ                                 | 2.02 \times 10^{-4} | 2.69 \times 10^{-5} | 3.03 \times 10^{-6} | 2.54 \times 10^{-4} | 4.11 \times 10^{-4} |
| |t|                                | 8.17      | 0.80         | 0.47         | 1.71         | 0.38         |
| Statistically different?          | TRUE      | FALSE        | FALSE        | FALSE        | FALSE        |

| D_{Agg,0} = 0.5 \text{ cm, (L/S)}_{0} = 0.10 | Average | 1.29 \times 10^{-3} | 2.97 \times 10^{-4} | 1.64 \times 10^{-5} | 1.26 \times 10^{-4} | 8.34 \times 10^{-4} | 1.20 \times 10^{-3} |
| Standard Deviation                | 1.40 \times 10^{-4} | 9.65 \times 10^{-5} | 5.82 \times 10^{-6} | 2.28 \times 10^{-4} | 3.22 \times 10^{-4} | 8.24 \times 10^{-4} |
| σ                                 | 1.48 \times 10^{-4} | 1.41 \times 10^{-5} | 2.33 \times 10^{-6} | 3.04 \times 10^{-4} | 6.87 \times 10^{-4} |
| |t|                                | 8.25      | 2.73         | 0.14         | 1.84         | 0.15         |
| Statistically different?          | TRUE      | FALSE        | FALSE        | FALSE        | FALSE        |

| D_{Agg,0} = 1.5 \text{ cm, (L/S)}_{0} = 0.62 | Average | 7.54 \times 10^{-5} | 1.54 \times 10^{-6} | 9.99 \times 10^{-7} | 1.00 \times 10^{-6} | 4.82 \times 10^{-5} | 1.09 \times 10^{-4} |
| Standard Deviation                | 1.77 \times 10^{-5} | 5.22 \times 10^{-7} | 3.54 \times 10^{-7} | 3.29 \times 10^{-6} | 1.49 \times 10^{-5} | 1.41 \times 10^{-7} |
| σ                                 | 1.40 \times 10^{-5} | 1.77 \times 10^{-5} | 1.78 \times 10^{-5} | 1.93 \times 10^{-5} | 1.77 \times 10^{-5} |
| |t|                                | 8.36      | 1.65         | 1.65         | 1.94         | 2.29         |
| Statistically different?          | TRUE      | FALSE        | FALSE        | FALSE        | FALSE        |
Figure 5-23. Fraction of injected liquid reaching the bottom of the Fluid Coker, effect of nozzle penetration (constant length) and initial agglomerate parameters

\( \rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3} \)

a) 5 Banks, b) 4 Banks (A off)
Figure 5-24. Fraction of injected liquid reaching the bottom of the Fluid Coker, effect of nozzle penetration (variable length) and initial agglomerate parameters

\( \rho_{\text{Agg},0} = 1100 \, \text{kg} \cdot \text{m}^{-3} \)

a) 5 Banks, b) 4 Banks (A off)
Table 5-6. Statistical analysis (Differences of means) – Effect of nozzle penetration for different initial agglomerate parameters

Constant penetration, Base case (5 banks)

<table>
<thead>
<tr>
<th>$D_{\text{Agg},0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0 cm</td>
<td>1 cm</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>1.16 $10^{-3}$</td>
<td>1.28 $10^{-3}$</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>2.69 $10^{-4}$</td>
<td>2.65 $10^{-4}$</td>
</tr>
<tr>
<td><strong>$\sigma$</strong></td>
<td>3.17 $10^{-4}$</td>
<td>2.93 $10^{-4}$</td>
</tr>
<tr>
<td>**$</td>
<td>t</td>
<td>$**</td>
</tr>
<tr>
<td><strong>Statistically different?</strong></td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>$D_{\text{Agg},0} = 0.5 \text{ cm}$, $(L/S)_0 = 0.10$</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0 cm</td>
<td>1 cm</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>7.75 $10^{-4}$</td>
<td>1.04 $10^{-3}$</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>7.14 $10^{-4}$</td>
<td>4.66 $10^{-3}$</td>
</tr>
<tr>
<td><strong>$\sigma$</strong></td>
<td>7.15 $10^{-4}$</td>
<td>7.29 $10^{-4}$</td>
</tr>
<tr>
<td>**$</td>
<td>t</td>
<td>$**</td>
</tr>
<tr>
<td><strong>Statistically different?</strong></td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>$D_{\text{Agg},0} = 1.5 \text{ cm}$, $(L/S)_0 = 0.62$</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0 cm</td>
<td>1 cm</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>7.54 $10^{-5}$</td>
<td>7.27 $10^{-5}$</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>1.77 $10^{-5}$</td>
<td>1.63 $10^{-6}$</td>
</tr>
<tr>
<td><strong>$\sigma$</strong></td>
<td>1.77 $10^{-5}$</td>
<td>1.78 $10^{-5}$</td>
</tr>
<tr>
<td>**$</td>
<td>t</td>
<td>$**</td>
</tr>
<tr>
<td><strong>Statistically different?</strong></td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>
Table 5-7. Statistical analysis (Differences of means) – Effect of nozzle penetration for different initial agglomerate parameters

Variable penetration, 4 banks case (A off and redistributed)

<table>
<thead>
<tr>
<th>4 banks (A off and redistributed)</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0 cm</td>
<td>1 cm</td>
</tr>
<tr>
<td>$\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>6</td>
<td>2</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}}$ (two-tailed, 99%)</td>
<td>3.71</td>
<td>3.50</td>
</tr>
<tr>
<td>$D_{\text{Agg},0} = 1.0 \text{ cm}$, $\text{(L/S)}_0 = 0.31$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>1.59 $10^4$</td>
<td>1.78 $10^4$</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>3.17 $10^5$</td>
<td>6.32 $10^7$</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>3.17 $10^5$</td>
<td>2.95 $10^5$</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>TRUE</td>
</tr>
<tr>
<td>$D_{\text{Agg},0} = 0.5 \text{ cm}$, $\text{(L/S)}_0 = 0.10$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>1.48 $10^4$</td>
<td>1.92 $10^4$</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.74 $10^4$</td>
<td>2.03 $10^5$</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>1.74 $10^4$</td>
<td>1.62 $10^4$</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
<tr>
<td>$D_{\text{Agg},0} = 1.5 \text{ cm}$, $\text{(L/S)}_0 = 0.62$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>1.54 $10^6$</td>
<td>4.99 $10^7$</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>5.22 $10^7$</td>
<td>1.77 $10^9$</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>5.22 $10^7$</td>
<td>4.77 $10^7$</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>TRUE</td>
</tr>
</tbody>
</table>

Figure 5-23, Table 5-6 and Table 5-7 show that, when changing the nozzle penetration (constant length for all nozzles), the relative variation of predicted liquid losses observed with the “representative” agglomerate is the same for most diameters and liquid-to-solid ratios tested. For the smallest and driest agglomerates, the variations become non-significant for the case with 4 banks (Bank A off and redistributed). More configurations are presented in Appendix M: Liquid losses predicted by the model for
multiple initial agglomerate parameters and alternative model parameters, with their statistical analysis in Appendix N: Statistical analysis of the model results. They show that the variations always a similar trend and are relatively small. These variations are not always significant, highlighting the small amplitude of the change implied by an increased nozzle penetration (for a constant length for all nozzles) and suggesting the need for a more thorough investigation of the effect of such changes for the whole wet-agglomerate population.

Figure 5-24, Table 5-6 and Table 5-7 show that, when using an uneven nozzle penetration (0 and 2 cm every other bank), the relative variation of predicted liquid losses observed with the “representative” agglomerate is the same for most diameters and liquid-to-solid ratios tested. The variations become non-significant for the case with 4 banks (Bank A off and redistributed) for the largest and wettest agglomerates. More configurations are presented in Appendix M: Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters, with their statistical analysis in Appendix N: Statistical analysis of the model results. The variations are significant except for the largest and wettest agglomerates, suggesting that they might be less sensitive to the changes applied for the nozzle rearrangement.

The following sections break down the effect of each initial agglomerate parameter (density, diameter, and liquid-to-solid ratio) when the lateral redistribution pattern is modified. As changing the nozzle penetration has a much smaller impact on the predicted liquid losses, the effect of each initial agglomerate parameter when the nozzle penetrations are modified is presented in Appendix M: Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters.

5.3.1.2 Agglomerate density

This section investigates the effect of the density of the experimental tracer-agglomerate and the density of the modelled wet-agglomerates on results.
5.3.1.2.1  Tracer experimental density

The tracer-agglomerate densities investigated in this research range between 900 and 1100 kg·m$^{-3}$ (Chapter 3 - Figure 3-14). The impact of this tracer-agglomerate density on measured production-to-stripper time distribution is presented in Figure 5-25.

![Cumulative distribution of the production-to-stripper time of tracer-agglomerates, the impact of tracer density (5 Banks, D_{Tracer} = 1.0 cm)](image)

**Figure 5-25.** Cumulative distribution of the production-to-stripper time of tracer-agglomerates, the impact of tracer density (5 Banks, D_{Tracer} = 1.0 cm)

The results indicate that, as expected, denser agglomerates move down through the bed significantly quicker than lighter ones. Therefore, for a given experimental configuration to study agglomerates with varying densities in the model, at least an experimental run is required for each density studied.

5.3.1.2.2  Agglomerate density in the model

Figure 5-26 presents the liquid losses predicted by the model for two agglomerates (and tracer-agglomerates) of varying initial densities.

In addition of the longer formation-to-stripper time, lighter agglomerates are associated with lower required time for drying $t_C$ (they dry more quickly) and lower simplified
breakage parameter $B_{Agg,0}$ (they break more easily). These combined mechanisms explain why they carry less liquid to the stripper.

The relative variation of predicted liquid losses observed with the “representative” agglomerate is the same for any density tested. The statistical analysis shows similar results for both densities as in Section 5.1.3.1 and is presented in Appendix N: Statistical analysis of the model results.

![Graph]

**Figure 5.26.** Fraction of injected liquid reaching the bottom of the Fluid Coker

($D_{Agg,0} = 1.0$ cm, (L/S)$_0 = 0.31$, $L_{IN} = 0$ cm)

5.3.1.3 Agglomerate diameter

This section investigates the effect of the diameter of the experimental tracer-agglomerate and the density of the modelled wet-agglomerates on results.

5.3.1.3.1 Tracer experimental diameter

The range of tracer-agglomerate diameters investigated in this research ranges between 0.6 and 1.4 cm (Chapter 3 - Figure 3-14). The impact of the tracer-agglomerate diameter on measured production-to-stripper time distribution is presented in Figure 5.27.
The results indicate that the cumulative distribution of the production-to-stripper time is similar for the range of tracer-agglomerate diameter used. Therefore, to study agglomerates with varying diameters in the model, there is no need for one experimental run per diameter studied for a given experimental configuration. The experiments can be conducted with a single tracer-agglomerate diameter, and the wet-agglomerates in the model can be used with other diameters (to control for drying, breakage, …) since the trajectories are similar. Therefore, to limit the required experimentation time, most experiments were conducted with a tracer agglomerate of diameter $D_{\text{Tracer}} = 1.0$ cm.

![Figure 5-27. Cumulative distribution of the production-to-stripper time of tracer-agglomerates, the impact of tracer diameter (5 Banks, $\rho_{\text{Tracer}} = 1100 \text{ kg}\cdot\text{m}^{-3}$)](image)

### 5.3.1.3.2 Agglomerate diameter in model

To be fully understood, changes of agglomerates initial diameter require considering the effect of agglomerate formation (see Section 5.1.3.2.1). The wet-agglomerates diameter has a significant impact on the fraction of liquid injected captured by initial wet-agglomerates (Li, 2016), independent of the fraction of liquid initially in wet-agglomerates reaching the bottom of the Fluid Coker. Figure 5-28 shows how much the fraction of the injected liquid captured by the initial wet-agglomerate varies when the wet-agglomerate initial diameter $D_{\text{Agg},0}$ is modified.
Figure 5-28. Variation of the fraction of liquid injected transferred to initial agglomerate as a function of its diameter (average of all banks for the 5 banks lateral distribution pattern)

Figure 5-29, Figure 5-30 and Figure 5-31 present the liquid losses predicted by the model for three agglomerates (and tracer-agglomerates) of varying diameters for various initial liquid-to-solid ratios. The figure includes the overall liquid losses (from injection to the stripper) and the liquid losses, excluding the impact of agglomerates formation (from agglomerates already formed to the stripper).

For all the agglomerate diameters tested, the statistical analysis shows results similar to those obtained for a “representative” agglomerate in Section 5.1.3.1. This analysis is presented in Appendix N: Statistical analysis of the model results.

The relative variation of predicted liquid losses observed with the “representative” agglomerate is the same for any diameter tested.
**Figure 5-29.** Fraction of liquid reaching the bottom of the Fluid Coker

\( \rho_{\text{Agg},0} = 1100 \text{ kg·m}^{-3}, \ (L/S)_0 = 0.10, \ L_{\text{IN}} = 0 \text{ cm} \)

a) Fraction of injected liquid reaching the bottom of the Fluid Coker,

b) Fraction of liquid in initial agglomerates reaching the bottom of the Fluid Coker
Figure 5-30. Fraction of liquid reaching the bottom of the Fluid Coker

\[ \rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, (L/S)_0 = 0.31, L_{\text{IN}} = 0 \text{ cm} \]

a) Fraction of injected liquid reaching the bottom of the Fluid Coker,
b) Fraction of liquid in initial agglomerates reaching the bottom of the Fluid Coker
Figure 5-31. Fraction of liquid reaching the bottom of the Fluid Coker

\[(\rho_{Agg,0} = 1100 \text{ kg·m}^{-3}, (L/S)_0 = 0.62, L_{IN} = 0 \text{ cm})\]

a) Fraction of injected liquid reaching the bottom of the Fluid Coker,

b) Fraction of liquid in initial agglomerates reaching the bottom of the Fluid Coker

Three competing mechanisms govern the magnitude of the liquid losses variations when the agglomerate diameter varies:
- Drying: larger agglomerates dry slower than smaller ones. If no breakage or re-wetting occurs, larger agglomerates will carry more liquid to the stripper. It is especially visible for the strongest agglomerates, such as those presented in Figure 5-29. In the sub-chart Figure 5-29-b), which considers only the liquid present in agglomerate initially formed reaching the bottom of the reactor, larger agglomerates carry more liquid to the bottom of the reactor.
- Bed shear & breakage: at a constant liquid content and density, larger agglomerates have a higher voidage. They are weaker and more prone to breakage, carrying less liquid to the bottom of the reactor. This is especially visible in Figure 5-31.
- Formation: smaller agglomerates capture a bigger fraction of the liquid injected (more small agglomerates are formed than larger ones). This mechanism explains the variations between the sub-charts a) and b) of Figure 5-29, Figure 5-30 & Figure 5-31.

When combining these mechanisms, smaller agglomerates will contribute the most to the liquid losses to the stripper.

5.3.1.4 Initial Liquid-to-Solid ratio (L/S) of agglomerate

Contrary to the agglomerate density and the agglomerate diameter, the initial (L/S) ratio cannot be simulated in the cold gas-solid experimental unit: it is entirely a model parameter. In addition, this parameter is important for the agglomerate saturation and, therefore, the resistance to breakage. It behaves in a non-monotonous way in this regard.

Figure 5-32 uses the simplified initial breakage coefficient $B_{Agg,0}$, introduced in Chapter 2, to present the resistance to breakage of agglomerates (higher $B_{Agg,0}$ means higher resistance to breakage) of a given size and density as a function of their initial liquid-to-solid ratio. A maximum strength is obtained for agglomerates of moderate wetness ($((L/S))_0 \approx 0.25$). The amplitude of the strength variation is more pronounced for smaller agglomerates than for large agglomerates.
Figure 5-32. Evolution of simplified breakage parameter with initial liquid-to-solid ratio, for various agglomerate size ($\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}$)

Figure 5-33 presents the liquid losses predicted by the model for two agglomerates of varying initial liquid-to-solid ratios.

\textit{a)}
For all the agglomerates wetness tested, the statistical analysis shows similar results as in Section 5.1.3.1 and is presented in Appendix N: Statistical analysis of the model results.
The relative variation of predicted liquid losses observed with the “representative” agglomerate is the same for any initial liquid-to-solid ratio tested.

Three competing mechanisms govern the amplitude of the liquid losses variations when the agglomerate initial liquid-to-solid ratio varies:

- Drying: wetter agglomerates dry slower (more liquid to evaporate)
- Bed shear & breakage: at a constant diameter & density, the resistance to breakage is maximum for a moderate wetness \((L/S) \approx 0.25\). This variation is more pronounced for smaller agglomerates, and there is almost no variation for the larger agglomerates.
- Liquid content: wetter agglomerates can potentially carry more liquid to the bottom of the Fluid Coker.

When combining these mechanisms, the impact of the agglomerates will vary depending on the agglomerate size.

- For small agglomerates: shear is negligible, and the dominant mechanism is the drying. Therefore, small wetter agglomerates carry more of the liquid injected in the bed to the bottom of the Fluid Coker.
- For the medium-sized agglomerates: bed shear and drying are competing. Therefore, medium dryer agglomerates carry marginally more liquid injected in the bed, reaching the bottom of the Fluid Coker.
- For the large agglomerates: bed shear is the dominant mechanism, overcoming drying. Therefore, large wetter agglomerates carry more liquid injected in the bed, reaching the bottom of the Fluid Coker.

5.4 Effect of important model parameters

As mentioned previously, the model includes some parameters whose values are not accurately known. This section tests the sensitivity of the liquid losses predictions to the variations of these model parameters. More configurations are presented in Appendix M:
Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters.

5.4.1 Coefficient of re-wetting $C_{\text{Rewet}}$

Figure 5-34 presents the liquid losses predicted by the model for two values of the model parameter $C_{\text{Rewet}}$.

![Figure 5-34](image_url)

**Figure 5-34.** Fraction of injected liquid reaching the bottom of the Fluid Coker as a function of the coefficient of re-wetting $C_{\text{Rewet}}$

\[
\rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, \ D_{\text{Agg,0}} = 1.0 \text{ cm}, \ (L/S)_{0} = 0.31, \ L_{\text{IN}} = 0 \text{ cm}
\]

The statistical analysis shows similar results for the two values of $C_{\text{Rewet}}$ tested as in Section 5.1.3.1 and is presented in Appendix N: Statistical analysis of the model results.

The coefficient of re-wetting $C_{\text{Rewet}}$ has a significant impact on the absolute value of predicted liquid losses. Therefore, future model improvement should include more work to get an accurate value for the coefficient of re-wetting $C_{\text{Rewet}}$. Nonetheless, the relative variations observed with the “representative” agglomerate are conserved.

5.4.2 Effect of bubble size & shear
\( D_{Agg,0} = 0.5 \text{ cm}, (L/S)_0 = 0.10 \)

\( d_b = 4.1 \text{ cm} \)

\( d_b = 5.0 \text{ cm} \)

\( D_{Agg,0} = 1.0 \text{ cm}, (L/S)_0 = 0.31 \)

\( d_b = 5.0 \text{ cm} \)

\( d_b = 4.1 \text{ cm} \)
The local bed shear used for the computation of agglomerate breakage varies significantly with the bubble diameter $d_b$. For instance, if $d_b$ increases from 4.1 cm to 5.0 cm, the bed shear decreases by about 20%. The value of the bubble diameter used is even more critical since the model assumes a constant value of $d_b$ in the bed. Figure 5-35 presents the liquid losses predicted by the model for two values of the model parameter $d_b$.

For the two bubble diameters $d_b$ tested, the statistical analysis shows similar results as in Section 5.1.3.1 and is presented in Appendix N: Statistical analysis of the model results. The relative variations observed with the “representative” agglomerate are conserved, but the amplitude of the variations is modified.

The bubble diameter $d_b$ has a variable impact on predicted liquid losses, depending on the initial agglomerate properties. For small and dry agglomerates, the variation of $d_b$ has no significant impact. The drying mechanism is dominant, and nearly all agglomerates are...
destroyed before reaching the bottom of the Fluid Coker. On the other hand, the impact of the bubble diameter $d_b$ is visible for larger and wetter agglomerates. As these agglomerates are more sensitive to bed shear, larger bubbles lead to more significant liquid losses with the equations used in the model. For example, if $d_b$ increases from 4.1 cm to 5.0 cm, the predicted liquid losses vary with a factor ranging between 2 and 6.

A more detailed discussion of the effect of the bubbles size on the model predictions was developed in Chapter 2 – Section 2.9.1.4 and Section 2.5.1.2.1.

5.4.3 Effect of empirical parameter $M$

Figure 5-36 presents the liquid losses predicted by the model for two values of the model parameter $M$.

Figure 5-36. Fraction of injected liquid reaching the bottom of the Fluid Coker as a function of the empirical parameter $M$

$\rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg,0}} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$, $L_{\text{IN}} = 0 \text{ cm}$
The empirical coefficient M has only a minor impact on predicted liquid losses. The relative variations observed with the “representative” agglomerate are conserved. The statistical analysis shows similar results for the two values of M tested as in Section 5.1.3.1 and is presented in Appendix N: Statistical analysis of the model results.

5.5 Main recommendations based on model results

Based on the model results, the main recommendations to reduce liquid losses to the bottom of a commercial Fluid Coker are the following:

1) Shut down the lowest spray nozzle bank (Bank A) and redistribute its liquid to the other banks. This method reduces predicted liquid losses for all tested initial agglomerate properties and model parameters (see detailed results in Appendix M: Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters) with a reduction of 79 % for the “representative” agglomerate. This result matches the observation from previous work (Wormsbecker, Wiens, et al., 2021). By analyzing the components of the model, the reduction of liquid losses is primarily due to two parameters. First, the larger minimum distance between the agglomerate production zones and the bottom of the Fluid Coker creates a longer drying time, and therefore lower liquid losses. Second, the redistribution creates more favourable bed hydrodynamics between the taper and the lowest injection bank. As a result, the agglomerate can be slowed down and carried up more efficiently by the bubbles and their wakes, moving them from the lower part of the bed (zones with a small bed shear) to the upper parts of the bed (zones with a higher bed shear, promoting breakage). Other redistribution patterns, with another bank than Bank A turned off and redistributed, do not offer significant improvement.

2) The nozzle penetration has a small impact (significantly smaller than the effect of the lateral redistribution), depending on the lateral redistribution profile (see detailed results in Appendix M: Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters). For the base case
(5 Banks), a longer penetration (constant penetration length) has a mostly negligible impact for the investigated range. For the case with 4 banks (A off and redistributed), a longer nozzle penetration (constant penetration length) offers a null to moderate liquid losses reduction for the investigated range. The amplitude of the liquid losses variations changes significantly depending on the agglomerate properties. Therefore, more work is needed to better understand the impact of such nozzle penetration variations on the whole wet-agglomerate population. An alternative nozzle penetration pattern, with different nozzles using different penetrations (0 and 2 cm every other bank), was tested but made liquid losses worse for most configurations tested (the largest and wettest agglomerates were not affected). Nonetheless, future studies should investigate other alternative nozzles penetration patterns (uneven nozzle penetration within a bank, for instance) to identify if some allow for liquid losses reduction.

5.6 Other sources of losses of valuable liquid

This section very briefly discusses how the suggested modifications impact other sources of losses of valuable liquid to the bottom of the Fluid Coker.

5.6.1 Vapour & Liquid carried-under

As mentioned in Chapter 1 – Section 1.2, valuable feedstock losses to the bottom of Fluid Coker can occur through several paths. The most significant one is due to unreacted liquid carried within wet-agglomerates. These losses are the focus of the model developed in this research and have been extensively discussed. Losses can also occur due to liquid and vapour carry-under (in the emulsion phase). The model does not consider these losses. However, they would be impacted by the suggested modifications.

Therefore, preliminary investigative experiments are conducted. The experimental setup and procedure are presented in Chapter 3 – Section 3.6.
5.6.2 Preliminary investigative experiments

In complement to the liquid losses predicted by the model using agglomerate experiments, some preliminary work was conducted to investigate vapour and liquid (in the emulsion) carry under in the experimental unit. A simplified schematic of the experimental setup is presented in Figure 5-37.

The first series of experiments conducted explored the impact of the injection location on the liquid and vapour (in the emulsion) carry-under. Liquid Varso™ was injected in one nozzle (out of 8 per bank) of one bank in successive steps of 5-10 minutes. The flowrate
injected at each step was slowly increased at the end of each step until VarSol™ was detected.

As described in Chapter 3 – Section 3.6, two measurement systems were used. The first system is a chemiresistor, located in the recirculation line. This sensor is very sensitive and able to detect a very small amount of vaporized VarSol™. Since VarSol™ entering the recirculation line is mixed with a significant flow of pure air, any liquid VarSol™ is instantly vaporized. Therefore, the chemiresistor measurements account for both vapour and liquid carry-under (in the emulsion). The second measurement system is a capacitance system, located just above the solids entrance into the recirculation line. This sensor is only sensitive to liquid VarSol™. It is therefore used in conjunction with the chemiresistor to differentiate vapour and liquid carry-under.

Figure 5-38. Evolution of injection flowrate required to detect VarSol™ as a function of the injection bank (5 Banks, \( L_{IN} = 0 \) cm, \( T = 21^\circ C, P = 1 \) atm)

Figure 5-38 shows the evolution of the required liquid VarSol™ flowrate to detect carry-under, with both measurement systems. A higher flowrate means that liquid or vapour are less likely to be carried-under. The expected flowrate to reach local saturation (and therefore the flowrate required to detect VarSol™ if there is no carry-under) is also
plotted. The values were calculated using a simplistic complete Plug Flow Regime (PFR) model for both the bubble and emulsion phase.

The results obtained show that the flowrate required for Varsol™ detection, by either the chemiresistor or the capacitance, is very small compared to the flowrate required to saturate the bed locally at the injection location (for a simplistic complete PFR bed configuration). Future work should include more sophisticated modelling to better understand solids and gas mixing in the bed. In addition, the difference between the results of the chemiresistor and the capacitance indicates that vapour Varsol™ is carried under significantly faster than unreacted liquid Varsol™. This result suggests good solids mixing allowing for efficient vaporization. Finally, when the injection bank is modified, there are only minor variations in the required minimum Varsol™ liquid flowrate to detect some carry-under. Therefore, there is only a limited impact of the distance between the injection bank and the recirculation line, suggesting good solid mixing again.

\( a) \)
The second series of critical experiments investigated the effect of redistributing the lateral injection and modifying the nozzle penetration. The injections were only conducted in Bank B.

For the lateral redistribution experiment, presented in Figure 5-38, the base case (5 Banks) was compared to the case with 4 banks (A off and redistributed). The nozzle penetration used was \( L_{IN} = 0 \) cm.

For the nozzle penetration experiments, presented in Figure 5-39, a penetration of 0 cm (0% of the bed radius) was compared with a penetration of 2.5 cm (21% of the bed radius). The redistribution pattern used was the base case.

**Figure 5-39.** Evolution of injection flowrate required to detect Varso!™, the effect of redistributing the injection profile (\( L_{IN} = 0 \) cm, \( T = 21^\circ C, P = 1 \) atm)

a) Chemiresistor measurement, b) Capacitance measurement
The second series of critical experiments showed a neglectable impact of redistribution and a neglectable impact of nozzle penetration. The emulsion phase seems to be much less impacted by the proposed changes than the wet-agglomerates.
It should be emphasized that these results are still at an exploratory stage, and more experimental and modelling work is required to obtain a better picture of the impact of a lateral injection redistribution on the liquid losses due to vapour and liquid carry-under (in the emulsion).
5.7 References


Chapter 6

6 Impact of baffle(s) addition on liquid losses

Similar to the analysis discussed in Chapter 5, the model developed in Chapter 2 to predict the liquid carry under to the stripper section of a Fluid Coker is used in conjunction with experiments in the pilot-scale cold gas-solid recirculated fluidized bed. The RPT experimental data are processed with the method introduced in Chapter 4. First, it is used to study the impact of adding a single baffle on the liquid losses to the bottom of a commercial Fluid Coker. The objective is to isolate the most favourable configuration. The effect of the presence or absence of flux-tubes and the nozzle penetration are also considered. Then, the analysis is extended to the impact of adding multiple baffles.

The analysis is carried out using a “representative” agglomerate, with initial properties, based on the analysis conducted in Chapter 2. Then, the results obtained are verified by investigating different initial agglomerate properties. In the same way, the main model parameters whose values are not known with precision are analyzed to ascertain the pertinence of the results.

Finally, a short section discusses the other sources of losses of valuable liquid, from liquid trapped in the emulsion phase or gas trapped in the emulsion or bubble phase, and how they could be affected by the proposed changes. The first results of preliminary investigative experiments using Varsol™ are presented.

6.1 Impact of adding a single baffle

In this section, the liquid losses model developed is used to investigate the impact of adding a single baffle. In addition, the mechanisms of the model are analyzed to understand the controlling parameters behind predicted improvements.

6.1.1 Use of baffle in commercial Fluid Coker

In fluidized beds, ring baffles are used to achieve several goals. These goals include: break and redistribute bubbles (Chitnis et al., 2013; Issangya et al., 2013; Kamienski et al., 2013),
modify local bubble distribution (Jahanmiri, 2017; Li et al., 2020), modify solid patterns (Sanchez Careaga, 2013), or improve the liquid spray distribution into the fluidized bed above the baffle (Jahanmiri, 2017; Li et al., 2018).

In a typical commercial Fluid Coker, ring baffles are used to mitigate liquid losses to the stripper and they are usually located near the lowest injection banks (Du et al., 2014). The ring baffles used in commercial Fluid Cokers typically incorporate flux-tubes that allow downward solids flow and upward gas flow to improve baffle performance (Wyatt Jr. et al., 2011; Kamienski et al., 2013; Wyatt Jr. et al., 2013; Wormsbecker, Wiens, et al., 2021)

6.1.2 Model results for a “representative” agglomerate, typical model parameters and standard experimental conditions

In a commercial Fluid Coker, the agglomerate population includes agglomerates with a wide range of size, density, and wetness (see Table 2-1). On the other hand, as explained in Chapter 2, the model developed in this work only considers agglomerates with a single set of initial properties. For the sake of clarity and as introduced in Chapter 5 – Section 5.1.3, as a first step, the model results are presented for an average “representative” agglomerate defined by the following initial properties: $\rho_{\text{Agg},0} = 1100 \text{ kg m}^{-3}$, $D_{\text{Agg},0} = 1 \text{ cm}$, $(L/S)_0 = 0.31$. In Section 6.3, the model results will be verified for other values of initial agglomerates properties.

Similarly, the model uses several parameters to predict the liquid losses at the bottom of a Fluid Coker. Some of these parameters are only known with moderate accuracy, or they can vary within a specific range of values given in the literature. The most critical of these parameters are: the coefficient of re-wetting $C_{\text{Rewet}}$, the empirical coefficient $M$ and the bubble diameter $d_b$. Initially, these parameters will be set as follows: $C_{\text{Rewet}} = 1$, $M = 7$, $d_b = 4.1 \text{ cm}$. Then, in Section 6.4, the model results will be verified for other values of these model parameters.
The experimental conditions used were selected based on the principles described in Chapter 3 – Section 3.1. A simplified schematic of the experimental system is presented in Figure 6-1. The solids recirculation flowrate is $F_S \in [0.42; 0.48] \text{ kg} \cdot \text{s}^{-1}$. The target value was set to $0.45 \text{ kg} \cdot \text{s}^{-1}$. The bed mass is $m_S \in [58; 61.5] \text{ kg}$. The target value was 60 kg. The variability of these two parameters is corrected by making the solids residence times dimensionless thanks to the experimental mean solids residence time $\tau_{\text{EXP}}$ (as defined in Appendix I: Spatial and Temporal scaling of the model). In addition, all the lateral injection nozzles were installed to provide a penetration $L_N = 0 \text{ cm}$ (0% of radius). The impact of different nozzle penetrations (up to 24% of the radius) is presented in Section 6.1.4. The experimental tracer-agglomerates used have the following properties: $D_{\text{Tracer}} \in [0.6; 1.4] \text{ cm}$, $\rho_{\text{Tracer}} \in [900; 1150] \text{ kg} \cdot \text{m}^{-3}$ (see Figure 3-14). The first results presented use a tracer-agglomerate with the following properties: $D_{\text{Tracer}} = 0.9 \text{ cm}$, $\rho_{\text{Tracer}} = 1100 \text{ kg} \cdot \text{m}^{-3}$. As
mentioned earlier, in Section 6.3, the model results will be verified with measurements with other values of tracer-agglomerates properties.

Based on results from Chapter 5, only two lateral redistribution patterns are considered. First, the base case, where the 5 lateral injection banks of the experimental unit are considered (later referred to as the “base case”). Second, the case with 4 lateral injections, where the lowest bank (Bank A) is turned off, and its liquid and atomization gas are evenly redistributed to the remaining nozzle banks.

Finally, the geometry of the baffle used in the experimental unit is described in detail in Chapter 3 – Section 3.1.4. For the experiments with a single baffle in the bed, the baffle is located so that the top of the baffle is located just below the bottom of the lateral injection nozzles of bank B. This position in the bed corresponds approximately to the typical ring baffle position in a commercial Fluid Coker when used to mitigate liquid losses to the stripper (Du et al., 2014). This configuration allowed for the investigation of the impact of maintaining some lateral injection active below the baffle location, via Bank A.

A significant limitation in the design of the experimental baffle is related to the scaling of the flux-tubes. On the one hand, the baffle is scaled down from the design of ring baffles used in a commercial Fluid Cokers, keeping the same open area. However, on the other hand, the tracer-agglomerate has a diameter similar to the size of real wet-agglomerates. Therefore, a tracer-agglomerate cannot easily move through the flux-tubes of the experimental baffle (flux-tubes diameter 0.75-1.5 times larger than the tracer-agglomerates), while real wet-agglomerates can easily go through the large flux-tubes of commercial ring baffles (flux-tubes diameter 15-30 times larger than the largest wet-agglomerates).

The first set of experiments results presented do not include flux-tubes. The impact of the presence or absence of flux-tubes is investigated in Section 6.1.3.
6.1.2.1 Liquid losses predicted by the model

Figure 6-2 presents the liquid losses predicted by the model in a commercial Fluid Coker. These liquid losses correspond to the fraction (on a mass basis) of liquid injected into the reactor that reached the top of the stripper zone by being carried within wet-agglomerates. Thus, these liquid losses do not include any liquid carried down as liquid in the emulsion or as vapour in the emulsion.

![Figure 6-2. Fraction of injected liquid reaching the bottom of the Fluid Coker](image)

\[\rho_{\text{Agg},0} = 1100 \text{ kg·m}^{-3}, \ D_{\text{Agg},0} = 1.0 \text{ cm}, \ (L/S)_0 = 0.31, \ L_{\text{IN}} = 0 \text{ cm}\]

Similar to the results of Chapter 5, a statistical analysis is conducted to verify the statistical significance of the results presented, using a test of hypotheses relative to the difference of means (Spiegel and Stephens, 2017: 277). The procedure involves a test against a two-tailed Student distribution with a 99% confidence interval and is described in Appendix N: Statistical analysis of the model results. The results are presented in Table 6-1.
Table 6-1. Statistical analysis (Differences of means) – Effect of single baffle addition for the “representative” agglomerate

<table>
<thead>
<tr>
<th>L\textsubscript{IN} = 0 cm</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td></td>
<td>No Baffle</td>
<td>I Baffle (No FT)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>6</td>
<td>10</td>
</tr>
<tr>
<td>\textit{t}\textsubscript{STUDENT} (two-tailed, 99%)</td>
<td>2.98</td>
<td>3.05</td>
</tr>
<tr>
<td>Average</td>
<td>9.71 \times 10^{-4}</td>
<td>4.59 \times 10^{-4}</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.15 \times 10^{-4}</td>
<td>8.41 \times 10^{-5}</td>
</tr>
<tr>
<td>\sigma</td>
<td>1.58 \times 10^{-4}</td>
<td>3.61 \times 10^{-5}</td>
</tr>
<tr>
<td></td>
<td>t</td>
<td></td>
</tr>
<tr>
<td>Statistically different?</td>
<td>TRUE</td>
<td>TRUE</td>
</tr>
</tbody>
</table>

The results in Figure 6-2 and Table 6-1 show that for the “representative” agglomerate considered, a significant reduction in liquid losses is achieved with a baffle addition (at Bank B, no flux-tubes). For the base case (5 banks), the liquid losses are reduced by a factor of 0.47. For the 4 banks case (A off and redistributed), the liquid losses reduction has a stronger factor of 0.26. For reference, changing the lateral redistribution profile from the base case to the 4 banks case (A off and redistributed) is associated with a liquid losses reduction factor of 0.21 (-79\%), and adding a baffle reducing losses further by a factor of 0.26 (-74\%), for an overall reduction factor of 0.06 (-94\%).

6.1.2.2 Impact of single baffle addition at Bank B (no flux-tubes) on mechanisms included in the model

This section investigates how adding a single baffle at Bank B (no flux-tubes) impacts the various mechanisms bundled in the model (as presented in Chapter 2). As in Chapter 5, it should be noted that these mechanisms are interconnected. For instance, a longer drying time will promote more breakage. Again, the objective of presenting these mechanisms one after the other is not to imply that they are independent but to identify important mechanisms.
6.1.2.2.1 Wetting at formation

As described in detail in Chapter 5 – Section 5.1.3.2.1, the model calculates how much of the liquid injected is trapped in the agglomerate formed using an empirical correlation (Chapter 2 – Section 2.3.2.2) varying with two parameters: the initial agglomerate diameter $D_{\text{Agg,0}}$ and the cross-sectional superficial velocity below the formation zone $u_{\text{G,Below}}$ (Li, 2016). Therefore, adding a baffle will have a small impact on the wetting. Everything else being equal, the cross-sectional superficial velocity below the formation zone $u_{\text{G,Below}}$ would be slightly increased, leading to a reduction of the amount of liquid injected captured by agglomerates on the order of 3-5 %.

6.1.2.2 Production-to-stripper time & drying

Once an agglomerate is formed and leaves the formation zone (at the tip of a lateral injection jet), it starts to dry. As presented in Chapter 2 – Section 2.4, the drying model used in the model is the Shrinking Core Model. Therefore, for two identical agglomerates, the drying level is only controlled by the production-to-stripper time (as defined in Chapter 2 – Section 2.4.1.1).

Figure 6-3 compares the cumulative distribution of the production-to-stripper time of the base case with the cumulative distribution of the production-to-stripper time after a single baffle addition (at Bank B, no flux-tubes).

Figure 6-3 shows that a single baffle addition (at Bank B, no flux-tubes) significantly lengthens the production-to-stripper time. The increase is more pronounced for the 4 banks (A off and redistributed) lateral distribution pattern than for the base case. The dimensionless $t/\tau_{\text{EXP}}$ required for the 50% of the agglomerates to travel from the production zone to the stripper zone increases from 0.134 to 0.221 (Factor of 1.65) for the base case and from 0.155 to 0.390 (Factor of 2.52) for the 4 banks case (A off and redistributed). For reference, changing the lateral redistribution profile from the base case to the 4 banks case (A off and redistributed) is associated with an increase in the dimensionless $t/\tau_{\text{EXP}}$ required for the 50% of the agglomerates to travel from the production zone to the stripper zone from 0.134 to 0.221 (Factor of 1.57).
Figure 6-3. Cumulative distribution of the production-to-stripper time of tracer-agglomerates ($D_{Tracer} = 0.9 \text{ cm}, \rho_{Tracer} = 1100 \text{ kg} \cdot \text{m}^{-3}, L_{IN} = 0 \text{ cm}, \text{ with replicates}$)

a) 5 banks – Baffle vs. Baffle at B (No Flux-tubes),

b) 4 banks (A off) – Baffle vs. Baffle at B (No Flux-tubes)

The dimensionless $t/\tau_{EXP}$ required for the 25% fastest agglomerates to travel from the production zone to the stripper zone increases from 0.048 to 0.083 (Factor of 1.73) for the
base case and from 0.080 to 0.190 (Factor of 2.38) for the 4 banks case (A off and redistributed). The addition of a baffle slows down all the agglomerates, from the fastest to the slowest, similarly. It should be noted that, on the other hand, the lateral distribution redistribution observed in Chapter 5 – Section 5.1.3.2.2 had a more pronounced effect on the fastest travelling agglomerates. This result suggests that the single baffle addition creates a global shift of the production-to-stripper time distribution, for all active banks.

This global shift is verified in Figure 6-4, which compares the contribution of each bank between the no baffle and the single baffle at Bank B (no flux-tubes) case (for the base case lateral distribution pattern). For each lateral injection bank, the production-to-stripper time distribution is shifted towards longer values.

**Figure 6-4.** Cumulative distribution of the production-to-stripper time of tracer-agglomerates, comparison of the contribution of each lateral injection bank when a single baffle (no flux-tubes) is added at Bank B

\(D_{\text{Tracer}} = 0.9 \text{ cm}, \ \rho_{\text{Tracer}} = 1100 \text{ kg}\cdot\text{m}^{-3}, \ L_{\text{IN}} = 0 \text{ cm}\)

Finally, it should be noted that the moderate variability between replicates observed in Figure 6-3 cannot be attributed to one specific lateral injection bank. These variations could come from variations in the recirculated solids flowrate (which is not totally constant
during a run) or inaccuracies in the solids mass, $m_S$, or the recirculated solids flowrate, $F_S$, measurements (which would lead to inaccuracies in experimental residence time, $\tau_{\text{EXP}}$).

These results indicate a modification of the bed hydrodynamics, which would significantly slow down wet-agglomerates moving downward. Figure 6-5 gives some insight into the reason behind these modifications of the agglomerate hydrodynamics.

Figure 6-5 clearly shows that adding a single baffle at Banks B (no flux-tubes) creates a choke-point for the agglomerates: a strong core-annulus structure is established, and the downward motion of the annulus is blocked by the baffle. For agglomerates located above the baffle, this baffle addition limits downward crossings of the wet-agglomerates. On the
other hand, the agglomerates located below the baffle have an increased chance of being carried back up by the bubbles and their wakes. In addition, the results for Bank A in Figure 6-4, with a longer production-to-stripper time distribution even for the Bank A (located below the baffle), suggest the establishment of a recirculation cell between the bottom of the taper and the baffle.

Figure 6-6. Average liquid content of agglomerates at the bottom of the Fluid Coker

\( \rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, \ D_{\text{Agg},0} = 1.0 \text{ cm}, \ (L/S)_{0} = 0.31, \ L_{\text{IN}} = 0 \text{ cm} \)

Figure 6-6 presents the average liquid content carried by agglomerates to the stripper zone. This parameter decreases with: 1) longer drying (longer production-to-stripper time), 2) more agglomerate destruction, 3) less re-wetting. These effects can be complementary or opposed. The results presented in Figure 6-6 follow the same trend as the results given by the study of the production-to-stripper time (Figure 6-3): adding a single baffle (at Bank B, no flux-tubes) reduces the liquid losses. It suggests that the three mechanisms aforementioned are complementary, or at least that the drying (controlled by the production-to-stripper time) is not neglectable. The following sections verify the relative impact of the agglomerate destruction and re-wetting.
It should also be noted that despite the significant shift in the production-to-stripper time distribution, the average liquid content carried by agglomerates to the stripper zone does only decrease moderately. This result suggests a strong influence of the re-wetting mechanisms. The impact of re-wetting will be discussed in Section 6.1.2.2.4.

6.1.2.2.3 Bed shear & breakage

In addition to drying, the other mechanism limiting the fraction of agglomerate formed reaching the bottom of the Fluid Coker is the breakage due to bed shear (see Chapter 2 – Section 2.5). This mechanism is connected to three parameters: agglomerate saturation, local bed shear, and agglomerate trajectories in the bed. It should be noted that the agglomerate saturation is a direct function of the agglomerate wetness. Therefore, drying impacts the breakage mechanism: dryer agglomerates break more easily. This relation is important as it highlights how interconnected the mechanisms of the model are.

The local bed hydrodynamics controls the local bed shear. Therefore, it varies with the injection profile, as shown in Chapter 5 - Figure 5-9.

As presented in Chapter 2 – Section 2.5.1.2, bed shear increases with the cross-sectional superficial gas velocity. The bed shear increases with the bed height, with clear separation at the location of the injection banks. It will also increase with cross-sectional area restrictions, such as those created by inserting a ring baffle in the bed.

On the other hand, bed shear is also a function of the local bed voidage. As described in Chapter 2 – Section 2.5.1.2.2, this local voidage was obtained through a combination of pressure-based and radiation-based measurements then fitted using an assumption of unimodal core annulus assumption (see Appendix K: Measurement of the bed voidage profile in the experimental unit). The comparison of measured bed voidage acquired on the experimental unit with and without a baffle showed two main differences: 1) the average bed voidage at a given height is lower in the zone above the baffle, 2) the radial voidage at a given height is flatter above the baffle. These two effects combined lead to a reduction of the bed shear above a baffle.
Therefore, when a baffle is added, the local increase of cross-sectional superficial gas velocity and the bed voidage have opposite effects. Figure 6-7 presents the bed shear as calculated in the model for both the no baffle and the single baffle (at Bank B, no flux-tubes) cases. Two configurations are considered: the base case (5 banks) and the case with 4 banks (A off and redistributed). The local increase of bed shear between the baffle tip is visible while restricted to a small volume (around 2-3 cm high). The overall lower bed shear value and the flatter bed shear profile above the baffle are also visible.

Figure 6-7. Evolution of local bed shear as calculated in the model, as a function of the injection profile used (d_b = 4.1 cm)

a) 5 banks & No baffle, b) 5 banks & Baffle at Bank B (No flux-tubes)

Figure 6-8 presents the fraction of agglomerates produced reaching the stripper with liquid and shows that adding a single baffle (at Bank B, no flux-tubes) leads to significant improvements compared to the base case. The number of agglomerates reaching the bottom of the Fluid Coker is reduced by a factor of 0.83 (- 17 %) for the base case (5 banks) and by a factor of 0.28 (- 72 %) for the case with 4 banks (A off and redistributed). For reference, changing the lateral redistribution profile from the base case to the 4 banks case (A off and redistributed) is associated with a reduction of the fraction of agglomerates produced reaching the stripper with liquid by a factor of 0.53 (- 47 %).
Figure 6-8. Fraction of agglomerates reaching the bottom of the Fluid Coker with liquid

\((\rho_{\text{Agg},0} = 1100 \, \text{kg} \cdot \text{m}^{-3}, D_{\text{Agg},0} = 1.0 \, \text{cm}, (L/S)_0 = 0.31, L_{\text{IN}} = 0 \, \text{cm})\)

This decrease in the fraction of agglomerates produced reaching the stripper with liquid can be partially explained by the increased drying due to longer production-to-stripper time. Regarding the effect of the modification of the bed shear, the analysis is less straightforward as there are two competing effects. First, most of the bed volume located above the baffle presents a lower average bed shear and a flatter bed shear radial profile: agglomerates spending time in this zone encounter less bed shear, especially near the core of the unit. On the other hand, there is a forced crossing of a zone with moderate bed shear at the baffle tip: when previously an agglomerate could move downward by staying in the low bed shear annulus, it is now forced to go at least once in a zone of moderate shear. This forced crossing of a zone of moderate bed shear might be especially critical for agglomerates having spent a significant amount of time in the upper zone (without being re-wetted) and, therefore, quite dry and weak.

As mentioned in Chapter 5 – Section 5.1.3.2.3, the average bed shear encountered by the wet-agglomerate is intrinsically correlated with the fraction of agglomerates produced reaching the stripper with liquid. Therefore, it does not bring additional information about the mechanisms causing more agglomerates breakage when a single baffle is added.
6.1.2.2.4 Re-wetting

As presented in Chapter 2 – Section 2.3.3 & Section 2.8 and Chapter 5 – Section 5.1.3.2.4, re-wetting plays a significant role in determining the amount of liquid carried under to the bottom of the Fluid Coker. Re-wetting can happen more than once, and the lowest location of re-wetting is especially critical as it almost behaves as a “pseudo-formation” zone.

Figure 6-9 presents the fraction of agglomerates re-wetted and the location of the last agglomerate re-wetting bank for the no baffle and single baffle (at Bank B, no flux-tubes) cases. Two configurations are considered: the base case (5 banks) and the case with 4 banks (A off and redistributed). It should be noted that similar to previous results, the overwhelming majority of agglomerates carrying liquid to the bottom of the Fluid Coker have been re-wetted (> 80%) for any configuration.

Figure 6-9 shows that adding a single baffle (at Bank B, no flux tubes) does not significantly modify the fraction of agglomerates re-wetted and the location of the last agglomerate re-wetting bank.

\[ a \]
6.1.2.2.5 Summary of the observations

The analysis of the impact of adding a single baffle (at Bank B, no flux tubes) on the mechanisms of the model (for two redistribution patterns, the base case, with 5 banks, and the case with 4 banks, with Bank A off and redistributed) can be summarized as follows.

- Wetting at formation: minor impact due to the local increase of $u_{G,\text{Below}}$ after a baffle is added (3-5 % reduction)
- Production-to-stripper time & drying: adding a single baffle (at Bank B, no flux-tubes) leads to longer production-to-stripper times (and therefore enhanced drying). The baffle modifies the bed hydrodynamics and creates a choke-point for the agglomerate. It prevents the agglomerates from moving down by enhancing the agglomerates/bubbles interaction in a restricted volume.
- Bed shear & breakage: moderate reduction in the number of agglomerates reaching the stripper when a single baffle (at Bank B, no flux-tubes) is added. The
choke-point forces agglomerates to go through a zone of moderate bed shear at least once. On the other hand, the baffle addition moderately reduces the intensity of the bed shear in the upper part of the bed.

- Re-wetting: this mechanism is not modified significantly.

6.1.3 Impact of flux-tubes

As mentioned in Chapter 3 – Section 3.1.4.3 and in Section 6.1.1, commercial ring baffles typically include flux-tubes (or “downcomers”) that allow downward solids flow and upward gas flow, improving baffle performance (Wyatt Jr. et al., 2011).

**Figure 6-10.** Fraction of injected liquid reaching the bottom of the Fluid Coker.

Impact of nozzle penetration (same distance for all nozzles)

\[ \rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}, \quad D_{Agg,0} = 1.0 \text{ cm}, \quad (L/S)_0 = 0.31, \quad L_{IN} = 0 \text{ cm} \]

This section investigates the impact of the presence or absence of flux-tubes on the liquid losses predicted by the model. Two significant limitations need to be considered. As mentioned in Section 6.1.2, the first one is due to the different methods of scaling used respectively for the baffle and the tracer-agglomerate. The baffle is scaled down while the agglomerate is not. Therefore, the experimental tracer agglomerate cannot move down through the flux-tubes of the experimental baffle, while real wet-agglomerates can easily
go through the large flux-tubes of commercial ring baffles. The second limitation is related to the calculation of the bed shear at the baffle tip. The model does not consider the extra open area due to the presence of flux tubes, overestimating the gas flow in the center area of the baffle and therefore overestimating the local bed shear (overestimation of shear). This second limitation can be modified relatively easily in the model equations for future studies.

Similar to the results of Section 6.1.2.1, a statistical analysis is conducted to verify the significance of the results. The procedure is described in Appendix N: Statistical analysis of the model results. The results are presented in Table 6-2.

With the limitations mentioned above, Figure 6-10 and Table 6-2 show almost no impact on the predicted liquid losses relative to the presence or absence of the flux-tubes on the baffle.

**Table 6-2.** Statistical analysis (Differences of means) – Effect of presence or absence of flux-tubes on a single baffle, for the “representative” agglomerate

<table>
<thead>
<tr>
<th>L_{IN} = 0 cm</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>l Baffle (No FT)</td>
<td>1 Baffle (FT)</td>
<td>1 Baffle (No FT)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>10</td>
<td>4</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}}$ (two-tailed, 99%)</td>
<td>3.05</td>
<td>3.17</td>
</tr>
<tr>
<td>Average</td>
<td>$4.59 \times 10^{-4}$</td>
<td>$4.34 \times 10^{-4}$</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>$8.41 \times 10^{-5}$</td>
<td>$1.37 \times 10^{-4}$</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>$1.10 \times 10^{-4}$</td>
<td>$6.42 \times 10^{-6}$</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

### 6.1.4 Impact of nozzle penetration

As mentioned in Chapter 5 – Section 5.1, in a commercial Fluid Coker, the nozzle penetration varies during a run (see Chapter 3 – Table 3-1). This variation is due to a build-up of coke on the walls, which reduces the nozzle penetration over time by reducing
the internal diameter of the Fluid Coker reactor. At the beginning of a run, the typical nozzle penetration is around 16% of the radius. In contrast, just before a maintenance shutdown, nozzles have an almost null effective penetration (0% of the radius) (Wormsbecker, McMillan, et al., 2021).

This section investigates the effect of the nozzle penetration on the predicted liquid losses, performing a similar analysis as the one conducted in Chapter 5 – Section 5.2. Only the configuration with the same penetration for all nozzles is tested (see Chapter 5 – Section 5.2.1), with penetration values ranging from 0 cm (0% of bed radius) to 2 cm (16% of bed radius).

The liquid losses predictions with various penetrations (same penetration value for all nozzles) are presented in Figure 6-11.

![Figure 6-11](image)

**Figure 6-11.** Fraction of injected liquid reaching the bottom of the Fluid Coker.

Impact of nozzle penetration (same distance for all nozzles)

\[
\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, \quad D_{\text{Agg},0} = 1.0 \text{ cm}, \quad (L/S)_0 = 0.31
\]

Similar to the results of Section 6.1.2.1, a statistical analysis is conducted to verify the significance of the results. The procedure is described in Appendix N: Statistical analysis of the model results. The results are presented in Table 6-3 and Table 6-4.
Figure 6-11, Table 6-3 and Table 6-4 show that no significant variations are visible when the nozzle penetration varies while a baffle is used. The variations are even less significant than for the case without a baffle (see Chapter 5).

**Table 6-3.** Statistical analysis (Differences of means) – Effect of nozzle penetration with a single baffle (no flux-tubes), for the “representative” agglomerate

<table>
<thead>
<tr>
<th>1 Baffle (No FT)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td></td>
<td>$L_{IN} = 0$ cm</td>
<td>$L_{IN} = 1$ cm</td>
</tr>
<tr>
<td>Number of samples</td>
<td>10</td>
<td>2</td>
</tr>
<tr>
<td>$t_{STUDENT}$ (two-tailed, 99%)</td>
<td>3.17</td>
<td>3.05</td>
</tr>
<tr>
<td>Average</td>
<td>4.59 $10^{-4}$</td>
<td>4.84 $10^{-4}$</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>8.41 $10^{-5}$</td>
<td>4.98 $10^{-5}$</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>8.70 $10^{-5}$</td>
<td>1.27 $10^{-4}$</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

**Table 6-4.** Statistical analysis (Differences of means) – Effect of nozzle penetration with a single baffle (flux-tubes), for the “representative” agglomerate

<table>
<thead>
<tr>
<th>1 Baffle (No FT)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td></td>
<td>$L_{IN} = 0$ cm</td>
<td>$L_{IN} = 1$ cm</td>
</tr>
<tr>
<td>Number of samples</td>
<td>4</td>
<td>2</td>
</tr>
<tr>
<td>$t_{STUDENT}$ (two-tailed, 99%)</td>
<td>4.60</td>
<td>4.60</td>
</tr>
<tr>
<td>Average</td>
<td>4.34 $10^{-4}$</td>
<td>5.22 $10^{-4}$</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.37 $10^{-4}$</td>
<td>2.64 $10^{-4}$</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>2.31 $10^{-4}$</td>
<td>1.40 $10^{-4}$</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>
6.2 Impact of adding several baffles

In this section, the liquid losses model developed is used to investigate the impact of adding more than one baffle simultaneously. In addition, the mechanisms of the model are analyzed to understand the controlling parameters behind predicted improvements. Simplified schematics of the experimental setup are presented in Figure 6-12. It should be noted that while ring baffles are already used in commercial Fluid Cokers, few configurations include more than one baffle (Kamienski et al., 2013).

Similar to the single baffle addition analysis, this analysis of the effect of adding multiple baffles is carried out using a “representative” agglomerate and typical model parameters. In addition, all results will be verified for some other initial agglomerate properties (Section 6.3) and other model parameters (see Section 6.4). It should be noted that fewer cases were studied compared to the analysis carried for the single baffle addition.

Figure 6-12. Simplified schematics of the experimental setup
6.2.1 Liquid losses predicted by the model

The liquid losses prediction while using no baffle, one baffle (at Bank B, with & without flux-tubes), and more than one baffle (at Banks B, C, D & E, no flux-tubes) are presented in Figure 6-13. A moderate extra reduction in liquid losses is achieved from going from one to four baffles, with a more pronounced variation for the 4 banks case.

Figure 6-13. Fraction of injected liquid reaching the bottom of the Fluid Coker

\( \rho_{\text{Agg,0}} = 1100 \, \text{kg} \cdot \text{m}^{-3}, \, D_{\text{Agg,0}} = 1.0 \, \text{cm}, \, (L/S)_0 = 0.31, \, L_{\text{IN}} = 0 \, \text{cm} \)

Similar to the results of Section 6.1.2.1, a statistical analysis is conducted to verify the significance of the results. The procedure is described in Appendix N: Statistical analysis of the model results. The results are presented in Table 6-5.

Figure 6-13 and Table 6-5 show that the liquid losses reduction does not significantly vary for the base case (5 banks) when more than one baffle is added (less than 1 % of relative variation). This result matches the CFD analysis performed by Kamienski et al. (2013) and presented in Figure 6-14. Their results did not show any significant improvements when using more than one baffle.
Table 6-5. Statistical analysis (Differences of means) – Effect of addition of one baffle vs. more than one baffle, for the “representative” agglomerate

<table>
<thead>
<tr>
<th>L_{IN} = 0 cm</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>1 Baffle (at B - No FT)</td>
<td>10</td>
<td>4</td>
</tr>
<tr>
<td>4 Baffles (at B, C, D, E - No FT)</td>
<td>3</td>
<td>3.11</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th></th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Number of samples</strong></td>
<td>10</td>
<td>3</td>
</tr>
<tr>
<td><strong>t_{STUDENT}</strong></td>
<td>3.11</td>
<td>3.17</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>4.59 x 10^4</td>
<td>4.18 x 10^4</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>8.41 x 10^5</td>
<td>7.17 x 10^7</td>
</tr>
<tr>
<td><strong>σ</strong></td>
<td>8.02 x 10^3</td>
<td>4.74 x 10^6</td>
</tr>
<tr>
<td>**</td>
<td>t</td>
<td>**</td>
</tr>
<tr>
<td><strong>Statistically different?</strong></td>
<td>FALSE</td>
<td>TRUE</td>
</tr>
</tbody>
</table>

Figure 6-14. CFD simulation of the impact of staging baffles for stripper fouling mitigation (Kamienski et al., 2013)
On the other hand, Figure 6-13 and Table 6-5 show that, for the case of 4 banks (A off and redistributed), adding more than one baffle leads to a significant extra-reduction in liquid losses. The factor of reduction relative to the base case goes from 0.26 with a single baffle to 0.18 with multiple baffles. This decrease represents a relative variation of around 30%. For reference, changing the lateral redistribution profile from the base case to the 4 banks case (A off and redistributed) is associated with a liquid losses reduction factor of 0.21.

6.2.2 Impact of multiple baffles addition (no flux-tubes) on mechanisms included in the model

This section investigates how adding multiple baffles (no flux-tubes) impacts the various mechanisms bundled in the model (as presented in Chapter 2). Again, it should be noted that these mechanisms are interconnected.

6.2.2.1 Wetting at formation

Similar to what was described in Section 6.1.2.2.1, adding a baffle will have a small to moderate impact on the wetting. For each individual baffle, everything else being equal, the cross-sectional superficial velocity below the formation zone $u_{G,Below}$ would be slightly increased, leading to a reduction of the amount of liquid injected captured by agglomerates on the order of 3-12%.

The baffles located at higher positions, where the cross-sectional superficial velocity is higher than at lower positions, are the ones with the most significant decrease of injected liquid captured by initial agglomerates.

6.2.2.2 Production-to-stripper time & drying

Figure 6-15 presents the comparison of the cumulative distribution of the production-to-stripper time when no baffle, a single baffle (at Bank B, no flux-tubes), and several baffles (at Bank B, C, D & E, no flux-tubes) are used. Two lateral redistribution patterns are presented, the base case (5 banks) and the case with 4 banks (A off and redistributed).
Figure 6-15. Cumulative distribution of the production-to-stripper time of tracer-agglomerates ($D_{\text{Tracer}} = 0.9 \text{ cm}, \rho_{\text{Tracer}} = 1100 \text{ kg} \cdot \text{m}^{-3}, L_{\text{IN}} = 0 \text{ cm}, \text{ with replicates})

a) 5 banks – Baffle vs. Baffle at B (No Flux-tubes) vs. 4 Baffles at Bank B, C, D & E (No flux-tubes),

b) 4 banks (A off) – Baffle vs. Baffle at B (No Flux-tubes) vs. 4 Baffles at Bank B, C, D & E (No flux-tubes)
Figure 6-15 shows that more than one baffle addition (at Bank B, C, D & E, no flux-tubes) lengthen even more the production-to-stripper time than just adding a single baffle (at Bank B, no flux-tubes). Once again, the increase is more pronounced for the 4 banks (A off and redistributed) lateral distribution pattern than for the base case. The dimensionless $t/\tau_{\text{EXP}}$ required for 50% of the agglomerates to travel from the production zone to the stripper zone increases from 0.134 to 0.221 (Factor of 1.65) for the base case and from 0.155 to 0.390 (Factor of 2.52) for the 4 banks case (A off and redistributed).

It should be noted that similarly to the single baffle addition, the shift of the production-to-stripper time distribution is global, for all active banks.

![Graph showing liquid content carried by agglomerates to the bottom of the Fluid Coker](image)

**Figure 6-16.** Average liquid content carried by agglomerates to the bottom of the Fluid Coker

$(\rho_{\text{Agg.0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg.0}} = 1.0 \text{ cm}, (L/S)_0 = 0.31, L_{\text{IN}} = 0 \text{ cm})$

Figure 6-16 presents the average liquid content carried by agglomerates to the stripper zone. This parameter decreases with: 1) longer drying (longer production-to-stripper time), 2) more agglomerate destruction, 3) less re-wetting. These effects can be complementary or opposed. The results presented in Figure 6-16 follow the same trend as the results given
by the study of the production-to-stripper time (Figure 6-3): adding more than one baffle (at Bank B, C, D & E, no flux-tubes) is better than adding a single baffle (at Bank B, no flux-tubes) reduces the liquid losses. It suggests that the three mechanisms aforementioned are complementary, or at least that the drying (controlled by the production-to-stripper time) is dominant.

Nonetheless, the significant difference in the amplitude of decrease between the base case (5 banks) and the case with 4 banks (Bank A off and redistributed) should be noted. The average liquid content carried by agglomerates to the stripper zone only decreases slightly for the base case, despite the significant shift in the production-to-stripper time distribution. This result suggests a strong influence of the re-wetting mechanisms. On the other hand, the case with 4 banks (Bank A off and redistributed) shows a substantial drop in the average liquid content carried by agglomerates to the stripper zone when more than one baffle is added. The impact of re-wetting will be discussed in Section 6.2.2.4.

6.2.2.3 Bed shear & breakage

Figure 6-17 presents the bed shear as calculated in the model for both the no baffle, the single baffle (at Bank B, no flux-tubes) and the multi-baffle (at Bank B, C, D & E, no flux-tubes) cases. Two configurations are considered: the base case (5 banks) and the case with 4 banks (A off and redistributed).

Figure 6-17 shows much more pronounced bed shear variations due to the multiple baffles addition. These stronger variations are primarily due to the installation in zones with a higher cross-sectional superficial velocity. Therefore, the baffles, especially the top three, act as “filters”, destroying most agglomerates going through them. As a result, only the strongest agglomerates formed above these baffles can go through the choke-points and reach the stripper with liquid. On the other hand, weak agglomerates formed below the lowest baffle can still easily reach the stripper and contribute significantly to liquid losses.
Figure 6-17. Evolution of local bed shear as calculated in the model, as a function of the injection profile used ($d_b = 4.1$ cm)

a) 5 banks & No baffle, b) 5 banks & Baffle at Bank B (No flux-tubes),

c) 5 banks & 4 Baffles at Bank B, C, D & E (No flux-tubes)

d) 4 banks (A off) & No baffle, e) 4 banks (A off) & Baffle at Bank B (No flux-tubes),

f) 5 banks & 4 Baffles at Bank B, C, D & E (No flux-tubes)
Figure 6-18 presents the fraction of agglomerates produced reaching the stripper with liquid.

**Figure 6-18.** Fraction of agglomerates reaching the bottom of the Fluid Coker with liquid

\( \rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, \ D_{\text{Agg,0}} = 1.0 \text{ cm}, \ (L/S)_0 = 0.31, \ L_{\text{IN}} = 0 \text{ cm} \)

Figure 6-18 shows two different patterns when more than one baffle is used. For the base case, when the lateral redistribution pattern includes 5 banks (therefore, with one bank located below the lowest baffle), only a minimal decrease in the fraction of agglomerates produced reaching the stripper with liquid is observed. On the other hand, when the lateral redistribution pattern uses 4 banks (A off and redistributed), there is a significant extra-reduction of the fraction of agglomerates produced reaching the stripper with liquid observed. This result highlights the importance of the lowest bank (especially if located below the lowest baffle) in liquid losses.
6.2.2.4 Re-wetting

(a)

Figure 6-19. Fraction of agglomerates re-wetted & location of last agglomerate re-wetting bank
\( \rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31, L_{\text{IN}} = 0 \text{ cm} \)

a) 5 Banks, b) 4 Banks (A off)

Figure 6-19 presents the fraction of agglomerates re-wetted and the location of the last agglomerate re-wetting bank for the first nozzle configuration, with various baffle configurations.

Figure 6-19 shows a substantial decrease in the overall re-wetting when more than one baffle is used. This decrease indicates that most agglomerates cannot move from one bank to another without being either dried or destroyed. The “relay” effect induced by re-wetting becomes least efficient. Most of the agglomerates reaching the bottom of the Fluid Coker reactor are produced in the lowest banks: higher banks become a null or almost null contributor to liquid losses.

6.2.2.5 Summary of the observations

The analysis of the impact of adding a single baffle (at Bank B, no flux tubes) on the mechanisms of the model (for two redistribution patterns, the base case, with 5 banks, and the case with 4 banks, with Bank A off and redistributed) can be summarized as follows.

- Wetting at formation: moderate reduction of liquid captured by initial agglomerates due to the local increase of \( u_{G,\text{Below}} \) after a baffle is added (3-12 % reduction, depending on the height of formation)

- Production-to-stripper time & drying: adding multiple baffles lengthens even more the production-to-stripper time (and drying time). This longer production-to-stripper time is due to a reinforced choke-point effect.

- Bed shear & breakage: extra-reduction in the number of agglomerates reaching the stripper with liquid ranging from small to strong when more than one baffles are used. The state of the lowest bank (Bank A, located below the lowest baffle) is critical: with Bank A active, the extra-reduction in the number of agglomerates reaching the stripper with liquid is characterized by a factor of 0.96 (- 4%), but with A inactive, the extra-reduction in the number of agglomerates reaching the stripper with liquid is characterized by a factor of 0.09 (- 91%)
- Re-wetting: its overall impact is significantly reduced when more than one baffle is used (-50% of agglomerates re-wetted). The impact of higher banks on liquid losses becomes neglectable. The importance of Bank A as a location of last re-wetting for the base case should be noted (50-70% of the last re-wetting).

Therefore, the difference between results obtained with the base case (5 banks) and the case with 4 banks (A off and redistributed) can be explained by a higher agglomerate survival rate for the base case. This higher agglomerate survival rate can be explained due to the important contribution of Bank A, located below the lowest baffle.

### 6.3 Effect of initial agglomerate properties

Similar to Chapter 5 – Section 5.3, this section investigates the possible impact of agglomerate properties on the results predicted by the model.

Figure 6-20 presents the liquid losses predicted by the model for two agglomerates (and tracer-agglomerates) of varying initial densities. Figure 6-21 shows the liquid losses predicted by the model for two agglomerates (and tracer-agglomerates) of varying initial densities. More configurations are also presented in Appendix M: Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters. Similar to the results of Section 6.1.2.1, a statistical analysis is conducted to verify the significance of the results. The procedure and results are described in Appendix N: Statistical analysis of the model results.

In Figure 6-20, the relative variation observed with the “representative” agglomerate is the same for any density tested. Similarly to what was observed in Chapter 5 - Section 5.3.1.2, lighter agglomerates carry less liquid to the stripper. This is due to the combination of longer formation-to-stripper time, lower required time for drying \( t_C \) (they dry more quickly) and lower simplified breakage parameter \( B_{Agg,0} \) (they break more easily).
Figure 6-20. Fraction of injected liquid reaching the bottom of the Fluid Coker

\( D_{\text{Agg},0} = 1.0 \text{ cm}, \ (L/S)_0 = 0.31, \ L_{\text{IN}} = 0 \text{ cm} \)

a) 5 Banks, b) 4 Banks (A off)
Figure 6-21. Fraction of injected liquid reaching the bottom of the Fluid Coker

\( \rho_{\text{Agg.0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, \ L_{\text{IN}} = 0 \text{ cm} \)

a) 5 Banks, b) 4 Banks (A off)
In Figure 6-21, the relative variations observed with the “representative” agglomerate are similar for any combinations of diameters and initial liquid-to-solid ratios tested. It should be noted that for larger and wetter agglomerates, the amplitude of the variation is minimal and becomes not significant. This is because most of these agglomerates are already destroyed in the base case. Similarly to what was observed in Chapter 5 – Section 5.3.1.3 and Section 5.3.1.4, the larger and wetter agglomerates are not the biggest contributors to the liquid losses due to their lower resistance to breakage due to bed shear.

6.4 Effect of important model parameters

Similar to Chapter 5 – Section 5.4, this section discusses the sensitivity of the liquid losses predictions to the variations of significant model parameters: the coefficient of re-wetting $C_{\text{Rewet}}$, the bubble diameter $d_b$, and the empirical parameter $M$.

The detailed results, as well as the comparison of more configurations, are presented in Appendix M: Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters. Similar to the results of Section 6.1.2.1, a statistical analysis is conducted to verify the significance of the results. The procedure is described in Appendix N: Statistical analysis of the model results.

The results show that varying the main model parameters (for the value tested) gives significant variations similar to the “representative” agglomerate.

6.5 Main recommendations based on model results

Based on the model results, the main recommendations to reduce liquid losses to the bottom of a commercial Fluid Coker are the following.

1) Add a baffle at Bank B (with or without flux-tubes). This modification allows a significant reduction in the predicted liquid losses for most tested initial agglomerate properties and model parameters (when no significant reduction is visible, the effect is null). The detailed results are presented in Appendix M: Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters. This type of internal is already in use at similar
locations in some industrial units (Wyatt Jr. et al., 2011; Kamienski et al., 2013; Wyatt Jr. et al., 2013; Du et al., 2014). By analyzing the components of the model, the reduction of liquid losses due to the addition of a single baffle is primarily due to two parameters. First, the production-to-stripper time is extended due to the apparition of choke-points (located at the baffle tip), slowing the agglomerates moving down and carrying them back up. Second, a local increase of bed shear at these choke-points enhancing agglomerates destructions. The amplitude of the improvement varies with the lateral redistribution profile used. For the base case (5 banks), a factor of reduction of 0.47 is achieved (53% reduction compared to the same configuration without a baffle). For the case with 4 Banks (A off and redistributed), a factor of reduction of 0.26 is achieved (74% reduction compared to the same configuration without a baffle). It should be noted that both remain smaller than the liquid losses reduction achieved by shutting Bank A and redistribution (factor of reduction of 0.21). Both the nozzle penetration and the presence or absence of flux-tubes have a negligible impact on the predicted liquid losses when a baffle is used.

2) The use of a single baffle combined with a favourable lateral redistribution of the injection profile has a cumulative effect: combining the lateral redistribution profile with 4 Banks (A off and redistributed) with a baffle (with or without flux-tubes) located at bank B gives a greater reduction in liquid losses than each modification individually. The combined reduction of liquid losses can reach 95% compared to the base case.

3) Adding more than one baffle allows for a significant extra-reduction of liquid losses (extra 30% reduction), but only if no active lateral injection bank is located below the lowest baffle. If an active lateral injection bank located below the lowest baffle is still active, no significant extra-reduction of predicted liquid losses is achieved.
6.6 Other sources of losses of valuable liquid

In complement to the liquid losses predicted by the model using agglomerates experiments, some preliminary work was conducted to investigate vapour and liquid (in the emulsion) carry-under in the experimental unit. The experimental setup and procedure are presented in Chapter 3 – Section 3.6 & Chapter 5 – Section 5.6.2.

\[ a) \]

\[ b) \]
A series of critical experiments were conducted to investigate the baffle effect on the vapour and liquid carry-under (in the emulsion). The experiments were conducted with the base case (5 banks lateral distribution, $L_{IN} = 0$ cm) and a single baffle (with flux-tubes) located at Bank B. The injections were only conducted in Bank A (below the baffle) and Bank B (above the baffle).

Figure 6-22 presents the evolution of the required liquid Varso® flowrate to detect carry-under, with both the chemiresistor and the capacitance measurement systems. The results show an apparent effect of the baffle addition. When liquid Varso® is injected above the baffle (Bank B, above the baffle top), a significant increase in the flowrate injected is required to detect carry-under in the recirculation line (for both the chemiresistor and the capacitance measurement systems). On the other hand, when liquid Varso® is injected below the baffle (Bank A), the results obtained are similar to the no-baffle case. These results indicate that the baffle could also act as a choke-point for the emulsion phase, reducing the mass transfer between the bed regions on either side of the baffle.

Similar to Chapter 5 - Section 5.6, these results are still at an exploratory stage. More experimental and modelling work is required to understand better the impact of a baffle addition on the liquid losses due to vapour and liquid carry-under (in the emulsion).
6.7 References


Jahanmiri M. 2017. Use of a baffle to enhance the distribution of a liquid sprayed into a gas-solid fluidized bed. The University of Western Ontario. https://ir.lib.uwo.ca/etd/4590/.


Chapter 7

7 Conclusions & recommendations

This chapter summarizes the results, points out the most important findings and gives conclusions drawn from this research. In addition, several actions are proposed for future studies.

7.1 Conclusions

1) A model was developed to predict liquid losses carried under by wet agglomerates to the bottom of commercial Fluid Cokers. This model aggregates several interconnected mechanisms impacting the properties of wet agglomerates. These mechanisms include wet agglomerates formation, thermal drying, breakage from interaction with the bed, and agglomerate re-wetting.

2) A lab-scale tapered cylindrical cold flow recirculating fluidized bed was designed, built, and used to acquire data on agglomerate trajectories required by the model. A Radioactive Particle Tracking (RPT) method was used to capture the tracer-agglomerate motion throughout the entire lab-scale unit.

3) A new data processing method was introduced to enhance the accuracy of the position calculated through the standard CARPT method. The new methods include a physics-based systematic correction for radiation absorption by the bed and unit walls. It also uses smoothing procedures to mitigate the random error caused by the stochastic nature of radiation emissions. A key feature of the new method compared to other sophisticated RPT data processing methods (such as a Monte-Carlo approach) is its relative speed, which allowed for the processing of data files of considerable size within a reasonable time frame.

4) Based on the model predictions, the following conclusions can be achieved regarding a typical commercial Fluid Coker operation:
   a. Redistributing the flow of the lowest feedstock injection bank towards upper banks reduces the liquid losses to the bottom of the Fluid Coker by a factor of 0.10 for a “representative” agglomerate (such as defined in
Chapter 5). Similar improvement was also verified for other types of wet agglomerates, with varying amplitude of variations.

b. Adding a single baffle located between the lowest and the second-lowest feed rings reduces the liquid losses to the bottom of the Fluid Coker by a factor of 0.46 for a “representative” agglomerate (such as defined in Chapter 5). Similar improvement was also verified for other types of wet agglomerates, with varying amplitude of variations. One interesting feature associated with baffles is the apparition of choke-points, defined as zones where the cross-sectional area is reduced, restricting the downward crossing of agglomerates, and promoting their breakage.

c. Combining the redistribution of the flow of the lowest lateral injection bank towards upper banks with a single baffle between the lowest and the second-lowest feed rings (therefore located below all remaining active feed rings) reduces the liquid losses to the bottom of the Fluid Coker by a factor of 0.05, for a “representative” agglomerate (such as defined in Chapter 5). Similar improvement was also verified for other types of wet agglomerates, with varying amplitude of variations.

d. The addition of more than one baffle is beneficial only if all active injection banks located below the lowest baffle are inactive. With other configurations, negligible reductions in liquid losses are achieved compared to a single baffle configuration.

e. The nozzle penetration has a null to a moderate impact on the liquid losses to the bottom of the Fluid Coker when no baffle is used, depending on the initial agglomerates properties and the feed redistribution profile used. When at least one baffle is used, this impact is reduced even further.

f. The baffle flux-tubes have a neglectable impact on the liquid losses to the bottom of the Fluid Coker. Similar results are obtained whether the baffle has flux-tubes or not.

5) Exploratory experiments measuring liquid and vapour carried to the bottom of the Fluid Coker by the emulsion suggest that the baffles also help reduce this type of
liquid loss. On the other end, the redistribution of lateral injection profile does not impact this type of liquid loss.

7.2 Recommendations

7.2.1 Possible model improvements

1) To validate the liquid losses model and adjust it if necessary, future work should test the model against an experimental system that allows for direct measurements of liquid losses. Such a system could be a warm pilot-scale recirculated bed, operating at temperatures where a vaporisable liquid tracer could be injected. A comparison of the results of the model with a CFD simulation could also help to refine it. This would also be important to verify how a CFD simulation could be applied to help scale up the model.

2) The following recommendations are suggested to improve the agglomerate formation component of the model:
   a) The equation used to calculate the fraction of injected liquid trapped in an agglomerate of a given size (Li, 2016) could be improved by incorporating results from future literature studies.
   b) The secondary growth of wet agglomerate fragments, through the capture of dry particles, could be studied and included in the model if experiments suggest it significantly affects the agglomerate properties.
   c) The jet tip positions need to be measured more accurately to validate values predicted from general literature correlations.

3) The following recommendations are suggested to improve the agglomerate drying component of the model:
   a) A more complex drying model could be used. For instance, it could be replaced by the lump model proposed by House (2007). This model considers a variable agglomerate diameter $D_{\text{Agg}}$ and more than one cut of hydrocarbons.
   b) The impact of the gas phase on limiting vapour mass transfer, and especially possible local saturation, should be investigated to obtain a more accurate representation of agglomerate drying.
4) The following recommendations are suggested to improve the agglomerate breakage component of the model:
   a) Some work should focus on acquiring a more accurate radial distribution of bubbles, especially in regions where the standard core-annular model may not apply perfectly (with a bimodal distribution, for instance).
   b) For a bed with an extensive range of gas velocities, some work should be done bundling equations and empirical relations to accurately describe bubble properties for both bubbling and turbulent regimes. For instance, the bubble size may vary with the bubble position and the bed regime at this position. Another possibility would be to build in the model a probability of encounter between a wet-agglomerate and a bubble at a time t, similar to the approach of Shi et al. (2017)
   c) It would be interesting to consider the fragments resulting from agglomerate breakage instead of assuming their instant disintegration. Indeed, they could contribute to liquid losses, especially for the largest agglomerates that may produce fragments of significant size.
   d) Some development could be done regarding the empirical coefficient $M_\text{Rewet}$ to obtain more specific values.
   e) In general, to validate the equations and assumptions used, future work should obtain direct measurements of bed shear rate. For instance, Parveen (2011) used the RFID (radio frequency identification) technology to study the separation of two half agglomerates under a given stress. Similar methods could be transposed to obtain information about local bed shear within a recirculated fluidized bed.

5) To improve the agglomerate re-wetting component of the model, some experimental work could be done to obtain and validate values of the coefficient of re-wetting, $C_\text{Rewet}$. In addition, it may be essential to detect and quantify the probability that an agglomerate re-entering the jet cavity can be destroyed instead of only assuming re-wetting.

7.2.2 Ideas that could lead to the reduction of liquid losses in commercial Fluid Cokers

These suggestions could be tested with lab experiments and scaled up using the model.
1) To extend the results of the lateral redistribution profiles, some work should be done studying profiles with an uneven feed redistribution. Examples of such solutions are:
   a) Reinjecting all the redistributed flow into a single bank.
   b) Inject an uneven flowrate between successive banks
   c) Inject an uneven flowrate between nozzles of the same bank
2) The results obtained with a variable even nozzle penetration configurations gave variable results, depending on the initial agglomerates properties. Therefore, this analysis needs to be extended to consider the whole wet-agglomerate population. This would allow to better estimate the importance of these even nozzle penetration variations on liquid losses in commercial Fluid Cokers.
3) The results obtained with an uneven nozzle penetration configuration should motivate the study of other nozzle penetration arrangements that might reduce liquid losses. For example, previous research suggests that fractal gas distributors can be beneficial for specific purposes (Peng et al., 2018).
4) To extend the baffle(s) results, some future studies should focus exclusively on the near periphery of the baffle to better understand how agglomerate and bubble trajectories are modified. Specific interest should be given to the interactions between the baffle(s) flux-tubes and the agglomerates.
5) Commercial Fluid Cokers are typically tapered. Two designs of taper shape are used, with a similar angle but different heights. This research used and short taper (with a tapered section located below the injection zone). It would be interesting to experimentally study how a longer taper shape (extending throughout the injection zone) would impact the liquid losses predicted by the model.
7.3 References


Appendices

Appendix A: Fluid coke particle sized distribution

The particle size distribution of coke used by Syncrude Canada Ltd. in their commercial Fluid Coker (Fort McMurray, Alberta, Canada) and used in this research was measured using a HELOS (H2316) particle size analyzer (Sympatec GmbH).

Solids were sampled directly from the barrel sent by Syncrude, after sieving to remove agglomerates (“Before fluidization”). Solids were also sampled from the experimental bed after a typical run: 18 hours of fluidization, with a superficial gas velocity of 0.9 m·s$^{-1}$ at the bed surface (“After fluidization”). Each configuration was measured two times.

![Particle size distribution](image)

**Figure A-1.** Coke particles size distribution (before and after fluidization) measured with a HELOS (H2316) particle size analyzer (Sympatec GmbH)
A moderate increase in Sauter-Mean diameter is observed due to the loss of fines in the unit cyclone. This small variation in the solids population balance did not lead to any measurable variations in the results recorded (travel time to stripper, liquid losses predicted, …).

In addition, fresh fluid coke was added regularly (at least once a day) into the experimental unit to keep the bed mass and the solids population balance constant.
Appendix B: Calibration of the solids recirculation line: gas and solids flowrate calibration

To reproduce the solids circulation between the reactor and the burner in the commercial Fluid Coker unit, the experimental unit used in this research used a pneumatic transport line to recirculate the solids from the bottom of the bed to the freeboard.

This appendix describes the gas & solids flow calibration procedures and results used to control the recirculation flowrate.

1) **Gas flow calibration**

The 7 mm ID sonic orifice (University Machine Shop, Western University) used to control the gas flow injected into the pneumatic transport line was calibrated using the following procedure.

![Figure B-1. Calibration set up for sonic nozzle calibration.](image)

The calibration can proceed with the system depicted in Figure B-1. Calibration set up for sonic nozzle calibration. as follows:

a. Fully open the downstream valve.
b. Set the upstream pressure regulator and record the reading $P_1$.
c. Determine the mass flowrate $W_g$ with the reference flowmeter.
d. Slowly close the valve to increase the downstream pressure. Record the mass flowrate $W_g$ and the downstream pressure reading $P_2$.
e. Go back to step B- and repeat as many times as necessary.
The nozzle can be considered sonic. For a given upstream pressure $P_1$, if the gas flowrate ratio $W_G/W_{G,\text{MAX}}$ remains nearly constant for increasing the ratio of downstream pressure to upstream pressure $P_2/P_1$.

Results presented in Figure B-2 show that the nozzle was in a sonic regime for the operating conditions (typical conditions: $P_1 = 72$ psig, $P_2 \leq 8$ psig). The separation between sonic and non-sonic regimes is presented in Figure B-3.

The calibration curve used to obtain the mass flowrate of gas is presented in Figure B-4.
Figure B-3. Sonic/Non-sonic regime separation (hollow points were extrapolated)

Figure B-4. Recirculation gas flow calibration curve

2) Solids flow calibration
The solids flow calibration was carried using the system presented in Figure B-5.

First, the solid mass within the defluidized bed height was measured using the laser distance system describe in Chapter 3. Using the mass-distance calibration curve presented...
in Chapter 3, the initial bed mass is calculated. Then, the bed is fluidized, and the solid recirculation flowrate is set. The process of setting a chosen recirculation flowrate requires to: 1) set a gas flowrate using the calibrated sonic nozzle, 2) set the pinch valve to a chosen opening. The elbow pressure signal is recorded for 5 minutes using a pressure transducer. The control valves are switched (valve 1 closed while valve 2 opened). At the same time, a timer is started. After a predefined time interval (2-5 minutes), the control valves are switched back (valve 2 closed while valve 1 opened), and the timer is stopped. The solids recirculation is stopped, and the bed is defluidized. Finally, the bed height is measured again, and the new bed mass is calculated. Therefore, the bed mass variation can be obtained (this value is verified against the mass recovered in the barrel). The mass flowrate is easily obtained as the ratio of bed mass variation by the timer duration.

Figure B-6 & Figure B-7 presents two calibration curves obtained for a gas flowrate $F_G = 3.57 \times 10^{-2}$ kg·s$^{-1}$ and $F_G = 2.75 \times 10^{-2}$ kg·s$^{-1}$

![Figure B-6. Solids flow recirculation calibration curve](image)

(Gas mass flowrate = $3.57 \times 10^{-2}$ kg·s$^{-1}$)
Figure B-7. Solids flow recirculation calibration curve (Gas mass flowrate = $2.75 \times 10^{-2} \text{ kg} \cdot \text{s}^{-1}$)
Appendix C: Standard RPT and CARPT calibration performed

This appendix presents the CARPT calibrations. Three calibrations were realized. For each of them, the position of calibration points, the calibrations curve obtained, and back-calculations of the calibration points positions are presented.
Figure C-1. Calibration position (calibration 1)
Figure C-2. Calibration position (calibration 2)
Figure C-3. Calibration position (calibration 3)
CAL 1: \( L_4 = 8.7845 \times 10^{-2} NC_4^{-0.48269} \)  
\( R^2 = 0.9473 \)

CAL 2: \( L_4 = 8.6960 \times 10^{-2} NC_4^{-0.51526} \)  
\( R^2 = 0.93731 \)

CAL 3: \( L_4 = 7.5760 \times 10^{-2} NC_4^{-0.57110} \)  
\( R^2 = 0.98071 \)

CAL 1: \( L_5 = 5.8157 \times 10^{-2} NC_5^{-0.62751} \)  
\( R^2 = 0.96676 \)

CAL 2: \( L_5 = 7.5388 \times 10^{-2} NC_5^{-0.56781} \)  
\( R^2 = 0.94904 \)

CAL 3: \( L_5 = 5.1084 \times 10^{-2} NC_5^{-0.66728} \)  
\( R^2 = 0.97611 \)

CAL 1: \( L_6 = 5.3309 \times 10^{-2} NC_6^{-0.68874} \)  
\( R^2 = 0.95467 \)

CAL 2: \( L_6 = 3.1276 \times 10^{-2} NC_6^{-0.51250} \)  
\( R^2 = 0.95446 \)

CAL 3: \( L_6 = 4.8196 \times 10^{-2} NC_6^{-0.72566} \)  
\( R^2 = 0.96311 \)
CAL 1: $L_7 = 5.6475 \times 10^{-2} \ NC_7^{-0.69254}$  
$R^2 = 0.91459$

CAL 2: $L_7 = 5.9035 \times 10^{-2} \ NC_7^{-0.48196}$  
$R^2 = 0.91973$

CAL 3: $L_7 = 4.9284 \times 10^{-2} \ NC_7^{-0.69128}$  
$R^2 = 0.94167$

CAL 1: $L_8 = 6.4860 \times 10^{-2} \ NC_8^{-0.56167}$  
$R^2 = 0.85046$

CAL 2: $L_8 = 8.4087 \times 10^{-2} \ NC_8^{-0.48526}$  
$R^2 = 0.91122$

CAL 3: $L_8 = 6.2888 \times 10^{-2} \ NC_8^{-0.61995}$  
$R^2 = 0.89584$

CAL 1: $L_9 = 7.4663 \times 10^{-2} \ NC_9^{-0.48529}$  
$R^2 = 0.90375$

CAL 2: $L_9 = 8.6183 \times 10^{-2} \ NC_9^{-0.47663}$  
$R^2 = 0.93919$

CAL 3: $L_9 = 7.1843 \times 10^{-2} \ NC_9^{-0.52155}$  
$R^2 = 0.93890$
Figure C-4. CARPT calibrations curves for the twelve scintillation detectors (for the three calibrations)
Figure C-5. Back-calculations of the position of calibration points (calibration 1)
Figure C-6. Back-calculations of the position of calibration points (calibration 2)
Figure C-7. Back-calculations of the position of calibration points (calibration 3)
Appendix D: Solids RTD validation experiments with dyed coke

These experiments were conducted in collaboration with Sara Alsaadi (2017) and Nicole Ing (2018).

These dyed coke experiments were conducted to validate the Residence Time Distribution (RTD) data obtained with Radioactive Particle Tracking (RPT).

The experiments’ concept was to use food colour dye (Water-soluble food dye – Vanilla Food Company) to coat coke particles to measure the actual Residence Time Distribution (RTD) of these particles and compare it with the RTD data obtained with a radioactive tracer and RPT. It was decided to use more than one colour to minimize the amount of coke used per experiment: a careful selection of the wavelength to analyze for each dye allows using several colours on the same coke. The various dyes used, with their UV-Vis spectrophotometer (Thermo Fischer Scientific – Evolution™ 220) scan & calibration curve, are presented below.
1) **Calibration**

1. Red Dye

Figure D-1. Spectrophotometer scan & calibration curve for red dye
2. Yellow Dye

![Spectrophotometer scan & calibration curve for yellow dye](image)

\[ \lambda_{\text{max}} = 425 \text{ nm, 1% Dye} \]

\[ y = 0.2286x - 0.0377 \]

\[ R^2 = 0.9516 \]

**Figure D-2.** Spectrophotometer scan & calibration curve for yellow dye
3. Blue Dye

![Spectrophotometer scan & calibration curve for blue dye](image)

**Figure D-3.** Spectrophotometer scan & calibration curve for blue dye

\[
\lambda_{\text{max}} = 630 \text{ nm, 1\% Dye}
\]

\[
y = 0.4267x + 0.0050 \\
R^2 = 0.9932
\]
4. Green Dye

![Spectrophotometer scan & calibration curve for green dye](image)

\[ \lambda_{\text{max}} = 630 \text{ nm, 1\% Dye} \]

**Figure D-4.** Spectrophotometer scan & calibration curve for green dye
5. Orange Dye

![Spectrophotometer scan & calibration curve for orange dye](image)

\[ \lambda_{\text{max}} = 482 \text{ nm, 1\% Dye} \]

\[ y = 0.2934x + 0.0287 \]

\[ R^2 = 0.9927 \]

**Figure D-5.** Spectrophotometer scan & calibration curve for orange dye
6. **Purple Dye**

![Spectrophotometer scan & calibration curve for purple dye](image)

- $\lambda_{\text{max}} = 629$ nm, 1% Dye

**Figure D-6.** Spectrophotometer scan & calibration curve for purple dye
2) **Experimental procedure**

The whole procedure for dyed coke experiments, from coke dying to data analysis, can be described as follow:

**Materials:**

- 1.5 kg coke
- 40 mL colour dye (Vanilla Food Company)
- 960 mL of Deionized (DI) water

**Coke dying procedure:**

1. Mix colour dye & water in a beaker.
2. Add the dye/water mixture & 1 kg of coke into the pyrolysis shaker reactor developed by Sanchez Careaga (2020).
3. While the shaker reactor is in motion, use the induction heating to evaporate the water from the mixture and dry the dyed coke (150 °C for 30-40 minutes).
4. Pause the shaker reactor, let it cool down for 5 minutes, and then add 0.5 kg of coke (helps the coke cool down and be used for extra analysis later). Resume shaking for 10 minutes.

**Injecting dyed coke and taking coke samples:**

1. Take defluidized bed height measurements. These will be used to calculate the bed mass.
2. Using a funnel, fill the dyed coke feeder apparatus with 1 kg of coke.
3. Turn on the fluidization in the experimental unit. Open the gas and solids valve to allow coke pneumatic recirculation.
4. Set recirculation flowrate of 0.55 kg/s. Let the system stabilize for 5-to 10 minutes.
5. Open the valve allowing air to pressurize the dyed coke feeder.
6. Once feeder pressure reaches 15 psi, have someone open the feeder valve.
At the same time, another experimenter starts the screw sampler (motor at 50% speed) and begins taking batch samples (using the mason jars) at 5 second intervals.

7. After all the dyed coke has been ejected from the feeder (≈ 5-10 seconds), its valve should be closed.

Analyzing coke samples:

1. Mix 10 mL of the dyed coke sample with 30 mL of DI water (total mixture volume = 40 mL)
   Shake the diluted mixture, then let sit for at least 30 seconds.
2. Take an aqueous sample of the mixture using a syringe with a syringe filter (Hydrophobic PTFE, 0.45 μm pore size – Biomed Scientific) to remove any solids in suspension.
3. Transfer the aqueous sample into a spectrophotometer cuvette. Dispose of syringe and syringe filter after a single-use.
4. Analyze the sample using "Scan" mode on a UV-Vis Spectrophotometer

   **Spectrophotometer settings:**
   - **Smooth:** None
   - **Derivative:** None
   - **Absorbance max:** 2
   - **Absorbance min:** -0.1
   - **Start wavelength:** 700 nm
   - **End wavelength:** 400 nm
   - **Bandwidth:** 2 nm
   - **Integration time:** 0.5s
   - **Data interval:** 1 nm
   - **Scan speed:** 120nm/min
5. Analyze scanned data.

   **Wavelength integration ranges:**
   - **Red:** 487-517 nm
   - **Yellow:** 410-440 nm
Blue: 615-645 nm  
Green: 615-645 nm  
Orange: 469-499 nm  
Purple: 614-644 nm

Figure D-7. Example of absorbance spectrum measured during dyed coke experiments

3) **Dyed coke tracers vs. Single-particle radioactive tracer: Convolution & Deconvolution**

In the case of a single tracer method, such as the Radioactive Particle Tracking (RPT), the tracer measurement allows direct measurement of the vessel RTD, excluding the recirculation loop. From a mathematical perspective, the measurement is performed in a close-close boundary system, allowing for direct measurements of E(t). On the other hand, multiple-particles tracer methods, such as coke dye coating, do not allow this direct E(t) measurement. When dyed coke is sampled from the bed, it is impossible to know if it had been recirculated zero, one, two or more times. The output tracer concentration measured is the convolution of the coke RTD for a various number of recirculation pass.
Therefore, it is not possible to directly compare RTD results from RPT and coke dye coating. It either requires a deconvolution of measured dye concentration or a convolution of measured RPT Residence Time Distribution. The latter method was chosen for the sake of mathematical simplicity.

Let us consider an input tracer signal $C_{IN}(t)$ passing through a vessel of Residence Time Distribution (RTD) $E(t)$. The output signal measured $C_{OUT}(t)$ can be described as (Levenspiel, 1999):

$$C_{OUT}(t) = \int_0^t C_{IN}(t - t')E(t') dt' = \int_0^t C_{IN}(t')E(t - t') dt'$$

with: $t'$: time that tracer spent in the vessel

![Figure D-8. Single vessel signal convolution (modified from Levenspiel, 1999)](image)

By definition, $C_{OUT}$ is the convolution of $E$ with $C_{IN}$ such as:

$$C_{OUT} = E * C_{IN} = C_{IN} * E$$

In the case of a succession of $N$ closed vessel in series, this equation becomes:

$$C_{OUT} = \prod_{i=1}^N E_i * C_{IN} = E_1 * E_2 * ... * E_N * C_{IN}$$
For a recirculated vessel, the equations become more complex. However, the system can be approximated as a succession of $N$ closed vessels corresponding to the $N$ pass of the tracer within the bed. With $N$ high enough, the output signal $C_{\text{OUT}}(t)$ can be approximated.

**Figure D-9.** Recirculated vessel signal convolution (modified from Levenspiel, 1999)

The Residence Time Distribution $E(t)$ measured with RPT is convoluted using the scheme presented in Figure D-9. Various numbers of pass $N$ are computed and combined until the resulting profiles stop to vary with increasing $N$.

For a given pass $i \in [1; N]$, as presented in Figure D-9,

$$\text{If } j \leq i, \alpha_j = 0$$
$$\text{If } j > i, \alpha_j = 1$$

Furthermore, for this pass $i \in [1; N]$: 
\[ C_{OUT} = \prod_{j=1}^{i} E_j \ast C_{IN} \]

Then, in the case of a total of N pass, the combined convoluted output becomes:

\[ C_{OUT} = \sum_{i=1}^{N} \prod_{j=1}^{i} E_j \ast C_{IN} \]

RPT residence time distributions are obtained by compiling the time intervals required for the tracer to move from just above the bed surface (located at \( z = 1.35 \) m) to the stripper zone (located at \( z = 0.15 \) m), as presented in Figure D-10.

An example Top-to-Stripper Time Distribution measured with RPT is presented in Figure D-11. The cumulative and differentiated distributions are both plotted. The convolution of this Top-to-Stripper Time Distribution is shown in Figure D-12.
Figure D-10. Top-to-Stripper boundaries
Figure D-11. Top-to-Stripper Time Distribution (cumulative & differentiated) of RPT measurement (Short taper, $U_{G,\text{TOP}} = 0.9 \text{ m} \cdot \text{s}^{-1}$, 5 injection banks, $L_{\text{IN}} = 0 \text{ cm}$, $F_S = 0.50 \text{ kg} \cdot \text{s}^{-1}$, $d_{\text{Agg}} = 1.3 \text{ cm}$, $\rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3}$)
**Figure D-12.** Convolution of the Top-to-Stripper Time Distribution of RPT measurement

(Short taper, \( U_{G,\text{TOP}} = 0.9 \, \text{m} \cdot \text{s}^{-1} \), 5 injection banks, \( L_{\text{IN}} = 0 \, \text{cm} \), \( F_S = 0.50 \, \text{kg} \cdot \text{s}^{-1} \),

\[ D_{\text{Agg}} = 1.3 \, \text{cm}, \rho_{\text{Agg}} = 1100 \, \text{kg} \cdot \text{m}^{-3} \])

a) Including original differentiated Top-to-Stripper Time Distribution

b) Only convoluted Top-to-Stripper Time Distribution for various number \( N \) of recirculation pass

The objective is to compare these dyed coke time profiles with Radioactive Particle Tracking (RPT) residence time distributions. It was necessary to compensate for the distance between the stripper \((z = +0.15 \, \text{m})\) and the location of the screw sampler used in dye experiments \((z = -0.10 \, \text{m})\). Therefore, a critical experiment was conducted by injecting dyed coke at the stripper. It was used to measure the time \( t_{\text{delay}} \) required to reach the screw sampler in a couple of typical operation configurations. This time is normalized using theoretical overall bed residence time \( \tau \) defined, such as in the equation below.

\[
\tau = \frac{m_S}{F_S}
\]

with:

- \( m_S \): bed mass (kg)
• \( F_S \): solids recirculated flow (kg·s\(^{-1}\))

The normalized delay time \( t_{\text{delay}}/\tau \) as a function of solids recirculated flow \( F_S \) is presented in Figure D-13.

![Figure D-13. Normalized delay time as a function of solids recirculated flow](image)

Finally, the Top-to-Sampler RPT dimensionless residence time is obtained by combining the measured Top-to-Stripper RPT dimensionless residence time and the calibrated dimensionless delay time \( t_{\text{delay}}/\tau \).

References:

Appendix E: Radiation absorption in RPT measurements

To better understand the effect of the different absorptions, let us simulate the steel wall absorption and the bed absorption when the source is moved in a 2D plan within the bed while keeping the distance source-detector L constant, as presented in Figure E-1.

![Figure E-1. Motion of tracer within the bed, in a 2D plan, at a constant source-detector distance L](image)

The effective thickness of the steel wall (thickness traversed by the source-detector ray) and the absorption in the steel wall are presented in Figure E-2. The absorption does not vary significantly for most radial positions. However, for the extreme left and right positions, the effective wall thickness increases so much that the absorption increases up to 300%.
It should be noted that in the actual experimental unit, similar variations of steel wall absorption occur in 3D. If the radioactive source is located above or below the detector, the steel wall absorption will vary. Besides, since the 1D calibration curves are obtained in the empty vessel, they include the steel wall absorption. The fitted calibration coefficients include the average steel wall absorption. This variable steel wall absorption can partially explain the variabilities of the calibration data point around the fitted calibration curves.

Figure E-2. Motion of tracer within the bed, in a 2D plan, at a constant source-detector distance L

a) Effective steel wall thickness, b) Steel wall absorption
Therefore, the correction required is only an adjustment of the average fitted value, and therefore moderate.

The distance source-detector and the absorption in the bed are presented in Figure E-3. The absorption variation is significant: with the bed voidage used (constant value of $\varepsilon_{\text{bed}} = 0.42$), it ranged from almost zero to a 50% bed absorption.

![Figure E-3](image)

**Figure E-3.** Motion of tracer within the bed, in a 2D plan, at a constant source-detector distance $L$

a) Distance source-wall b) Bed (coke) absorption

It should be noted that, again, in the actual experimental unit, similar variations of bed absorption occur in 3D. If the radioactive source is located above or below the detector,
the bed absorption will vary. Nevertheless, contrary to the steel wall absorption, the 1D calibration curves do not include any bed absorption effect (they were acquired in an empty vessel). Therefore, the correction due to bed absorption will be more significant.
Appendix F: Overview of the smoothing methods considers for the RPT enhancements: description and analysis

1. **Smoothing of counts rate**
   
   1.1. **Average**

   The first method to increase the sampling time of each recorded counts rate consists of averaging the counts rate recorded for two or more sampling time.

   Let us define a raw data sequence of counts rate recorded with ND detectors, represented by \( CR(t) = \left[ CR_1, \ldots, CR_{ND} \right] \) (t), beginning at time \( t = 0 \) and including \( N \) events such as \( CR(t) = \{ CR_1, CR_2, \ldots, CR_i, \ldots, CR_N \} \) with \( i \in [1; N] \).

   For \( i \in [1; N/M] \), with \( M \) an integer such as \( M > 1 \):

   \[
   \overline{CR}_i = \frac{\sum_{j=(i-1)M+1}^{iM} \Delta t_j CR_j}{\sum_{j=(i-1)M+1}^{iM} \Delta t_j}
   \]  
   (F.1)

   By increasing the time interval used to calculate a position, this method would reduce the fraction of random error in measured counts rate by 30% or more, as presented in Chapter 4 – Section 4.3. On the other hand, it reduces the number of differentiated time events (division by a factor \( M \)), reducing the time resolution of the measurement.

   A significant advantage of average smoothing is its linear time complexity: a series of \( N \) events requires \( M.N \) smoothing operations.

   Figure F-1 presents the x-component of the tracer position vector obtained through CARPT with the average smoothing methods (using a window of two points) applied to the radiation counts detected for a length of 10 seconds (836 events).
Figure F-1. x-component of tracer position obtained by CARPT after Average Smoothing of radiation counts.

1.2. Average with linear interpolation
To keep the benefit of averaging counts rate while maintaining the time resolution of the measurement, a possible method is to perform a linear regression to re-calculate intermediate points in-between the averaged counts rate. This method is based on the idea that when the tracer physically moves within the bed, its trajectory is regular when the time variation is small enough (limited acceleration variations). For instance, if the counts rate is averaged every two time intervals (with an average time interval of 12 ms), the intermediate point is located within 12 ms of averaged data. It can therefore be reasonably re-calculated by interpolation (Suarez-Navarro et al., 2019).

Let us define a raw data sequence of counts rate recorded with ND detectors, represented by \( CR(t) = \left[ CR_1, ..., CR_{ND} \right](t) \), beginning at time \( t = 0 \) and including \( N \) events such as \( CR(t) = \{ CR_1, CR_2, ..., CR_i, ..., CR_N \} \) with \( i \in [1; N] \).

For \( i \in [1; (N - (M - 1))] \), with \( M \) an integer such as \( M > 1 \):

- If \( I = \frac{(i + M - 1)}{M} \) is an integer,
  \[
  \overline{CR}_i = \frac{\sum_{j=(i-1)M+1}^{iM} \Delta t_j \cdot CR_j}{\sum_{j=(i-1)M+1}^{iM} \Delta t_j}
  \]
- If \( I = \frac{(i + M - 1)}{M} \) is NOT an integer, then \( I_L = I(k) \) is defined as the closest lower integer \( (i = k) \) and \( I_H = I(l) \) is defined as the closest higher integer \( (i = l) \).

\[
\overline{CR}_i = \frac{CR_i - CR_k}{t_i - t_k} \cdot (t_i - t_k) + CR_k
\]

A significant advantage of average with linear interpolation smoothing is its linear time complexity: a series of \( N \) events requires \((M + 1)N\) smoothing operations.

Figure F-2 presents the x-component of the tracer position vector obtained through CARPT with the moving average (with linear interpolation) smoothing methods (using a moving window of two points) applied to the radiation counts detected for a length of 10 seconds (836 events).
Figure F-2. *x*-component of tracer position obtained by CARPT after Average Smoothing (with linear interpolation) of radiation counts rate

1.3. Central moving average
Another method to reduce the error by increasing the time sampling while maintaining the
time resolution of the measurement is to use a moving average: the counts rates at any time
are averaged using the neighbouring values.

Let us define a raw data sequence of counts rate recorded with ND detectors, represented
by \( CR(t) = \begin{bmatrix} CR_1 \\ \vdots \\ CR_{ND} \end{bmatrix}(t) \), beginning at time \( t = 0 \) and including \( N \) events such as
\( CR(t) = \{ CR_1, CR_2, \ldots, CR_i, \ldots, CR_N \} \) with \( i \in [1; N] \).

For \( i \in [1 + (M' + 1); N - (M' - 1)] \), with \( M' \geq 1 \):

\[
\overline{CR}_i = \frac{\sum_{j=i-M'}^{i+M'} \Delta t_j CR_j}{\sum_{j=i-M'}^{i+M'} \Delta t_j}
\]  

(F.3)

A significant advantage of central moving average smoothing is its linear time complexity:
a series of \( N \) events requires \( M'.N \) smoothing operations.

**Figure F-3.** x-component of tracer position obtained by CARPT after Moving Average
Smoothing of radiation counts rate
Figure F-3 presents the x-component of the tracer position vector obtained through CARPT with the moving average smoothing methods (using a moving window of two points) applied to the radiation counts detected for a length of 10 seconds (836 events).

2. Smoothing of coordinates

2.1. Central moving average

One can apply a central moving average on the calculated coordinates. Let us define a raw data sequence represented by $X(t) = \begin{bmatrix} x \\ y \\ z \end{bmatrix}(t)$, beginning at time $t = 0$ and including $N$ events such as $X(t) = \{X_1, X_2, \ldots, X_i, \ldots, X_N\}$ with $i \in [1; N]$.

For $i \in [1 + (M' + 1); N - (M' - 1)]$, with $M' \geq 1$:

$$\bar{X}_i = \frac{\sum_{j=1}^{M' + M} X_j}{(2 \cdot M' + 1)}$$

(F.4)

A significant advantage of central moving average smoothing is its linear time complexity: a series of $N$ events requires $M' \cdot N$ smoothing operations.

Figure F-4. Moving Average Smoothing on x-component of the tracer position vector
Figure F-4 presents the moving average smoothing methods (using a moving window of five points) applied to the x-component of the tracer position vector for a length of 10 seconds (836 events).

2.2. Exponential smoothing

Exponential smoothing is a standard window function used to smooth the time series of data. While methods such as moving averages give the same weight $1/N$ to each of the $N$ observations, exponential smoothing assigns exponentially decreasing weight to older observations. The rapidity of this decrease is controlled by one or more smoothing parameter(s). Several exponential smoothing methods exist: single exponential smoothing, double exponential smoothing, or triple exponential smoothing (NIST/SEMATECH, 2003).

A limitation of exponential smoothing is related to its 1D nature: it is applied independently on each component of the $[x, y, z](t)$ position vector. Besides, the first event of the series will not be significantly smoothed. The smoothing reaches a “steady state” after a couple hundred to a thousand data points. Due to the significant size of the data to process, it is not a problem since only a small fraction of the data points would not be in “steady-state smoothing”. Besides, a significant advantage of exponential smoothing is its linear time complexity: a series of $N$ events requires $N$ smoothing operations.

Let us define a raw data sequence represented by $X(t)$, beginning at time $t = 0$ and including $N$ events such as $X(t) = \{X_1, X_2, ..., X_i, ..., X_N\}$ with $i \in [1; N]$. The output of the exponential smoothing algorithm is noted $S(t)$, including $N$ events.

2.2.1. Single Exponential smoothing (Holt linear)

For the initial event ($i = 1$), the signal is not modified:

$S_1 = X_1 \quad \text{(F.5)}$

For the subsequent events ($i \in [2; N]$) the following equation is used, with $0 < \alpha < 1$:

$S_i = \alpha . X_i + (1 - \alpha) . X_{i-1} \quad \text{(F.6)}$
The smoothing parameter $\alpha$ is used to control the shape of the exponential decrease of past events weight: when $\alpha$ increases, the decrease is faster (if $\alpha = 1$, the output series is averaged over only the current observation), and when $\alpha$ increases, the decrease is slower (if $\alpha = 0$, the output series becomes a simple moving average). This simple form of exponential smoothing is also known as “exponentially weighted moving average (EWMA)”.

Figure F-5 presents the single exponential smoothing methods applied to the x-component of the tracer position vector for a length of 10 seconds (836 events). Three values of smoothing factor $\alpha$ are presented: 0.10, 0.25 & 0.50. It should be noted that the data series presented is an excerpt of a much larger series with a significant number of events processed before $t = 0$: the exponential smoothing can be considered in “steady-state”.

![Diagram](image-url)
Figure F-5. Single Exponential Smoothing on x-component of the tracer position vector
a) \( \alpha = 0.10 \), b) \( \alpha = 0.25 \), c) \( \alpha = 0.50 \)

2.2.2. Double Exponential smoothing

Single exponential smoothing performs poorly when there is a trend in the data. Therefore, recursive application of an exponential filter twice was introduced (NIST/SEMATECH,
The concept of double exponential smoothing is to introduce an additional term to consider possible underlying data slope. This slope component is itself updated via exponential smoothing. The literature describes two main methods: Brown’s linear double exponential smoothing and Holt-Winters double exponential smoothing.

- **Brown’s linear exponential smoothing**
  This method does not directly introduce a second exponential factor but uses a different set of equations to simultaneously use $\alpha$ as the first and second-order smoothing factor.
  For the first two initial events ($i = 1 \& i = 2$), the signal is not modified:
  \[
  S_1 = X_1 \\
  \text{with: } S'_1 = X_1 \\
  S''_1 = X_1
  \]
  For the subsequent events ($i \in [3; N]$) the following equation is used, with $0 < \alpha < 1$:
  \[
  S_i = 2.S'_i - S''_i
  \]
  \[
  \text{with: } S'_i = \alpha.X_i + (1 - \alpha).S'_{i-1} \\
  S''_i = \alpha.X_i + (1 - \alpha).S''_{i-1}
  \]

- **Holt-Winters double exponential smoothing**
  This method introduces a trend smoothing factor $\beta$ in addition to the first-order smoothing factor $\alpha$.
  For the first two initial events ($i = 1 \& i = 2$), the signal is not modified:
  \[
  S_1 = X_1 \\
  S_2 = X_2
  \]
  \[
  \text{with: } b_2 = X_2 - X_1
  \]
  For the subsequent events ($i \in [3; N]$) the following equation is used, with $0 < \alpha < 1$ & $0 < \beta < 1$:
  \[
  S_i = \alpha.X_i + (1 - \alpha). (S_{i-1} + b_{i-1})
  \]
  \[
  \text{with: } b_i = \beta. (S_i - S_{i-1}) + (1 - \beta). b_{i-1}
  \]

Figure F-6 presents the double exponential smoothing methods applied to the $x$-component of the tracer position vector for a length of 10 seconds (836 events). The two methods
discussed above are presented. Again, the exponential smoothing can be considered in a “steady-state”.

![Graph](image)

**Figure F-6.** Double Exponential Smoothing on x-component of the tracer position vector

a) Brown's linear ($\alpha = 0.25$), b) Holt-Winters ($\alpha = 0.25; \beta = 0.25$)

**2.2.3. Triple exponential smoothing (Holt Winters)**
Triple exponential smoothing was developed for data series, including trend and seasonality (NIST/SEMATECH, 2003). Since it did not position vector do not show seasonality, it was not investigated.

2.3. Kernel Smoother

A kernel smoother is a statistical method creating a continuous function of real values from the weighted average of neighbouring observed data. The weight of each data point is defined by the aforementioned kernel, such that closer points are given higher weights (Hastie et al., 2008). Several kernel functions are discussed in the literature. The calculated function produced by the kernel smoother represents a smoothed average of the overall data series. The level of smoothness is set by a single parameter.

An advantage of Kernel Smoothers is their ability to perform simultaneous smoothing on the three components of the tracer position vector. Let us define a raw data sequence represented by $X(t) = \begin{bmatrix} x \ y \ z \end{bmatrix}(t)$, beginning at time $t = 0$ and including $N$ events such as $X(t) = \{X_1, X_2, \ldots, X_i, \ldots, X_N\}$ with $i \in [1; N]$. The output of a Kernel Smoothers is a continuous function $\bar{X}(t)$. Within this continuous function, one can extract the $N$ smoothed points corresponding to the original series. The general equation of a Kernel Smoother to estimate the smoothed position at $j$ can be described as follow:

To estimate the smoothed position at $j$, for $i \in [1; N] \& j \in [1; N]$:

$$\bar{X}_j = \frac{\sum_{i=1}^{N} K_{h_\lambda}(X_j, X_i) \cdot X_i}{\sum_{i=1}^{N} K_{h_\lambda}(X_j, X_i)}$$  \hspace{1cm} (F.11)

With a kernel $K_{h_\lambda}$ being defined as:

$$K_{h_\lambda}(X_j, X_i) = D \left( \frac{\|X_i, X_j\|}{h_\lambda(X_j)} \right)$$  \hspace{1cm} (F.12)

with:

- $h_\lambda(X_j)$: kernel parameter (kernel radius)
• $D \left( L_{X_{i-j}} \right)$: positive real-valued function selected to decrease when the distance $L_{X_{i-j}} = \|X_i - X_j\|$ (Euclidian norm between $X_i$ and $X_j$) increases.

Contrary to exponential smoothing, kernel smoothers do not have a linear time complexity. For a series of $N$ events, they require $N \cdot N = N^2$ smoothing operations: kernel smoothers have a polynomial time complexity. This research produces significant datasets (regularly 1 to 20 million data points per run). Therefore, the polynomial time complexity represents a significant drawback with the data set of this research.

To mitigate this limitation, and since the importance of neighbouring data points decreases with distance, one can introduce a maximum number of points $2 \cdot N_{MAX}$ to be considered for smoothing. The equations presented above stay the same, but the parameter $i$ is no longer bounded by $i \in [1; N]$ but by $i \in [j - N_{MAX}; j + N_{MAX}]$. With a number of points $N_{MAX}$ small enough, the kernel function time complexity becomes linear: for a series of $N$ events, they require $2 \cdot N_{MAX} \cdot N$ smoothing operations (with $N_{MAX} = \text{Constant} \leq N$). To estimate the smoothed position at $j$, for $i \in [j - N_{MAX}; j + N_{MAX}]$ & $j \in [1; N]$:

$$\bar{X}_j = \frac{\sum_{i=j-N_{MAX}}^{j+N_{MAX}} K_{h_{\lambda}}(X_j, X_i) \cdot X_i}{\sum_{i=j-N_{MAX}}^{j+N_{MAX}} K_{h_{\lambda}}(X_j, X_i)}$$  \hspace{1cm} (F.13)

2.3.1. Pseudo-Gaussian Kernel Smoother

Gaussian Kernels are one of the most widely used kernels. They use a Gaussian function as a weighting function for the smoothing (Brett, 2014). For a 3D smoothing, the kernel can be described as:

$$K_{h_{\lambda}}(x_j, x_i) = \frac{1}{\left(\sqrt{2 \cdot \pi \cdot \sigma} \right)^3} \cdot \exp \left( - \frac{\|X_i - X_j\|^2}{2 \cdot \sigma^2} \right)$$  \hspace{1cm} (F.14)

with:

• $\sigma = \frac{\text{FWHM}}{\sqrt{\ln(2)}}$, Population standard deviation.

• FWHM: Full Width at Half Maximum (m). It is the width of the kernel, at half of the maximum of the height of the Gaussian (Locci-Lopez et al., 2018).
The Gaussian function is defined within \([- \infty; \infty]\): a Gaussian Kernel would theoretically require an infinite window length (all points in the data series are to be considered). This infinite window length is particularly detrimental in applications with a significant number of data points. However, thanks to its rapid decrease, it is often reasonable to apply the Gaussian only within a relatively narrow rectangular calculation window. This calculation window needs to be large enough to keep the general shape of the Gaussian smoothing while being small enough to significantly reduce the required computational time (Locci-Lopez et al., 2018). This modified Gaussian Kernel corresponds to the limitation of points to consider for smoothing to \(2.N_{\text{MAX}}\), as presented in the previous section. It is referred to as the Pseudo-Gaussian Kernel method.

Figure F-7 presents the Gaussian Kernel methods applied to the x-component of the tracer position vector for a length of 10 seconds (836 events). Two different FWHM parameters are presented: one close to the bed radius, one close to the bed parameter. The calculation window was set for \(N_{\text{MAX}} = 5000\) data points.
2.3.2. Nearest Neighbors Kernel Smoother

Nearest Neighbors Kernels are another kind of popular kernels. They are based on a simple principle: for each point, average the values of these neighbours of the M nearest neighbours (with a constant weight). Therefore, these kernels are sometimes referred to as Flat Kernel or Uniform Kernel. They can be described as (Schmidt-Thieme, 2007):

$$K_{h,\lambda}(X_j, X_i) = \begin{cases} 
\frac{1}{k} & , \text{if } t \leq 1 \\
0 & , \text{otherwise}
\end{cases}$$ (F.15)

with:

- \( t = \frac{\|X_i X_j\|}{\|X_i X_F\|} \) the normalized distance between point F & point j, with F the \( k \)th closest neighbour from J-

Figure F-8 presents the Nearest Neighbors Kernel methods applied to the x-component of the tracer position vector for a length of 10 seconds (836 events). Two different numbers
of neighbours to consider are presented. The calculation window was set for $N_{\text{MAX}} = 5000$ data points.

**Figure F-8.** Nearest Neighbors Kernel Smoother on x-component of the tracer position vector

a) $N_{\text{NEIGHBOURS}} = 25$, b) $N_{\text{NEIGHBOURS}} = 50$
2.3.3. Weighted Average Kernel Smoother

This family of kernels is based on the idea of defining a constant distance size $\lambda$ (kernel radius) and use a specific function to compute a weighted average for all data points within this window. Popular functions include the Epanechnikov Kernel or the Tri-Cube Kernel. Their shapes, compared to a Gaussian shape and Flat shape, are presented in Figure F-9.

- **Epanechnikov Kernel**

  \[
  K_{h\lambda}(X_j, X_i) = \begin{cases} 
  \frac{3}{4} \left(1 - t^2\right), & \text{if } t \leq 1 \\
  0, & \text{otherwise}
  \end{cases} 
  \] (F.16)

  with:
  - $t = \frac{\|X_i - X_j\|}{\lambda}$
  - $\lambda$: radius of the smoothing window (m)

- **Tri-Cube Kernel**

  \[
  K_{h\lambda}(X_j, X_i) = \begin{cases} 
  (1 - |t|^3)^3, & \text{if } t \leq 1 \\
  0, & \text{otherwise}
  \end{cases} 
  \] (F.17)

  with:
  - $t = \frac{\|X_i - X_j\|}{\lambda}$
  - $\lambda$: radius of the smoothing window (m)

Figure F-10 presents the Weighted Average Kernel methods applied to the x-component of the tracer position vector for a length of 10 seconds (836 events). The Epanechnikov kernel and the Tri-Cube kernel are presented for a parameter $\lambda = 10$ cm (one radius). The calculation window was set for $N_{\text{MAX}} = 5000$ data points.
Figure F-9. Comparison of shape of various Kernels. Each curve has been calibrated to integrate to 1 (Wong et al., 2012).

\[ \text{Smoothed data} \]

\[ \text{Raw data} \]

\[ \text{Epanechnikov} \]

\[ \lambda = 10 \text{ cm} \]
Figure F-10. Weighted Average Kernel Smoother on x-component of tracer position vector ($\lambda = 10$ cm)

a) Epanechnikov function, b) Tri-Cube function

2.4. Wavelets

Another approach discussed in the literature is wavelet filtering (Daubechies orthonormal wavelets) to reduce noise in positions calculated by CARPT (Degaleesan et al., 2002).

This method requires a constant frequency within the original dataset. As described before, the data acquisition system used in this research was designed to acquire the data as fast as possible (Sanchez Careaga, 2013), resulting in variable data acquisition frequency. Therefore, wavelet filtering required a pre-treatment of the CARPT coordinates datasets to interpolate them at a constant time interval: a linear interpolation method was used to obtain data with a constant time interval equal to the average time interval of the original dataset. It should be noted that this extra step is considered when determining the required computing time to perform the method.

The wavelet function used was orthogonal wavelets, specifically Daubechies wavelets (Daubechies, 1992) with two vanishing moments (D4/db2). The smoothing parameter used
was the fraction of coefficients to eliminate $F_R$, such as $0 \leq F_R \leq 1$. The closer $F_R$ to 1, the more aggressive the smoothing is.

**Figure F-11.** Wavelet filter on x-component of the tracer position vector

a) Fraction of coefficients to eliminate $F_R = 80\%$, b) Fraction of coefficients to eliminate $F_R = 90\%$
Figure F-11 presents the Wavelet Filter methods applied to the x-component of the tracer position vector for a length of 10 seconds (836 events). Two fractions of coefficients to eliminate $F_R$ were tested: 80% & 90%.

References:


Appendix G: Selection of the RPT enhancements processing methods: smoothing and iterative correction

Methods smoothing the counts measured and methods smoothing the coordinates obtained (through simple CARPT or using iterative methods) can be combined to obtain optimal results.

The selection criteria used are the ones described in Chapter 4 - Section 4.1.1. The criteria are:

- To reduce the proportion of positions detected outside bed limits.
- Limit the creation of zones of artificially low tracer presence within the bed.
- Reduce the average value of the moving standard deviation of the tracer position (reduce sharp unrealistic variations of positions).
- Limit required computing time.

At first, to be in a worst-case scenario, the smoothing procedures are applied to the data originally recorded, without the iterative correction. Once the optimal smoothing method is selected, it will be applied to data corrected with the iterative procedure.

1. **Smoothing of counts rate**

The smoothing of counts rate, also later referred to as “pre-processing” (since being applied before the CARPT procedure), was investigated to select the best method using the criteria described in Chapter 4 - Section 4.1.1.

The following methods are presented:

- Average counts rate.
- Moving-average of counts rate.
- Average counts rate, with linear interpolation of removed counts rate.

Several parameters are tested for each type of smoothing, such as the number of points used to perform the average.
Figure G-1 presents the variation of the proportion of positions detected outside the bed for various pre-processing procedures. Figure G-2 presents the relative variation of the proportion of zones of artificially low tracer presence within the bed for various pre-processing procedures (the reference is the data originally recorded). It can be observed that the more a smoothing process reduces the proportion of positions detected outside the bed, the more the relative variation of the proportion of zones of artificially low tracer presence within the bed increases. It should be noted that increasing the number of events used for these three processes both decreases the proportion of positions detected outside the bed and increase the fraction of zones of artificially low tracer presence within the bed: a compromise value needs to be selected.

Therefore, to better understand which smoothing to select, Figure G-3 presents the relative variation of the proportion of positions detected outside the bed as a function of relative variation of the proportion of zones of zones of artificially low tracer presence within the bed, for various pre-processing procedures (for both axes, the reference is the data originally recorded).

**Figure G-1.** Variation of the proportion of positions detected outside the bed for various pre-processing procedures.
**Figure G-2.** Relative variation of the proportion of zones of artificially low tracer presence within the bed for various pre-processing procedures. The reference is the data originally recorded.

Figure G-3 clearly shows that averaging the counts rate is a poor smoothing method compared to using a moving average or an average coupled with linear interpolation of removed data. Indeed, the reduction in the proportion of positions detected outside the bed is less efficient, while more zones with artificially low tracer presence are created within the bed volume. The moving average and average with linear interpolation procedure show the same trend of performances.

Figure G-4 presents the relative variation of the average value of moving standard deviation on calculated cartesian coordinates \([x, y, z]\) for various pre-processing procedures. It can be observed that the smoothing is much less efficient on the \(z\)-component of the coordinates. It is likely due to higher position fluctuations along the \(z\)-axis compared to the \(x\)- and \(y\)-axis. Again, the average procedure performs less efficiently than the others, with even increased noise on the \(z\)-component. Moving average and average with linear interpolation procedures perform similarly. For the three procedures, increasing the number of events used to averaged decreased the average value of the moving standard deviation.
Finally, Figure G-5 presents the variation of required computational time for various pre-processing procedures. It clearly shows that the smoothing procedures tested do not significantly impact the required computational time. The average procedure even reduced the required time (due to the reduction in the number of events to process through CARPT). For the other two procedures, increasing the number of events used to average did not significantly modify the required computational time.

In conclusion, regarding the smoothing of the counts rate, the best methods are the moving average procedure and the average with linear interpolation procedure. Using an average over two events allows for a significant reduction of the proportion of counts outside the bed (-35%) while limiting the relative increase of artificially low tracer presence inside the bed (+35-40%). Coordinates are also less irregular (-40% for x- & y-components, -15% for z-component). Finally, the required computational time is not significantly increased.
Figure G-4. Relative variation of the average value of moving standard deviation on calculated coordinates for various pre-processing procedures. The reference is the data originally recorded.

Figure G-5. Variation of required computational time for various pre-processing procedures.
2. **Smoothing of coordinates**

The smoothing of coordinates, also later referred to as “post-processing” (since being applied after the CARPT procedure), was investigated to select the best method using the criteria described in Chapter 4- Section 4.1.1.

The following methods are presented:

- Moving-average of coordinates
- Exponential smoothing
  - Simple
  - Double (Holt-Winters & Brown)
- Kernel smoothing
  - Pseudo-Gaussian
  - Nearest Neighbours
  - Weighted average (Epanechnikov & Tri-Cube)
- Wavelet smoothing

Several parameters are tested for each type of smoothing, such as the number of points used to perform the moving average, the coefficient(s) used for exponential smoothing, the number of events used for kernel calculations, or the kernel radius to consider.

Figure G-6 presents the variation of the proportion of positions detected outside the bed for various post-processing procedures. Figure G-7 presents the relative variation of the proportion of artificially low tracer presence within the bed for various post-processing procedures (the reference is the data originally recorded). Similar to what was observed with the smoothing of counts rate, the more a smoothing process reduces the proportion of positions detected outside the bed, the more the relative variation of the proportion of artificially low tracer presence within the bed increases: a compromise value needs to be selected. Besides, the following remarks can be made:
• Moving-average of coordinates: an increased number of events used both decrease the proportion of positions detected outside the bed and increase the fraction of zone of artificially low tracer presence created within the bed.

• Exponential smoothing: increasing the value of α and/or β increases the proportion of positions detected outside the bed and decreases the fraction of zone of artificially low tracer presence created within the bed.

• Kernel smoothing:
  o increasing the number of events used for kernel calculations decreases the proportion of positions detected outside the bed and increases the fraction of zone of artificially low tracer presence created within the bed.
  o increasing the kernel radius (number of neighbours to consider for the nearest neighbour kernel) increases the proportion of positions detected outside the bed and decreases the fraction of zone of artificially low tracer presence created within the bed.

• Wavelet smoothing: increasing the fractions of coefficients to remove F_R decreases the proportion of positions detected outside the bed and increases the fraction of zone of artificially low tracer presence created within the bed.

Therefore, to better understand which smoothing to select, Figure G-8 presents the relative variation of the proportion of positions detected outside the bed as a function of relative variation of the proportion of zone of artificially low tracer presence within the bed, for various post-processing procedures (for both axes, the reference is the data originally recorded).
Figure G-6. Variation of the proportion of positions detected outside the bed for various post-processing procedures.

Figure G-7. Relative variation of the proportion of zone of artificially low tracer presence within the bed for various post-processing procedures. The reference is the data recorded initially.
Figure G-8 clearly shows that the moving average of coordinates and the various kernel smoothing are inferior smoothing methods compared to exponential smoothing (simple or double) or wavelet smoothing. Indeed, the reduction in the proportion of positions detected outside the bed is less efficient, while more zones of artificially low tracer presence are created within the bed volume. The simple exponential smoothing method is more aggressive than the double exponential smoothing methods or wavelet smoothing, leading to a higher reduction of positions outside at the cost of an increased augmentation of zones of artificially low tracer presence.

![Figure G-8](image)

**Figure G-8.** Relative variation of the proportion of zone of artificially low tracer presence within the bed as a function of the variation of the proportion of positions detected outside the bed, for various post-processing procedures.

Figure G-9 presents the relative variation of the average value of moving standard deviation on calculated cartesian coordinates [x, y, z] for various post-processing procedures. Again, the smoothing is much less efficient on the z-component of the coordinates. The nearest
neighbour kernel smoothing stands out as the method offering the most significant reduction in the average of moving standard deviation. All the other methods offer similar smoothing of coordinates.

![Graph showing relative variation of average value of moving standard deviation on calculated coordinates for various post-processing procedures.](image)

**Figure G-9.** Relative variation of the average value of moving standard deviation on calculated coordinates for various post-processing procedures. The reference is the data originally recorded.

Finally, Figure G-10 presents the variation of required computational time for various pre-processing procedures. It clearly shows that the moving average and exponential smoothing (simple and double) smoothing procedures do not significantly impact the required computational time. On the other hand, kernel smoothing procedures come at the cost of significantly higher required computational time. This required computational time increase with the number of events used for kernel calculations: doubling the number of events to consider more than double the required computational time. Regarding wavelets smoothing, it should be noted that the higher required computational is likely due to the extra processing step required (interpolation of data to obtain constant time interval). All the other smoothing parameters impact only marginally the required computational time.
In conclusion, regarding the smoothing of coordinates, the best methods are the exponential smoothing procedures, either simple or double. The double exponential smoothing methods (Holt-Winters & Brown) allow for more control on the level of smoothing. Using coefficients $\alpha$ (and $\beta$ when necessary) of 0.25 allows for a significant reduction of the proportion of counts outside the bed (- 30-40%) while limiting the relative increase of zones of artificially low tracer presence inside the bed (+ 45-50%). Coordinates are also less irregular (- 35% for x- & y-components, - 15% for z-component). Finally, the required computational time is not significantly increased.

3. **Smoothing of counts rate & smoothing of coordinates combined**

This section combines the best smoothing of the counts rate with the best smoothing of coordinates.

Smoothing of counts rate considered:

- Moving average counts rate.
- Average counts rate, with linear interpolation of removed counts rate.

Smoothing of coordinates considered:

- Single Exponential smoothing.
- Double Exponential smoothing (Holt-Winters & Brown).
- Nearest Neighbour Kernel.
- Wavelet smoothing.

Several parameters are tested for each type of smoothing, such as the number of points used to perform the average, or the coefficient(s) used for exponential smoothing.

Again, the smoothing was investigated to select the best method using the criteria described in Chapter 4 - Section 4.1.1.

**Figure G-11.** Variation of the proportion of positions detected outside the bed for various pre- & post-processing procedures combined.

Figure G-11 presents the variation of the proportion of positions detected outside the bed for various pre- & post-processing procedures. Figure G-12 presents the relative variation
of the proportion of zones of artificially low tracer presence within the bed for various pre-
& post-processing procedures (the reference is the data recorded initially).

Therefore, to better understand which smoothing to select, Figure G-13 presents the
relative variation of the proportion of positions detected outside the bed as a function of
relative variation of the proportion of zones of artificially low tracer presence within the
bed, for various pre-processing procedures (for both axes, the reference is the data
originally recorded).

Figure G-12. Relative variation of the proportion of impossible zones of artificially low
tracer presence within the bed for various pre- & post-processing procedures combined.

The reference is the data recorded initially.
Figure G-13. Relative variation of the proportion of zones of artificially low tracer presence within the bed as a function of the variation of the proportion of positions detected outside the bed, for various pre- & post-processing procedures combined.

Figure G-13 clearly shows that the best performances are obtained for a combination of average counts rate (with linear interpolation of removed counts rate) and double exponential smoothing (either Holt-Winters or Brown). A close second is the combination of average counts rate (with linear interpolation of removed counts rate) and wavelet smoothing. In the third position comes the combination of moving average counts rate and double exponential smoothing (either Holt-Winters or Brown). Indeed, the reduction in the
proportion of positions detected outside the bed is efficient, while few zones of artificially low tracer presence are created within the bed volume.

Figure G-14 presents the relative variation of the average value of moving standard deviation on calculated cartesian coordinates \([x, y, z]\) for various pre- & post-processing procedures. Again, the smoothing is much less efficient on the \(z\)-component of the coordinates. The smoothing of coordinates is the most efficient for combinations involving nearest neighbour kernel smoothing and single exponential smoothing. Nonetheless, combinations including double exponential or wavelet smoothing offer very similar results.

![Relative variation of the average value of moving standard deviation on calculated coordinates](image)

**Figure G-14.** Relative variation of the average value of moving standard deviation on calculated coordinates for various pre- & post-processing procedures combined. The reference is the data originally recorded.

Finally, Figure G-15 presents the variation of required computational time for various pre-processing procedures. It clearly shows that, outside the combinations, including nearest neighbours kernel smoothing, the smoothing procedures tested do not significantly impact the required computational time. Again, for wavelet smoothing, the extra required time is due to the extra processing step of interpolation.
In conclusion, regarding the combination of smoothing of counts rate and smoothing of coordinates, the best methods are the combination of average counts rate (with linear interpolation of removed counts rate) and double exponential smoothing (either Holt-Winters or Brown). Using an average over two events and a parameter $\alpha$ (and $\beta$ if necessary) of 0.25 allows for a significant reduction of the proportion of counts outside the bed (- 42-47%) while limiting the relative increase of zones of artificially low tracer presence inside the bed (+ 45-50%). Coordinates are also less irregular (- 40% for x- & y-components, - 15% for z-component). Finally, the required computational time is not significantly increased.
Appendix H: Calibration of pressure transducers used

The pressure transducers used in this thesis were from the ASDX Series Silicon Pressure Sensors (Honeywell, 2012). They were calibrated against a U-tube water manometer.

\[ P = 3456.498 \ U - 8648.493 \]
\[ R^2 = 0.999 \]

\[ P = 16636.044 \ U - 42022.259 \]
\[ R^2 = 0.930 \]

**Figure H-1.** Differential 0-1.0 psi pressure sensor (ASDXRRX001PDAA5) calibration against U-tube water manometer

**Figure H-2.** Differential 0-15 psi pressure sensor (ASDXRRX015PDAA5) calibration against U-tube water manometer
Figure H-3. Differential 0-10 inch H₂O pressure sensor (ASDXRRX010NDAA5) calibration against U-tube water manometer

References:

Appendix I: Spatial and Temporal scaling of the model

By imposing the same gas velocities (bottom & top) and the same recirculated solid flux, the hydrodynamics are similar in the experimental unit and the full-scale unit. But since the experimental is significantly smaller than the commercial coker, its solid residence time is significantly smaller.

1) Spatial scale-up factors $\gamma_X$, $\gamma_Y$ & $\gamma_Z$

$$
\begin{align*}
\gamma_X &= \gamma_Y = \frac{D_{COK}}{D_{EXP}} \\
\gamma_Z &= \frac{H_{COK}}{H_{EXP}}
\end{align*}
$$

with:

- $\gamma_X$, $\gamma_Y$ & $\gamma_Z$: spatial scale-up factor, respectively on x-, y-, and z-axis
- $H_{EXP}$: height of experimental unit (m)
- $H_{COK}$: height of commercial coker (m)
- $D_{EXP}$: diameter of experimental unit (m)
- $D_{COK}$: diameter of commercial coker (m)

Since R/D kept constant

$$
\gamma = \gamma_X = \gamma_Y = \gamma_Z
$$

with:

- $\gamma$: overall spatial scale-up factor

By selecting carefully the origin of each coordinate systems, the following relation can be established:

$$
\begin{bmatrix}
x' \\
y' \\
z'
\end{bmatrix} = \delta \cdot \begin{bmatrix}
x \\
y \\
z
\end{bmatrix}
$$

with:

- $[x, y, z]$: agglomerate position in experimental unit (m)
• \([x', y', z']\): agglomerate position in commercial coker (m)

2) **Temporal scale-up factors** \(T\)

\[
\tau_i = \frac{m_{S,i}}{F_{S,i}}
\]

(I.4)

with:

• \(\tau_{\text{EXP}}\): mean solids residence time in commercial coker or experimental unit (s)
• \(m_{S,i}\): mass of solids in commercial coker or experimental unit (kg)
• \(F_{S,i}\): mass flowrate of solids in commercial coker or experimental unit (kg·s\(^{-1}\))

\[
t' = t \cdot \frac{\tau_{\text{COK}}}{\tau_{\text{EXP}}} = t \cdot T
\]

(I.5)

with:

• \(t\): agglomerate time in experimental unit (s)
• \(t'\): agglomerate time in commercial coker (s)
• \(\tau_{\text{EXP}}\): mean solids residence time in experimental unit (s)
• \(\tau_{\text{COK}}\): mean solids residence time in commercial coker (s)

3) **Radioactive Particle Tracking (RPT) scale-up procedure**

Measured RPT data in experimental unit:

• Agglomerate position: \(\begin{bmatrix} x \\ y \\ z \end{bmatrix}(t)\)
• Agglomerate velocity: \(u_{\text{Agg}}(t)\)
• Bubble diameter & velocity: \(d_b\) & \(u_B(t)\)

Equivalent data in commercial unit:

• Agglomerate position: \(\begin{bmatrix} x' \\ y' \\ z' \end{bmatrix}(t') = \delta \cdot \begin{bmatrix} x \\ y \\ z \end{bmatrix}(t \cdot T)\)
• Similar for jets & internals position

• Same gas velocity profile and recirculated solids flux
  - Same agglomerate velocity: $u_{\text{Agg}}(t)$
  - Same bubble diameter & velocity: $d_b$ & $u_B(t)$
Appendix J: Reduction of residual slugging in the experimental unit

Calculation of predicted bubble size using Darton (1977) model with the following configuration:

- Fluid coke and air
- Sparger loop with 32 holes (Sanchez Careaga, 2013)
- $D_{BED,\text{Bottom}} = 19.68 \text{ cm}$
- $u_{G,\text{Spg}} \in [0.1; 1.0] \text{ m}\cdot\text{s}^{-1}$

This system corresponds to the worst-case scenario for slugging in the tapered unit (smallest diameter, highest gas velocities).

![Figure J-1](attachment:figure.png)

**Figure J-1.** Predicted bubble size as a function of bed height

Measurement system:

- Visual detection of slugging
  - Light & digital camera in a black box
- 10 minutes of video recording per run

- Pressure measurements
  - Eight pressure taps along the bed height, measurements with a digital manometer
  - Two taps with a pressure transducer

The original sparger setup studied was the one used by Sanchez Careaga (2013):
• Sparger double loop
• 32 holes (d_H = 1.6 mm)
• Upward injection

Figure J-3. Original sparger setup

Results obtained with the original configuration show intermittent slugs for low gas velocities (bubbling regime).

• Clear presence of slugs (detected visually)
• Low frequency of slugs when detected (less than two slugs per minute on average, for a 10 minutes recording)
• Disappearance of slugs at the transition Bubbling-Turbulent, around 0.5-0.6 m·s^{-1}
Figure J-4. Example of slug recorded by the camera, successive frames

\((u_G = 0.35 \text{ m/s}^1)\)
Figure J-5. Number of slugs visually detected
    (original sparger configuration)

The ratio of pressure drop indicates a poor sparger operation for low gas velocities.

Figure J-6. Pressure drop ratio between the bed and the sparger
    (original sparger configuration)
To increase the sparger pressure drop and reduce the residual slugging, the sparger configuration is modified.

The modified sparger setup studied is:

- Sparger double loop (kept the same)
- 24 holes ($d_H = 1.6$ mm, kept the same)
- Downward injection

![Figure J-7. Modified sparger setup](image)

Results obtained with the modified configuration still show a presence of intermittent slugs for low gas velocities (bubbling regime) but reduced.

- Still present (detected visually)
- Reduced frequency of slugs when detected (less than one slug per minute on average, for a 10 minutes recording)
- Transition Bubbling-Turbulent still in the same range of gas velocities
Figure J-8. Number of slugs visually detected  
(original & modified sparger configuration)

The ratio of pressure drop was increased, explaining the improvement.

Figure J-9. Pressure drop ratio between the bed and the sparger  
(original & modified sparger configuration)
The new sparger configuration offers a better gas distribution (less slugging). In addition, experiments show that for the range of velocities that will be investigated in the taper unit:

- Only intermittent slugging (less than one slug·min⁻¹) may appear at low velocities ([0.2; 0.5] m·s⁻¹)
- No slugging will appear at high velocities ([0.5; 0.9] m·s⁻¹)
  - No grinding of Fluid Coke required (to go from Group B to Group A powder)

It should be noted that another solution could have been to design an injection system with two spargers, one used for the low velocity range and the other used for the high velocity range.

References:

Appendix K: Measurement of the bed voidage profile in the experimental unit

1) Axial variation

The variation of the average voidage $\varepsilon_{\text{bed,AVG}}(z)$ is expressed as a function of height $z$. To obtain this average voidage $\varepsilon_{\text{bed,AVG}}(z)$, the first step consists in calculating the variation of average voidage $\varepsilon_{\text{bed,AVG}}(z)$ as a function of superficial gas-velocity $u_G$. This can be obtained from pressure-based measurements (Botterill et al., 1982; Lucas et al., 1986), acquired in the well-established zone of the bed, between $z_L$ & $z_H$ (see Figure K-1).

First, the bed average density $\rho_{\text{bed,AVG}}(z)$ is calculated as follow:

$$\rho_{\text{bed,AVG}}(z = z_C) = \frac{P_{\text{bed}}(z_L) - P_{\text{bed}}(z_H)}{g \cdot (z_H - z_L)} = \frac{\Delta P_{\text{bed,AVG}}(z_C)}{g \cdot \Delta H_{\text{bed}}(z_C)}$$  \hspace{1cm} (K.1)

with:

- $\rho_{\text{bed,AVG}}(z)$: average bed density in the pressure cell of center $z$ (Pa)
- $z$: axial location (m)
- $\Delta H_{\text{bed}}(z)$: height of the pressure cell of center $z$ (Pa)
  - $z_L$: axial location of the lowest pressure taps (m)
  - $z_H$: axial location of the highest pressure taps (m)
  - $z_C = (z_L + z_H)/2$: axial location of the pressure cell (m)
- $\Delta P_{\text{bed,AVG}}(z)$: bed pressure drop in the pressure cell of center $z$ (Pa)
  - $P_{\text{bed}}(z)$: bed pressure at location $z$ (Pa)

Then the average voidage $\varepsilon_{\text{bed,AVG}}(z)$ is expressed as:

$$\varepsilon_{\text{bed,AVG}}(z) = 1 - \frac{\rho_{\text{bed,AVG}}(z)}{\rho_S - \rho_G}$$  \hspace{1cm} (K.2)

Figure K-2 shows the variation of average bed voidage $\varepsilon_{\text{bed,AVG}}(z)$ with the superficial gas velocity, with the pressure sensors located in a well-established zone. The measurements were done in a bed with and without a ring baffle (located at Bank B, without flux-tubes).
Figure K-1. Location of pressure-based average bed voidage measurements (performed with/without the baffle)
**Figure K-2.** Variation of average bed voidage measured with the superficial gas velocity, with the pressure sensors located in a well-established zone (Without baffle & With a baffle located below bank B)

In Figure K-2, two zones are clearly visible, before and after transition bubbling turbulent. Before the transition, the voidage increases quickly, after it is mostly stable. This transition velocity is identified at about \([0.4; 0.6] \text{ m\cdot s}^{-1}\) (see Chapter 2 – Section 2.5.4). A local minimum of average bed voidage can be identified near the transition. It should be noted that strong differences are visible between the results acquired with and without a ring baffle. With a baffle, the voidage decrease near the transition is much more pronounced.

Using these results, an empirical correlation was developed to obtain the average bed voidage \(\varepsilon_{\text{bed,AVG}}\) as a function of the cross-sectional superficial gas velocity \(u_G\).

\[
\varepsilon_{\text{bed,AVG}}(u_G) = A \cdot u_G^6 + B \cdot u_G^5 + C \cdot u_G^4 + D \cdot u_G^3 + E \cdot u_G^2 + F \cdot u_G + G
\]  \(\text{(K.3)}\)

Table K-1 shows the empirical coefficients obtained for the configuration with and without a ring baffle.
Table K-1. Parameters of empirical polynomial correlation between average bed voidage and superficial gas velocity

<table>
<thead>
<tr>
<th></th>
<th>A</th>
<th>B</th>
<th>C</th>
<th>D</th>
<th>E</th>
<th>F</th>
<th>G</th>
<th>$R^2$</th>
</tr>
</thead>
<tbody>
<tr>
<td>No Baffle</td>
<td>-1.33</td>
<td>7.09</td>
<td>-0.147</td>
<td>14.7</td>
<td>-7.37</td>
<td>1.74</td>
<td>0.410</td>
<td>0.957</td>
</tr>
<tr>
<td>Baffle</td>
<td>-2.54</td>
<td>13.0</td>
<td>-25.4</td>
<td>23.4</td>
<td>-10.1</td>
<td>1.79</td>
<td>0.420</td>
<td>0.917</td>
</tr>
</tbody>
</table>

By using the correlation described by Equation (K.3) and the cross-sectional gas velocity profile, the average bed voidage $\varepsilon_{\text{bed, AVG}}$ can be expressed as a function of the bed height $z$. For the bed configuration with a ring baffle, the following assumptions are made: below the baffle location, the empirical coefficients used are the ones obtained for the bed without a baffle, and above the baffle location, the empirical coefficients used are the ones obtained for the bed with a baffle. Figure K-3 presents the evolution of the average bed voidage $\varepsilon_{\text{bed, AVG}}$ as a function of the bed height $z$, with and without a baffle.

---

---
In Figure K-3, the profile without a baffle shows a mostly flat average voidage $\varepsilon_{\text{bed,AVG}}$ along the whole bed height. On the other hand, when a baffle is present, a substantial decrease of the average voidage $\varepsilon_{\text{bed,AVG}}$ is visible at the baffle location. Then, the average voidage increases again, reaching similar value as for the no baffle case in the upper part of the bed. This sudden average voidage decrease at the baffle level is due to the local minimum identified in Figure K-2.

2) **Local variation of radial voidage**

The relative variation of voidage $\varepsilon_{\text{bed}}(r)/\varepsilon_{\text{bed,AVG}}(z)$ is expressed as a function of the height $z$ and the radial position $r$. To obtain this relative variation of voidage $\varepsilon_{\text{bed}}(r)/\varepsilon_{\text{bed,AVG}}(z)$, the first step consists in calculating the average pseudo 1D bed voidage $\varepsilon_{\text{bed,LIN}}(z)$ as a function of superficial gas-velocity $u_G$. This pseudo 1D voidage is calculated from radiation-based measurements (Bhowmick et al., 2015; Pant et al., 2017) and is expressed based on equation (3.9), considering only the bed absorption:
\[
\frac{\text{Counts}(u_G)}{\text{Counts}_{EMPTY}(u_G)} = \exp(-\alpha_j \cdot L_j \cdot \rho_j) \\
= \exp\left(-\alpha_j \cdot L_j \cdot \rho_S \cdot \left(1 - \varepsilon_{\text{bed,LIN}}(u_G)\right)\right)
\]

(K.4)

It can be rearranged as:

\[
\ln\left(\frac{\text{Counts}(u_G)}{\text{Counts}_{EMPTY}(u_G)}\right) = A(u_G) \cdot \left(1 - \varepsilon_{\text{bed,LIN}}(u_G)\right)
\]

(K.5)

This average voidage is “pseudo 1D” because the measurement zone restricted to the volume between the radioactive source and the radiation sensor. Due to the small distance source-detector, this volume is approximated as an arc of the cylindrical cross-section of the bed.

The radiation measurements can be correlated to a voidage using the two boundary cases of measurements: for an empty bed, no solids are presents and therefore \(\varepsilon_{\text{bed,LIN}} = 1.0\), and for a packed bed \((u_G = 0)\), \(\varepsilon_{\text{bed,LIN}}\) is approximated equal to 0.42 (voidage at minimum fluidization). A linear regression is used with the data points from the empty & packed bed case to calculate the coefficient \(A(u_G)\) in Equation (K.5). Then, using the calibrated coefficient \(A(u_G)\), the pseudo 1D bed voidage measured with radiation transmission is obtained for any fluidization velocity \(u_G\) with:

\[
\varepsilon_{\text{bed,LIN}}(u_G, X) = 1 - A(u_G) \cdot \ln\left(\frac{\text{Counts}(u_G, X)}{\text{Counts}_{EMPTY}(u_G, X)}\right)
\]

(K.6)

All radiation transmission measurements were acquired in the well-established zone of the bed, near \(z_C\) and the location where the pressure based voidage measurements were conducted (see Figure K-1).
Figure K-4 presents the average pseudo 1D bed voidage $\varepsilon_{\text{bed,LIN}}(z)$, from radiation transmission measurements, and compares it with the average volumetric bed voidage $\varepsilon_{\text{bed,AVG}}(z)$, from pressure based measurements.

**Figure K-4.** Variation of average voidage measured with pressure sensors at a given height $z$ (with & without baffle)

a) From pressure-based measurements, b) From radiation-based measurements

In Figure K-4, it is clearly visible that the pseudo 1D voidage (from radiation transmission measurements), are always bigger than the volumetric voidage (from pressure-based measurements). This is due to the difference between 1D measurements along an arc vs. 3D measurement in a bed with a cylindrical cross-section. It should also be noted that the addition of a baffle leads to significantly more noise in the measurements.

The next step consists in getting the empirical correlation correlating the pseudo 1D voidage $\varepsilon_{\text{bed,LIN}}(z)$ (from the radiation transmission measurements) as a function of the cross-sectional superficial gas velocity $u_G$

$$
\varepsilon_{\text{bed,LIN}}(u_G) = A \cdot u_G^6 + B \cdot u_G^5 + C \cdot u_G^4 + D \cdot u_G^3 + E \cdot u_G^2 + F \cdot u_G + G \quad (K.7)
$$
Table K-2 shows the empirical coefficients obtained for the configuration with and without a ring baffle.

<table>
<thead>
<tr>
<th></th>
<th>A</th>
<th>B</th>
<th>C</th>
<th>D</th>
<th>E</th>
<th>F</th>
<th>G</th>
<th>( R^2 )</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>No Baffle</strong></td>
<td>0.877</td>
<td>-3.26</td>
<td>3.19</td>
<td>1.12</td>
<td>-3.31</td>
<td>1.72</td>
<td>0.412</td>
<td><strong>0.983</strong></td>
</tr>
<tr>
<td><strong>Baffle</strong></td>
<td>-0.158</td>
<td>0.888</td>
<td>-2.18</td>
<td>3.06</td>
<td>-2.58</td>
<td>1.21</td>
<td>0.41</td>
<td><strong>0.849</strong></td>
</tr>
</tbody>
</table>

Then, one can compare of the values of the volumetric average voidage \( \varepsilon_{\text{bed,AVG}}(z) \), measured with pressure drop (3D measurement) and the pseudo 1D voidage \( \varepsilon_{\text{bed,LIN}}(z) \), measured with radiation transmission (pseudo-1D measurements). These values are compared using Equations (K.3) and (K.7), both expressed as a function of superficial gas-velocity \( u_G \).

The difference between these two values are linked to different bed volumes considered in each measurement method. With pressure measurement, the whole 3D volume located in-between the two measurements taps is considered. Therefore, it results in a “true” 3D measurement (with a 3D cylindrical integration of the 3D voidage profile). On the other hand, radiation transmission measurement can only measure phenomena in the section of the bed located in between the radioactive source and the radiation detectors (see Figure K-5-a)). For small distance, this can be approximated as pseudo 1D measurements. In a cylindrical bed, this can lead to an overestimation of the measured voidage, due to a linear integration of the voidage 3D profile in a circular cross-section (see Figure K-5-b)).

As shown in Figure K-5-c), these two measurements should give the same voidage values only if the 3D voidage profile is flat. In other case, the pseudo 1D measurements will always overestimate the voidage, and the more significant the overestimation, the more the voidage profile is concentrated in the core (assuming a monotonic profile).
a) 

![Diagram](image)

b) 

![Graph](image)
Figure K-5. Variation between average voidage from pressure-based (3D cylindrical integration) & radiation-based (1D integration) measurements

a) Example of the measurement setup, b) Example of voidage profiles & resulting average, c) Gas repartition center/periphery (50-50 area basis) as a function of the relative overestimation of bed voidage from radiation-based measurements

Figure K-6 presents the relative overestimation of bed voidage from radiation based measurements with the superficial gas velocity (with the radiation & pressure sensors located in a well-established zone). For the case without a baffle, this overestimation increases for cross-sectional superficial gas velocities ranging from 0 to 0.3 m·s⁻¹, which is indicative of voidage profile less and less flat. Then it reaches a local maximum near 0.45 m·s⁻¹ (near the transition velocity from a bubbling regime to a turbulent regime). Finally, a slow decrease of the overestimation is visible, indicative of flatter voidage profiles. For the case with a baffle (located at bank B), a similar profile is visible, with a local maximum also located near 0.45 m·s⁻¹ (near transition B/T) but a quicker increase and decrease around this maximum.
By using the correlation described by Equations (K.3) and (K.7), and the cross-sectional gas velocity profile, the average bed voidage $\varepsilon_{\text{bed,AVG}}$ and the pseudo 1D bed voidage $\varepsilon_{\text{bed,LIN}}$ can be expressed as a function of the bed height $z$. For the bed configuration with a ring baffle, the following assumptions are made: below the baffle location, the empirical coefficients used are the ones obtained for the bed without a baffle, and above the baffle location, the empirical coefficients used are the ones obtained for the bed with a baffle. Then the values of $\varepsilon_{\text{bed,AVG}}$ and $\varepsilon_{\text{bed,LIN}}$ are compared to obtain the relative overestimation of bed voidage from radiation based measurements with the bed height $z$. Figure K-7 presents this relative overestimation of bed voidage for the case with and without a baffle.
Figure K-7. Relative overestimation of bed voidage from radiation-based measurements with the bed height
a) Without baffle, b) With a baffle located below bank B
Figure K-7 shows that for the case without a baffle, the taper has an important impact on the voidage shape, strongly redistributing gas to the periphery. The voidage profile is the most concentrated to the center near a cross-sectional superficial gas velocity of 0.45 m·s\(^{-1}\) (near the transition velocity from the bubbling regime to the turbulent regime). For the case with a baffle (located at bank B), the effect of the taper is also clearly visible. In addition, the effect of the taper is also significant, with a strong redistribution to the periphery near the baffle tip and a stronger concentration near the core above the baffle. The voidage profile is still the most concentrated to the center near a cross-sectional superficial gas velocity of 0.45 m·s\(^{-1}\) (near the transition velocity from the bubbling regime to the turbulent regime). It should be noted that a relatively flat profile is present at the highest bed locations.

The last step consists in obtaining the correlation giving the voidage shape \(\varepsilon_{\text{bed}}(r/R)\) as a function of dimensionless radius (\(r/R\)). The correlation is constructed based on the typical shape of voidage radial profile (Song et al., 2004; Song et al., 2006). It also assumes that the emulsion voidage \(\varepsilon_{\text{emulsion}}\) is equal to the voidage at minimum fluidization, \(\varepsilon_{\text{mf}} = 0.42\).

\[
\varepsilon_{\text{bed}}(r/R) \left(\frac{r}{R}, u_G\right) = A \cdot \sum \left[ B \cdot \left| \frac{r}{R} \right| \right]^C + (1 - A) \cdot \sum \left[ D \cdot \left(1 - \left| \frac{r}{R} \right| \right) \right]^E + F \quad (K.8)
\]

with: A, B, C, D, E & F varying with \(u_G\)

Table K-3 presents the empirical coefficients obtained.

<table>
<thead>
<tr>
<th>(u_G) (m·s(^{-1}))</th>
<th>(0.00)</th>
<th>(0.01)</th>
<th>(0.02)</th>
<th>(0.03)</th>
<th>(0.04)</th>
<th>(0.05)</th>
<th>(0.06)</th>
<th>(0.07)</th>
<th>(0.08)</th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>0.04972</td>
<td>0.05000</td>
<td>0.04777</td>
<td>0.07215</td>
<td>0.13167</td>
<td>0.16702</td>
<td>0.19661</td>
<td>0.24016</td>
<td>0.27222</td>
</tr>
<tr>
<td>C</td>
<td>0.06084</td>
<td>0.07017</td>
<td>0.08961</td>
<td>0.01642</td>
<td>0.09642</td>
<td>0.03568</td>
<td>0.05095</td>
<td>0.04884</td>
<td>0.05634</td>
</tr>
<tr>
<td>D</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
</tr>
<tr>
<td>E</td>
<td>1.00669</td>
<td>0.99371</td>
<td>1.00668</td>
<td>1.00092</td>
<td>1.01939</td>
<td>1.02679</td>
<td>1.04085</td>
<td>1.05543</td>
<td>1.06682</td>
</tr>
<tr>
<td>F</td>
<td>1.00669</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
<td>1.00000</td>
</tr>
</tbody>
</table>

Table K-3. Parameters correlated for the voidage profiles.
3) Voidage maps

By combining the correlations obtained for the average bed voidage $\varepsilon_{\text{bed,AVG}}(z)$ and for the voidage shape $\varepsilon_{\text{bed}}(r)/\varepsilon_{\text{bed,AVG}}(z)$, voidage maps $\varepsilon_{\text{bed}}(r, z)$ such as the ones presented in Figure K-8 can be obtained.

In Figure K-8, for the case without a baffle, a relatively sharp voidage profile is visible at any height. This voidage profile starts to flatten in the higher zone of the bed. For the case with a baffle (located at bank B), a relatively voidage sharp profile is visible below the
baffle. On the other hand, a relatively flat profile is visible above the baffle. An overall drop in voidage is visible just above the taper location.

![Figure K-8. Bed voidage maps](image)

a) Without baffle, b) With a baffle located below bank B

References:


Appendix L: Detailed analysis of the contribution of each component of the liquid losses prediction model

1) Agglomerates drying only

This section presents model results when only the drying components of the model is used (with the Shrinking Core Model used to calculate the thermal drying).

1.1 Formation-to-stripper time distribution

![Graph of formation-to-stripper time distribution](image)

**Figure L-1.** Formation-to-stripper time distribution recorded with the RPT measurements
(Short Taper, Baffle @ B, No Flux-Tubes, $L_{\text{Nozzle}} = 0$ cm, $L_{\text{jet}} = 9$ cm,
$u_G \in [0.3; 0.9]$ m·s$^{-1}$, $F_S = 0.45$ kg·s$^{-1}$)

Figure L-1 presents the formation-to-stripper time distribution recorded with the RPT measurements. It should be noted that more than 50% of tracer-agglomerate travels are under $\tau = 1$ (mean solids residence time). This is explained as wet-agglomerates travel down faster than the average particle population. In addition, as expected, the lower the injection bank, the shorter the residence time. Finally, it should be pointed out that the figure presents logarithmic variation. Therefore, the contribution of Bank A (lowest
injection bank) is much worse than bank B, which is much worse than bank C, and so on. This highlights how wet-agglomerates bypassing is much worse with lower injection banks.

1.2 Fraction of agglomerates reaching stripper

\[ \rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3} \]
\[ (L/S)_0 = 0.31 \]
\[ D_{\text{Agg}} = 1 \text{ cm} \]

![Graph showing fraction of agglomerates reaching stripper with liquid](image)

**Figure L-2.** Fraction of agglomerates formed at Bank i reaching the stripper with liquid, using only the drying equations and assuming no agglomerate breakage, and no re-wetting

(Short Taper, Baffle @ B, No Flux-Tubes, \( L_{\text{Nozzle}} = 0 \text{ cm} \), \( L_{\text{jet}} = 9 \text{ cm} \), \( u_G \in [0.3; 0.9] \text{ m} \cdot \text{s}^{-1} \), \( F_S = 0.45 \text{ kg} \cdot \text{s}^{-1} \))

Figure L-2 shows the fraction of agglomerates formed at the injection Bank i reaching the stripper with liquid, using only the drying equations and assuming no agglomerate breakage, and no re-wetting. It should be noted that most of the agglomerates reaching the stripper with liquid are generated at the lowest injection bank (Bank A), confirming the importance of this bank in the liquid losses. On the other hand, there is only a small difference between 3 highest injection banks.
Figure L-3. Fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid, with the drying equations only, and assuming no agglomerate breakage, and no re-wetting for various initial agglomerates properties (Short Taper, Baffle @ B, No Flux-Tubes, \( L_{\text{Nozzle}} = 0 \text{ cm} \), \( L_{\text{jet}} = 9 \text{ cm} \), 
\[ u_G \in [0.3; 0.9] \text{ m} \cdot \text{s}^{-1}, \text{ } F_S = 0.45 \text{ kg} \cdot \text{s}^{-1} \])

Figure L-3 shows the fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid. Since only drying is considered, the time required for full drying \( t_c \) (defined in Chapter 2 – Section 2.4.2.2) is the controlling parameter. The bigger \( t_c \), the more agglomerates reach the stripper while carrying liquid (with a power-law increase). The value of \( t_c \) increases with \( \rho_{\text{Agg},0}, D_{\text{Agg}} \) & \( (L/S)_0 \), making the largest, densest, and wettest agglomerates more prone to carry-under liquid.

1.3 Liquid carried to stripper
Figure L-4. Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, with the drying equations only and assuming no agglomerate breakage, and no re-wetting

(Short Taper, Baffle @ B, No Flux-Tubes, $L_{\text{Nozzle}} = 0 \text{ cm}$, $L_{\text{jet}} = 9 \text{ cm}$,
\[ u_G \in [0.3; 0.9] \text{ m}\cdot\text{s}^{-1}, \quad F_S = 0.45 \text{ kg}\cdot\text{s}^{-1} \]

a) Fraction of liquid initially in agglomerate reaching stripper (from model),

b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)

Figure L-4 presents the fraction of liquid carried by agglomerates formed at Bank i towards the stripper. Figure L-4-a) presents the fraction of liquid initially carried by agglomerates formed at Bank i towards the stripper. Figure L-4-b) presents the fraction of liquid injected captured and carried by agglomerates formed at Bank i towards the stripper. As expected, most of the liquid carried by agglomerates towards the stripper is generated at the lowest bank (Bank A). This is due to a shorter travel-time leading to less drying. This effect is amplified by the fact that the lower gas velocity in the lower banks allows for more liquid to be captured by an agglomerate formed (for a given agglomerate size). Again, there is only a small difference between the three highest banks.

**a)**

\[ \rho_{Agg} = 1100 \text{ kg}\cdot\text{m}^{-3} \]
Figure L-5. Fraction of liquid carried by agglomerates towards the stripper (for the whole bed), with the drying equations only and assuming no agglomerate breakage, and no re-wetting, for various initial agglomerates properties (Short Taper, Baffle @ B, No Flux-Tubes, \(L_{\text{Nozzle}} = 0\) cm, \(L_{\text{jet}} = 9\) cm, \(u_G \in [0.3; 0.9] \text{ m}\cdot\text{s}^{-1}, F_S = 0.45 \text{ kg}\cdot\text{s}^{-1}\))

a) Fraction of liquid initially in agglomerate reaching stripper (from model),

b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)

Figure L-5 shows the fraction of liquid carried by agglomerates towards the stripper (for the whole bed). Figure L-5-a) presents the fraction of liquid initially carried by agglomerates towards the stripper (for the whole bed). Figure L-5-b) presents the fraction of liquid injected captured and carried by agglomerates towards the stripper (for the whole bed). Since only drying is considered, the bigger \(t_C\), the more liquid reaches the stripper. In addition, bigger agglomerates capture less of the liquid injected (as explained in Chapter 2 - Section 2.3.2.2). Therefore, agglomerates with a bigger \(t_C\) combined with a moderate \(D_{\text{Agg}}\) carry the most liquid to the stripper.
2) **Impact of agglomerates re-wetting**

This section presents model results when the drying components of the model is used (with the Shrinking Core Model used to calculate the thermal drying) combined with the re-wetting component of the model (with a coefficient of re-wetting $C_{\text{Rewet}} = 1$).

### 2.1 Fraction of agglomerates reaching stripper & location of (last) re-wetting

**a)**

![Graph showing fraction of agglomerates reaching stripper wet and rewetted.

**b)**

![Graph showing last rewet locations at different banks.}
Figure L-6. Fraction of agglomerates formed at Bank i reaching the stripper with liquid & rewetted, with the drying and re-wetting equations only, assuming no agglomerate breakage

(Short Taper, Baffle @ B, No Flux-Tubes, \( L_{\text{Nozzle}} = 0 \) cm, \( L_{\text{jet}} = 9 \) cm,
\( u_G \in [0.3; 0.9] \) m\( \cdot \)s\(^{-1}\), \( F_S = 0.45 \) kg\( \cdot \)s\(^{-1}\))

a) Fraction of agglomerates wet & rewetted at stripper for agglomerates formed at Bank i,
b) Location of last re-wetting bank for agglomerates formed at Bank i

Figure L-6 presents the fraction of agglomerates formed at Bank i reaching the stripper with liquid & the fraction of agglomerates formed at Bank i reaching the stripper after being re-wetted. First, by comparing Figure L-2 and Figure L-6, it can be noted that re-wetting increases significantly the number of agglomerates reaching the stripper with liquid (by a factor 10). In addition, in Figure L-6, it is visible that all banks have now similar contributions to the liquid losses. The higher banks contribute even more than the lower banks. A critical change is that virtually all agglomerates reaching stripper are re-wetted (> 99%). In addition, the last re-wetting occurs primarily in the two lowest banks (> 65%). Therefore, these banks act as a “relay”: by re-wetting the agglomerates, they regularly increase the required drying time, allowing agglomerates to reach the stripper with liquid.
Figure L-7. Fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid & rewetted, with the drying & re-wetting equations only, for various initial agglomerates properties

(Short Taper, Baffle @ B, No Flux-Tubes, L_{Nozzle} = 0 cm, L_{jet} = 9 cm,
\( u_G \in [0.3; 0.9] \text{ m} \cdot \text{s}^{-1}, F_S = 0.45 \text{ kg} \cdot \text{s}^{-1} \))

a) Fraction of agglomerates wet & rewetted at stripper for agglomerates formed at Bank i,

b) Location of last re-wetting bank for agglomerates formed at Bank i

Figure L-7 presents the fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid & the fraction of agglomerates formed (for the whole bed) reaching the stripper after being re-wetted. When comparing with Figure L-3, it can be noted that re-wetting significantly increases the number of agglomerates reaching the stripper with liquid. It should be noted that no increase is visible for agglomerates with a very small \( t_C \) (they have not enough time to move from their wetting to a potential re-wetting without drying first). Re-wetting has a moderate impact for agglomerates with a very high \( t_C \) (by a factor 3): this is because these agglomerates already had enough time to reach the stripper with liquid before drying. Finally, the most significant impact is for agglomerates with a medium \( t_C \) (by a factor 6). These are the agglomerates which could have dried before reaching stripper but are now re-wetted, contributing to liquid losses. It should also be
noted that the lower banks are the banks of the last re-wetting (with Bank A, the lowest bank, being the most frequent location of last re-wetting).

2.2 Liquid carried to stripper

\[ a) \]

\[ b) \]

\[ \rho_{Agg} = 1100 \text{ kg·m}^{-3} \]
\[ (L/S)_0 = 0.31 \]
\[ D_{Agg} = 1 \text{ cm} \]
**Figure L-8.** Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, with the drying and re-wetting equations only (Short Taper, Baffle @ B, No Flux-Tubes, L_{Nozzle} = 0 cm, L_{jet} = 9 cm, \( u_G \in [0.3; 0.9] \) m·s\(^{-1}\), \( F_S = 0.45 \) kg·s\(^{-1}\))

a) Fraction of liquid initially in agglomerate reaching stripper (from model).

b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)

Figure L-8 presents the fraction of liquid carried by agglomerates formed at Bank i towards the stripper. Figure L-8-a) presents the fraction of liquid initially carried by agglomerates formed at Bank i towards the stripper. Figure L-8-b) presents the fraction of liquid injected captured and carried by agglomerates formed at Bank i towards the stripper. When compared with Figure L-4, re-wetting significantly increases the amount of liquid carried by agglomerates towards the stripper. This is true for all injection banks. In addition, with re-wetting considered, all banks have similar contributions. The only difference comes from the variable amount of liquid captured by the agglomerates formed at different banks.

\[
\text{\( \rho_{Agg} = 1100 \text{ kg} \cdot \text{m}^{-3} \)}
\]

<table>
<thead>
<tr>
<th>Fraction of liquid transferred, from generated agglomerate to stripper</th>
<th>0.00001</th>
<th>0.0001</th>
<th>0.001</th>
<th>0.01</th>
<th>0.1</th>
<th>1</th>
</tr>
</thead>
<tbody>
<tr>
<td>(L/S)<em>0, D</em>{Agg} (cm) &amp; t_c (s)</td>
<td>0.10</td>
<td>0.31</td>
<td>0.62</td>
<td>0.5</td>
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</tr>
<tr>
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<td>65.4</td>
<td>238.2</td>
<td></td>
<td></td>
<td></td>
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</tbody>
</table>
Figure L-9. Fraction of liquid carried by agglomerates towards the stripper (for the whole bed), with the drying and re-wetting equations only, for various initial agglomerates properties (Short Taper, Baffle @ B, No Flux-Tubes, L_{Nozzle} = 0 cm, L_{jet} = 9 cm, $u_G \in [0.3; 0.9] \text{ m}\cdot\text{s}^{-1}$, $F_S = 0.45 \text{ kg}\cdot\text{s}^{-1}$)

a) Fraction of liquid initially in agglomerate reaching stripper (from model),
b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)

Figure L-9 presents the fraction of liquid carried by agglomerates towards the stripper (for the whole bed). Figure L-9-a) presents the fraction of liquid initially carried by agglomerates towards the stripper (for the whole bed). Figure L-9-b) presents the fraction of liquid injected captured and carried by agglomerates towards the stripper (for the whole bed). When compared with Figure L-5, re-wetting increases significantly the amount of liquid carried by agglomerates towards the stripper. The impact of the re-wetting addition is the most significant impact for agglomerates with a medium $t_c$ (factor of 40); For agglomerate with a high $t_c$, the change is more moderate (factor of 20). Finally, no impact
is visible for agglomerates of low $t_C$ (because these agglomerates get dry before a possible re-wetting).

3) **Impact of agglomerates shear destruction**

This section discusses the impact of the addition of the bed shear component in the model, with and without considering re-wetting at the same time.

**3.1 No re-wetting considered**

This section presents model results when the drying components of the model is used (with the Shrinking Core Model used to calculate the thermal drying) combined with the breakage components from the bed shear (with $M = 7$, and $d_b = 4.1$ cm).

**3.1.1 Fraction of agglomerates reaching stripper**

![Graph](image)

**Figure L-10.** Fraction of agglomerates formed at Bank i reaching the stripper with liquid, with the drying and shear destruction equations only

(Short Taper, Baffle @ B, No Flux-Tubes, $L_{Nozzle} = 0$ cm, $L_{jet} = 9$ cm, $u_G \in [0.3; 0.9]$ m·s$^{-1}$, $F_S = 0.45$ kg·s$^{-1}$)
Figure L-10 shows the fraction of agglomerates formed at Bank i reaching the stripper with liquid. When compared with Figure L-2, the inclusion of the breakage mechanism due to bed shear divides the number of agglomerates reaching the stripper with liquid by 2. Similarly to the model used only with the drying component, most of the agglomerates reaching the stripper with liquid are generated at the lowest bank (Bank A).

The controlling parameter used for the analysis of the breakage component is the simplified initial agglomerate yield strength $Br_{Agg,0}$ (presented in more details in Chapter 2 - Section 2.8). This parameter is based on the combination of:

- the agglomerate yield strength $\tau_{Agg} = f(\varepsilon_{Agg}, S_{Agg})$ defined in Chapter 2 - Section 2.5.1.1
- the critical agglomerate yield strength $\tau_{Agg}^* = f(D_{Agg}^2)$ defined in Chapter 2 - Section 2.5.1.2

These parameters are bundled in a simple parameter used to estimate the agglomerate probability to survive breakage for a given solid (coke) and liquid (bitumen), for a given bed shear $\gamma$ (for a given bed hydrodynamics) and a given reaction parameters ($T, P, \ldots$). The new parameter is controlled through three variables: $\varepsilon_{Agg}, S_{Agg} & D_{Agg}$. Finally, these variables are associated to three initial agglomerates parameters: $\rho_{Agg,0}, (L/S)_0 & D_{Agg}$.

$$Br_{Agg,0} = 1.0 \times 10^{-4} \left(1 - \frac{\varepsilon_{Agg,0}}{\varepsilon_{Agg,0}}\right) \frac{S_{Agg,0}}{D_{Agg}^2} \quad (L.1)$$

Figure L-11 presents the fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid. When compared with Figure L-3, the inclusion of the breakage mechanism due to bed shear reduces the amount of liquid carried by agglomerates towards the stripper. This reduction is more significant for agglomerates of low $Br_{Agg,0}$: these are the larger agglomerates. This is an interesting phenomenon as it shows how increased breakage rate can overcome the impact of the longer $t_C$ achieved by these larger and wetter agglomerates. Consequently, the agglomerate carrying the most liquid to the stripper become the agglomerates of medium size and liquid content.
Figure L-11. Fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid, with the drying and shear destruction equations only, for various initial agglomerates properties

(Short Taper, Baffle @ B, No Flux-Tubes, L_{Nozzle} = 0 cm, L_{jet} = 9 cm, 
\( u_G \in [0.3; 0.9] \text{ m} \cdot \text{s}^{-1}, F_S = 0.45 \text{ kg} \cdot \text{s}^{-1} \))

3.1.2 Liquid carried to stripper

Figure L-12 presents the fraction of liquid carried by agglomerates formed at Bank i towards the stripper. Figure L-12-a) presents the fraction of liquid initially carried by agglomerates formed at Bank i towards the stripper. Figure L-12-b) presents the fraction of liquid injected captured and carried by agglomerates formed at Bank i towards the stripper. When compared with Figure L-4, adding the breakage component due to bed shear decreases moderately the amount of liquid carried by agglomerates towards the stripper. This decrease is more significant for the higher banks (factor 3.5).
Figure L-12. Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, with the drying and shear destruction equations only
(Short Taper, Baffle @ B, No Flux-Tubes, \(L_{\text{Nozzle}} = 0\) cm, \(L_{\text{jet}} = 9\) cm, 
\(u_G \in [0.3; 0.9]\) m·s\(^{-1}\), \(F_\text{S} = 0.45\) kg·s\(^{-1}\))
a) Fraction of liquid initially in agglomerate reaching stripper (from model),
b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)
Figure L-13. Fraction of liquid carried by agglomerates towards the stripper (for the whole bed), with the drying and shear destruction equations only, for various initial agglomerates properties

(Short Taper, Baffle @ B, No Flux-Tubes, L_{Nozzle} = 0 cm, L_{jet} = 9 cm,
\quad u_G \in [0.3; 0.9] \text{ m}\cdot\text{s}^{-1}, F_S = 0.45 \text{ kg}\cdot\text{s}^{-1})

a) Fraction of liquid initially in agglomerate reaching stripper (from model),
b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)

Figure L-13 presents the fraction of liquid carried by agglomerates towards the stripper (for the whole bed). Figure L-13-a) presents the fraction of liquid initially carried by agglomerates towards the stripper (for the whole bed). Figure L-13-b) presents the fraction of liquid injected captured and carried by agglomerates towards the stripper (for the whole bed). When compared with Figure L-5, the inclusion of the breakage mechanism due to bed shear reduces the amount of liquid carried by agglomerates towards the stripper. This reduction is more significant for agglomerates of low Br_{Agg,0}: these are the larger agglomerates. This is interesting as it shows how an increased breakage rate can overcome the impact of the longer t_c achieved by these larger and wetter agglomerates. Consequently, the agglomerate carrying the most liquid to the stripper become the agglomerates of medium size and liquid content.

3.2 Re-wetting considered

This section presents model results when the drying components of the model is used (with the Shrinking Core Model used to calculate the thermal drying) combined with the breakage components from the bed shear (with M = 7, and d_b = 4.1 cm) and the re-wetting component of the model (with a coefficient of re-wetting C_{Rewet} = 1).

3.2.1 Fraction of agglomerates reaching stripper & location of (last) re-wetting
Figure L.14. Fraction of agglomerates formed at Bank i reaching the stripper with liquid & rewetted, with the drying, re-wetting and bed shear

(Short Taper, Baffle @ B, No Flux-Tubes, $L_{\text{Nozzle}} = 0$ cm, $L_{\text{jet}} = 9$ cm, $u_G \in [0.3; 0.9]$ m·s$^{-1}$, $F_S = 0.45$ kg·s$^{-1}$)
a) Fraction of agglomerates wet & rewetted at stripper for agglomerates formed at Bank i,

b) Location of last re-wetting bank for agglomerates formed at Bank i

Figure L-14 presents the fraction of agglomerates formed at Bank i reaching the stripper with liquid & the fraction of agglomerates formed at Bank i reaching the stripper after being re-wetted. When comparing with Figure L-2, the combined inclusion of breakage due to bed shear & re-wetting increases moderately the number of agglomerates reaching the stripper with liquid. Similarly to results obtained with only the drying components and with results obtained with drying component combined only with the breakage component due to bed shear, most of the agglomerates reaching the stripper with liquid generated at the lowest bank (Bank A). Virtually all agglomerates reaching stripper are rewetted (> 99%). These re-wetting act as a “relay”: they regularly increase the required drying time, allowing agglomerates to reach the stripper with liquid. The last re-wetting occurs primarily in the two lowest banks (> 80%).

Figure L-15 presents the fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid & the fraction of agglomerates formed (for the whole bed) reaching the stripper after being re-wetted. When comparing with Figure L-3, the combined inclusion of breakage due to bed shear & re-wetting leads to the “medium” agglomerates being the most likely to carry liquid towards the stripper. Indeed, small agglomerates undergo quick drying (characterized by a short $t_C$). On the other hand, big agglomerate (with a slow drying, characterized by a high $t_C$) are more prone to destruction due to shear (characterized by a low $Br_{Agg,0}$). Combined with the decreasing amount of liquid injected captured by larger agglomerates, this makes the “medium” agglomerates the biggest individual contributors to liquid losses to the stripper.

Regarding re-wetting, either all agglomerates are dried, or virtually all agglomerates reaching the stripper are rewetted (> 99%). The last re-wetting occurs primarily in the two lowest banks (> 80%).
Figure L-15. Fraction of agglomerates formed (for the whole bed) reaching the stripper with liquid & rewetted, with the drying, re-wetting and bed shear equations, for various initial agglomerates properties.
(Short Taper, Baffle @ B, No Flux-Tubes, L\textsubscript{Nozzle} = 0 cm, L\textsubscript{jet} = 9 cm, \n \quad u_G \in [0.3; 0.9] \text{ m}\cdot\text{s}^{-1}, F_S = 0.45 \text{ kg}\cdot\text{s}^{-1})

a) Fraction of agglomerates wet & rewetted at stripper for agglomerates formed at Bank i,

b) Location of last re-wetting bank for agglomerates formed at Bank i

### 3.2.2 Liquid carried to stripper

Figure L-16 presents the fraction of liquid carried by agglomerates formed at Bank i towards the stripper. Figure L-16-a) presents the fraction of liquid initially carried by agglomerates formed at Bank i towards the stripper. Figure L-16-b) presents the fraction of liquid injected captured and carried by agglomerates formed at Bank i towards the stripper. When compared with Figure L-4, the combined inclusion of breakage due to bed shear & re-wetting leads to the “medium” agglomerates being the most likely to carry liquid towards the stripper. Indeed, small agglomerates undergo quick drying (characterized by a short $t_c$). On the other hand, big agglomerate (with a slow drying, characterized by a high $t_c$) are more prone to destruction due to shear (characterized by a low $\text{Br}_{\text{Agg},0}$). Combined with the decreasing amount of liquid injected captured by larger agglomerates, this makes the “medium” agglomerates the biggest individual contributors to liquid losses to the stripper.

Figure L-17 presents the fraction of liquid carried by agglomerates towards the stripper (for the whole bed). Figure L-17-a) presents the fraction of liquid initially carried by agglomerates towards the stripper (for the whole bed). Figure L-17-b) presents the fraction of liquid injected captured and carried by agglomerates towards the stripper (for the whole bed). When compared with Figure L-5, after the addition of the combined re-wetting component and breakage component due to bed shear the “medium” agglomerates becomes the most likely to carry liquid towards the stripper. Indeed, small agglomerates undergo quick drying (characterized by a short $t_c$). On the other hand, big agglomerate (with a slow drying, characterized by a high $t_c$) are more prone to destruction due to shear (characterized by a low $\text{Br}_{\text{Agg},0}$). Combined with the decreasing amount of liquid injected captured by larger agglomerates, this makes the “medium” agglomerates the biggest individual contributors to liquid losses to the stripper.
Figure L-16. Fraction of liquid carried by agglomerates formed at Bank i towards the stripper, with the drying, re-wetting and shear destruction equations (Short Taper, Baffle @ B, No Flux-Tubes, L\textsubscript{Nozzle} = 0 cm, L\textsubscript{jet} = 9 cm, 
\(u_G \in [0.3; 0.9] \text{ m} \cdot \text{s}^{-1}, \ F_s = 0.45 \text{ kg} \cdot \text{s}^{-1}\))

a) Fraction of liquid initially in agglomerate reaching stripper (from model),

\(\rho_{\text{Agg}} = 1100 \text{ kg} \cdot \text{m}^{-3}\)

\([L/S]\_0 = 0.31\)

\(D_{\text{Agg}} = 1 \text{ cm}\)
b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)

\[ \rho_{Agg} = 1100 \text{ kg}\cdot\text{m}^{-3} \]

\begin{tabular}{cccccccc}
0.10 & 0.15 & 0.18 & 0.21 & 0.31 & 0.47 & 0.62 \\
0.5 & 0.625 & 0.7 & 0.75 & 1.00 & 1.25 & 1.50 \\
6.3 & 14.1 & 20.2 & 26.5 & 65.4 & 137.2 & 238.2 \\
2.981 & 2.194 & 1.808 & 1.611 & 0.906 & 0.534 & 0.334 \\
\end{tabular}

\[(L/S)_0 \& D_{Agg} (\text{cm})\]

**Figure L-17.** Fraction of liquid carried by agglomerates towards the stripper (for the whole bed), with the drying, re-wetting and shear destruction equations, for various initial...
agglomerates properties

(Short Taper, Baffle @ B, No Flux-Tubes, L_{Nozzle} = 0 cm, L_{jet} = 9 cm,
\[ u_G \in [0.3; 0.9] \, \text{m} \cdot \text{s}^{-1}, \, F_S = 0.45 \, \text{kg} \cdot \text{s}^{-1} \])

a) Fraction of liquid initially in agglomerate reaching stripper (from model),
b) Fraction of liquid injected trapped into original agglomerates (Li, 2016) & Fraction of liquid injected reaching stripper (through agglomerates)

4) **Competing effect of drying vs. shear breakage**

This section performs a brief analysis of the evolution of the drying and breakage characterizing parameters to identify the type of agglomerates the most likely to contribute to liquid carry-under.

\( a) \)
Figure L-18. Variation of the time required for full drying $t_c$ & the simplified initial agglomerate strength $B_{Agg,0}$, for various initial agglomerate parameters

a) Variable initial agglomerate density $\rho_{Agg,0}$, b) Variable initial Liquid-To-Solid Ratio $(L/S)_0$, c) Variable agglomerate diameter $D_{Agg}$
Figure L-18 shows the variation of the time required for full drying $t_c$ & the simplified initial agglomerate strength $B_{\text{Agg},0}$, for various initial agglomerate parameters. Regarding the time required for full drying $t_c$, it increases (slower drying) with: larger $\rho_{\text{Agg},0}$, larger $(L/S)_0$ and larger $D_{\text{Agg},0}$. Regarding the simplified initial agglomerate strength $B_{\text{Agg},0}$, it increases (less breakage) with: larger $\rho_{\text{Agg},0}$, smaller $D_{\text{Agg},0}$ and varies unevenly with $(L/S)_0$ (it goes through a local maximum). The “worst” agglomerates, i.e. the agglomerates responsible for the larger individual liquid carry-under are therefore, of high density, of moderate size and of moderate $(L/S)$.

5) **Agglomerates repartition in bed**

This section briefly discusses the axial repartition of wet-agglomerates in the bed, for the agglomerates reaching stripper with liquid and for agglomerates destroyed or dried before reaching the stripper.

Figure L-19 shows the repartition of agglomerates positions in-between formation and stripper or destruction, with drying, re-wetting & bed shear equations. For the agglomerates reaching stripper with liquid, it appears that they spend most of their time either: near the bank A or below (no matter where they were formed), or near the top two banks (which is the zone where the tracer spend the most time). For the agglomerate dried or destroyed before reaching the stripper, they spend most of their time above their production level. Overall, the agglomerates dried or destroyed before reaching the stripper spend the most time near the top two banks (which is the zone where the tracer spend the most time).
Figure L-19. Repartition of agglomerates positions in-between formation and stripper or destruction, with drying, re-wetting & bed shear equations
(Short Taper, Baffle @ B, No Flux-Tubes, L_{Nozzle} = 0 cm, L_{jet} = 9 cm,
\begin{align*}
u_G &\in [0.3; 0.9] \text{ m}\cdot\text{s}^{-1}, \quad F_S = 0.45 \text{ kg}\cdot\text{s}^{-1}, \quad \rho_{Agg} = 1100 \text{ kg}\cdot\text{m}^{-3}, \quad (L/S)_0 = 0.31, \quad D_{Agg} = 1\text{ cm}, \\
C_{Rewet} & = 1, \quad d_b = 4.0 \text{ cm}, \quad M = 7)
\end{align*}
a) Zones, b) Repartition of agglomerates reaching stripper with liquid, b) Repartition of agglomerates dried or destroyed

6) Location of agglomerate destruction

Figure L-20 presents the location of agglomerates destruction due to bed shear (it does not include agglomerates being fully dry without breaking). The colour scale represents the liquid content in the agglomerate at the moment of destruction, relative to the initial liquid content at formation.
Figure L-20. Location & liquid content of agglomerates destroyed

(Short Taper, Baffle @ B, No Flux-Tubes, L_{Nozzle} = 0 cm, L_{jet} = 9 cm, u_G \in [0.3; 0.9] \text{ m} \cdot \text{s}^{-1}, F_S = 0.45 \text{ kg} \cdot \text{s}^{-1}, \rho_{Agg} = 1100 \text{ kg} \cdot \text{m}^{-3}, (L/S)_0 = 0.31, D_{Agg} = 1\text{ cm}, C_{Rewet} = 1, d_b = 4.0 \text{ cm}, M = 7)

a) Formed at bank A, b) Formed at bank B, c) Formed at bank C, d) Formed at bank D, e) Formed at bank E

In Figure L-20, it appears that the agglomerate destruction occurs mostly near the top two banks (in region of highest shear). This corresponds to agglomerates with destroyed when they had moderately low liquid content (moderately strong). Significant destruction also happens within the taper region, but mostly for agglomerates with a very low liquid content (weak). In addition, for the parameter used for this figure, destruction occurs mostly for agglomerate already fairly dry \(((L/S)/(L/S)_0 < 0.35)\).
Appendix M: Liquid losses predicted by the model for multiple initial agglomerate parameters and alternative model parameters

1) Lateral redistribution results

Table M-1. Evolution of the amount of liquid carried-under to the bottom of the Fluid Coker, relative to the 5 banks case & for various initial agglomerate properties and various nozzle penetrations ($p_{\text{Agg,0}} = 1100$ kg·m$^{-3}$)

<table>
<thead>
<tr>
<th>(L/S)$_0$</th>
<th>$D_{\text{Agg}}$ (cm)</th>
<th>L$_{\text{IN}}$ (cm)</th>
<th>0.5</th>
<th>1</th>
<th>1.5</th>
<th>0.5</th>
<th>1</th>
<th>1.5</th>
</tr>
</thead>
<tbody>
<tr>
<td>A off</td>
<td>0.10</td>
<td>0.37</td>
<td>0.22</td>
<td>0.28</td>
<td>0.23</td>
<td>0.21</td>
<td>0.05</td>
<td>0.11</td>
</tr>
<tr>
<td></td>
<td>0.31</td>
<td>0.41</td>
<td>0.21</td>
<td>0.23</td>
<td>0.21</td>
<td>0.21</td>
<td>0.05</td>
<td>0.11</td>
</tr>
<tr>
<td></td>
<td>0.62</td>
<td>0.26</td>
<td>0.04</td>
<td>0.16</td>
<td>0.01</td>
<td>0.16</td>
<td>0.01</td>
<td>0.11</td>
</tr>
<tr>
<td>B off</td>
<td>0.10</td>
<td>0.98</td>
<td>1.17</td>
<td>1.19</td>
<td>1.05</td>
<td>1.07</td>
<td>0.66</td>
<td>1.78</td>
</tr>
<tr>
<td></td>
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<td>0.77</td>
<td>0.92</td>
<td>1.16</td>
<td>1.07</td>
<td>0.66</td>
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<td>1.78</td>
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<tr>
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<td>0.92</td>
<td>1.28</td>
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<td>0.82</td>
<td>1.25</td>
<td>0.82</td>
<td>0.82</td>
</tr>
<tr>
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<td>0.99</td>
<td>0.97</td>
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<td>0.93</td>
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<td>0.94</td>
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<td>0.75</td>
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<td>1.32</td>
<td>1.12</td>
<td>1.03</td>
<td>0.94</td>
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</tr>
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<td>0.62</td>
<td>1.01</td>
<td>1.32</td>
<td>1.12</td>
<td>1.09</td>
<td>1.03</td>
<td>0.94</td>
<td>1.97</td>
</tr>
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<td>1.01</td>
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<td>1.15</td>
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<td>1.50</td>
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<td>1.43</td>
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<td>1.17</td>
</tr>
<tr>
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<td>2.00</td>
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<td>3.60</td>
<td>1.65</td>
<td>3.60</td>
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<td>1.15</td>
</tr>
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<td>1.15</td>
<td>1.24</td>
<td>2.09</td>
<td>1.15</td>
<td>1.91</td>
<td>2.30</td>
<td>1.15</td>
</tr>
</tbody>
</table>
Table M-2. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, relative to the 5 banks case & for various agglomerate densities and various nozzle penetrations ($D_{Agg,0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$)

<table>
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<th>1100</th>
<th>900</th>
<th>1100</th>
</tr>
</thead>
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</tr>
<tr>
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<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>2</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$D_{Agg} = 1.0 \text{ cm}$</td>
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<td></td>
<td></td>
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<tr>
<td>$(L/S)_0 = 0.31$</td>
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<td></td>
<td></td>
</tr>
<tr>
<td>$A_{off}$</td>
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<td>0.21</td>
<td>0.05</td>
<td>0.05</td>
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<tr>
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<td>0.92</td>
<td>0.58</td>
<td>0.66</td>
</tr>
<tr>
<td>$C_{off}$</td>
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<td>1.01</td>
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<tr>
<td>$D_{off}$</td>
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<td>1.30</td>
<td>1.30</td>
<td>1.43</td>
</tr>
<tr>
<td>$E_{off}$</td>
<td>1.33</td>
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</tr>
</tbody>
</table>

Table M-3. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, relative to the 5 banks case & for bubble diameter $d_b$ and various nozzle penetrations ($\rho_{Agg,0} = 1100 \text{ kg·m}^{-3}$, $D_{Agg,0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$)

<table>
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<th>$d_b$ (cm)</th>
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<td></td>
<td></td>
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</tr>
<tr>
<td>2</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$D_{Agg} = 0.5 \text{ cm}$</td>
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<td></td>
<td></td>
</tr>
<tr>
<td>$(L/S)_0 = 0.10$</td>
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<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$A_{off}$</td>
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<td>0.28</td>
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<tr>
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<td>1.19</td>
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<td>3.60</td>
<td>2.69</td>
</tr>
<tr>
<td>$D_{Agg} = 1.0 \text{ cm}$</td>
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<tr>
<td>$(L/S)_0 = 0.31$</td>
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<td></td>
</tr>
<tr>
<td>$A_{off}$</td>
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<td>0.05</td>
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<tr>
<td>$B_{off}$</td>
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<td>0.53</td>
<td>0.97</td>
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<tr>
<td>$(L/S)_0 = 0.62$</td>
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</tr>
<tr>
<td>$A_{off}$</td>
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</tr>
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<td>1.21</td>
<td>1.15</td>
<td>2.12</td>
</tr>
</tbody>
</table>

2) Jet penetration results
Table M-4. Evolution of the amount of liquid carried-under to the bottom of the Fluid Coker, relative to various nozzle penetration & for various initial agglomerate properties and various redistribution profiles ($\rho_{\text{Agg},0} = 1100 \text{ kg}\cdot\text{m}^{-3}$)

<table>
<thead>
<tr>
<th>$D_{\text{Agg}}$ (cm)</th>
<th>$\langle L/S \rangle_0$</th>
<th>$0 \rightarrow 1$</th>
<th>$1 \rightarrow 2$</th>
<th>$2 \rightarrow 3$</th>
<th>$3 \rightarrow 0/2$</th>
</tr>
</thead>
<tbody>
<tr>
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<td>1.15</td>
<td>1.42</td>
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<td>6.27</td>
</tr>
<tr>
<td>1</td>
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<td>2.49</td>
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</tr>
<tr>
<td>1</td>
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<td>0.99</td>
<td>1.31</td>
<td>1.13</td>
<td>5.32</td>
</tr>
<tr>
<td>0.5</td>
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</tr>
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</tr>
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<td>1.5</td>
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<td>0.21</td>
<td>0.17</td>
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</table>

Table M-5. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, relative to various nozzle penetration & for various agglomerate densities and various redistribution profiles ($D_{\text{Agg},0} = 1.0 \text{ cm}$, $\langle L/S \rangle_0 = 0.31$)

<table>
<thead>
<tr>
<th>$\rho_{\text{Agg}}$ (kg/m$^3$)</th>
<th>$D_{\text{Agg}}$ (cm)</th>
<th>$\langle L/S \rangle_0$</th>
<th>$0 \rightarrow 1$</th>
<th>$1 \rightarrow 2$</th>
<th>$2 \rightarrow 3$</th>
<th>$3 \rightarrow 0/2$</th>
</tr>
</thead>
<tbody>
<tr>
<td>5 Banks</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1100</td>
<td>1</td>
<td>0.3</td>
<td>1.15</td>
<td>1.42</td>
<td>1.24</td>
<td>6.27</td>
</tr>
<tr>
<td>900</td>
<td>1</td>
<td>0.3</td>
<td>0.96</td>
<td>1.61</td>
<td>1.10</td>
<td>6.76</td>
</tr>
<tr>
<td>4 Banks (A off)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1100</td>
<td>1</td>
<td>0.3</td>
<td>0.83</td>
<td>0.33</td>
<td>0.06</td>
<td>26.29</td>
</tr>
<tr>
<td>900</td>
<td>1</td>
<td>0.3</td>
<td>0.79</td>
<td>0.76</td>
<td>0.05</td>
<td>40.05</td>
</tr>
</tbody>
</table>
Table M-6. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, relative to various nozzle penetration & for various bubble diameter $d_b$ and various redistribution profiles ($\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{Agg,0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$)

<table>
<thead>
<tr>
<th>$D_{Agg} \text{ (cm)}$</th>
<th>$(L/S)_0$</th>
<th>0 → 1</th>
<th>0 → 2</th>
<th>0 → 3</th>
<th>0 → 0/2</th>
<th>0 → 1</th>
<th>0 → 2</th>
<th>0 → 3</th>
<th>0 → 0/2</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>0.31</td>
<td>1.15</td>
<td>1.42</td>
<td>1.24</td>
<td>6.27</td>
<td>1.11</td>
<td>1.13</td>
<td>0.49</td>
<td>1.60</td>
</tr>
<tr>
<td>0.5</td>
<td>0.1</td>
<td>1.09</td>
<td>1.64</td>
<td>1.60</td>
<td>17.86</td>
<td>0.82</td>
<td>1.97</td>
<td>1.72</td>
<td>17.00</td>
</tr>
<tr>
<td>1.5</td>
<td>0.62</td>
<td>0.78</td>
<td>0.21</td>
<td>0.17</td>
<td>1.18</td>
<td>0.64</td>
<td>0.27</td>
<td>0.14</td>
<td>1.03</td>
</tr>
<tr>
<td>2</td>
<td>0.31</td>
<td>0.83</td>
<td>0.33</td>
<td>0.06</td>
<td>26.29</td>
<td>0.98</td>
<td>1.11</td>
<td>0.39</td>
<td>4.65</td>
</tr>
<tr>
<td>0.5</td>
<td>0.1</td>
<td>0.84</td>
<td>0.80</td>
<td>0.17</td>
<td>9.81</td>
<td>0.84</td>
<td>0.80</td>
<td>0.17</td>
<td>10.04</td>
</tr>
<tr>
<td>1.5</td>
<td>0.62</td>
<td>0.19</td>
<td>0.07</td>
<td>0.02</td>
<td>4.60</td>
<td>0.62</td>
<td>0.11</td>
<td>0.01</td>
<td>2.02</td>
</tr>
</tbody>
</table>

3) Single & multiple baffle results

3.1 Baffle(s) addition

Table M-7. Evolution of the amount of liquid carried-under to the bottom of the Fluid Coker, relative to the case without baffle & for various initial agglomerate properties and various nozzle penetrations ($\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $L_{IN} = 0 \text{ cm}$)

<table>
<thead>
<tr>
<th>$D_{Agg} \text{ (cm)}$</th>
<th>$(L/S)_0$</th>
<th>0.5</th>
<th>1</th>
<th>1.5</th>
</tr>
</thead>
<tbody>
<tr>
<td>5 banks</td>
<td>1 baffle @ B (No FT)</td>
<td>0.10</td>
<td>0.59</td>
<td>0.28</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.31</td>
<td>0.55</td>
<td>0.47</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.62</td>
<td>0.54</td>
<td>0.92</td>
</tr>
<tr>
<td></td>
<td>1 baffle @ B (FT)</td>
<td>0.10</td>
<td>0.52</td>
<td>0.27</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.31</td>
<td>0.49</td>
<td>0.45</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.62</td>
<td>0.45</td>
<td>0.50</td>
</tr>
<tr>
<td></td>
<td>4 baffles @ B,C,D,E (No FT)</td>
<td>0.10</td>
<td>0.49</td>
<td>0.19</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.31</td>
<td>0.47</td>
<td>0.43</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.62</td>
<td>0.44</td>
<td>0.75</td>
</tr>
<tr>
<td>4 banks (A off)</td>
<td>1 baffle @ B (No FT)</td>
<td>0.10</td>
<td>0.05</td>
<td>0.28</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.31</td>
<td>0.08</td>
<td>0.26</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.62</td>
<td>0.28</td>
<td>1.09</td>
</tr>
<tr>
<td></td>
<td>1 baffle @ B (FT)</td>
<td>0.10</td>
<td>0.04</td>
<td>0.27</td>
</tr>
<tr>
<td></td>
<td></td>
<td>0.31</td>
<td>0.08</td>
<td>0.26</td>
</tr>
</tbody>
</table>
Table M-8. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, relative to the case without baffle & for various agglomerate densities and various nozzle penetrations ($D_{Agg,0} = 1.0$ cm, $(L/S)_0 = 0.31$, $LIN = 0$ cm)

<table>
<thead>
<tr>
<th>$\rho_{Agg}$ (kg·m$^{-3}$)</th>
<th>900</th>
<th>1100</th>
</tr>
</thead>
</table>
| 5 banks
| $D_{Agg} = 1.0$ cm $(L/S)_0 = 0.31$ |
| 1 baffle @ B (No FT) | 0.65 | 0.47 |
| 1 baffle @ B (FT) | 0.41 | 0.45 |
| 4 baffles @ B,C,D,E (No FT) | 0.40 | 0.43 |
| 4 banks
(A off) |
| 1 baffle @ B (No FT) | 0.39 | 0.26 |
| 1 baffle @ B (FT) | 0.26 | 0.26 |
| 4 baffles @ B,C,D,E (No FT) | 0.16 | 0.17 |

Table M-9. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, relative to the case without baffle & for bubble diameter $d_b$ and various nozzle penetrations

($\rho_{Agg,0} = 1100$ kg·m$^{-3}$, $D_{Agg,0} = 1.0$ cm, $(L/S)_0 = 0.31$, $LIN = 0$ cm)

<table>
<thead>
<tr>
<th>$d_b$ (cm)</th>
<th>4.1</th>
<th>5</th>
</tr>
</thead>
</table>
| 5 banks
| $D_{Agg} = 0.5$ cm $(L/S)_0 = 0.10$ |
| 1 baffle @ B (No FT) | 0.59 | 0.70 |
| 1 baffle @ B (FT) | 0.52 | 0.60 |
| 4 baffles @ B,C,D,E (No FT) | 0.49 | 0.58 |
| 4 banks
(A off) |
| 1 baffle @ B (No FT) | 0.05 | 0.07 |
| 1 baffle @ B (FT) | 0.04 | 0.05 |
| 4 baffles @ B,C,D,E (No FT) | 0.03 | 0.08 |
| $D_{Agg} = 1.0$ cm $(L/S)_0 = 0.31$ |
| 1 baffle @ B (No FT) | 0.47 | 0.52 |
| 1 baffle @ B (FT) | 0.45 | 0.51 |
| 4 baffles @ B,C,D,E (No FT) | 0.43 | 0.50 |
| 4 banks
(A off) |
| 1 baffle @ B (No FT) | 0.26 | 0.31 |
| 1 baffle @ B (FT) | 0.26 | 0.29 |
### 3.2 Presence/absence of flux-tubes

**Table M-10.** Evolution of the amount of liquid carried-under to the bottom of the Fluid Coker, with one baffle at bank B. Addition of flux-tubes relative to the case without flux-tubes for various initial agglomerate properties 

\[(\rho_{\text{Agg},0} = 1100 \text{ kg·m}^{-3}, L_{\text{IN}} = 0 \text{ cm})\]

<table>
<thead>
<tr>
<th>(D_{\text{Agg}} ) (cm)</th>
<th>((L/S)_0)</th>
<th>0.5</th>
<th>1</th>
<th>1.5</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>5 banks</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>4 baffles @ B,C,D,E (No FT)</td>
<td>0.17</td>
<td>0.23</td>
<td></td>
</tr>
<tr>
<td></td>
<td>1 baffle @ B (No FT)</td>
<td>0.92</td>
<td>1.06</td>
<td></td>
</tr>
<tr>
<td></td>
<td>1 baffle @ B (FT)</td>
<td>0.75</td>
<td>0.86</td>
<td></td>
</tr>
<tr>
<td></td>
<td>4 baffles @ B,C,D,E (No FT)</td>
<td>0.75</td>
<td>0.87</td>
<td></td>
</tr>
<tr>
<td>4 banks (A off)</td>
<td>1 baffle @ B (No FT)</td>
<td>1.09</td>
<td>1.15</td>
<td></td>
</tr>
<tr>
<td></td>
<td>1 baffle @ B (FT)</td>
<td>1.23</td>
<td>1.26</td>
<td></td>
</tr>
<tr>
<td></td>
<td>4 baffles @ B,C,D,E (No FT)</td>
<td>1.23</td>
<td>1.27</td>
<td></td>
</tr>
</tbody>
</table>

\[D_{\text{Agg}} = 1.5 \text{ cm} \]
\[(L/S)_0 = 0.62\]
Table M-11. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, with one baffle at bank B. Addition of flux-tubes relative to the case without flux-tubes for various agglomerate densities 
\( (D_{Agg,0} = 1.0 \text{ cm}, (L/S)_0 = 0.31, L_{IN} = 0 \text{ cm}) \)

<table>
<thead>
<tr>
<th>( \rho_{Agg} ) (kg·m(^{-3}))</th>
<th>900</th>
<th>1100</th>
</tr>
</thead>
<tbody>
<tr>
<td>( D_{Agg} = 1.0 \text{ cm} ) (( L/S )_0 = 0.31) 5 banks</td>
<td>0.63</td>
<td>0.95</td>
</tr>
<tr>
<td>( D_{Agg} = 0.5 \text{ cm} ) (( L/S )_0 = 0.10) 4 banks (A off)</td>
<td>0.67</td>
<td>1.02</td>
</tr>
</tbody>
</table>

Table M-12. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, with one baffle at bank B. Addition of flux-tubes relative to the case without flux-tubes for various bubble diameter \( d_b \) 
\( (\rho_{Agg,0} = 1100 \text{ kg·m}^{-3}, D_{Agg,0} = 1.0 \text{ cm}, (L/S)_0 = 0.31, L_{IN} = 0 \text{ cm}) \)

<table>
<thead>
<tr>
<th>( d_b ) (cm)</th>
<th>4.1</th>
<th>5</th>
</tr>
</thead>
<tbody>
<tr>
<td>( D_{Agg} = 0.5 \text{ cm} ) (( L/S )_0 = 0.10) 5 banks</td>
<td>0.88</td>
<td>0.97</td>
</tr>
<tr>
<td>( D_{Agg} = 1.0 \text{ cm} ) (( L/S )_0 = 0.31) 4 banks (A off)</td>
<td>0.83</td>
<td>0.91</td>
</tr>
<tr>
<td>( D_{Agg} = 1.5 \text{ cm} ) (( L/S )_0 = 0.62) 5 banks</td>
<td>0.82</td>
<td>0.91</td>
</tr>
<tr>
<td>( D_{Agg} = 1.5 \text{ cm} ) (( L/S )_0 = 0.62) 4 banks (A off)</td>
<td>1.13</td>
<td>1.29</td>
</tr>
</tbody>
</table>

3.3 Impact of nozzles penetration

Table M-13. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, relative to various nozzle penetration & for agglomerate initial diameter and
liquid content, various redistribution profiles and various baffle configurations

\((\rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg,0}} = 1.0 \text{ cm}, (L/S)_0 = 0.31)\)

<table>
<thead>
<tr>
<th>(D_{\text{Agg}}(\text{cm}))</th>
<th>((L/S)_0)</th>
<th>1 Baf @ B (No FT)</th>
<th>1 Baf @ B (FT)</th>
</tr>
</thead>
<tbody>
<tr>
<td>(5 \text{ Banks})</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1</td>
<td>0.31</td>
<td>1.05</td>
<td>1.08</td>
</tr>
<tr>
<td>0.5</td>
<td>0.10</td>
<td>0.68</td>
<td>0.77</td>
</tr>
<tr>
<td>1.5</td>
<td>0.62</td>
<td>1.00</td>
<td>0.95</td>
</tr>
<tr>
<td>(4 \text{ Banks}) (A off)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1</td>
<td>0.31</td>
<td>1.13</td>
<td>1.26</td>
</tr>
<tr>
<td>0.5</td>
<td>0.10</td>
<td>1.06</td>
<td>1.17</td>
</tr>
<tr>
<td>1.5</td>
<td>0.62</td>
<td>1.10</td>
<td>1.19</td>
</tr>
</tbody>
</table>

**Table M-14.** Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, relative to various nozzle penetration & for various agglomerate densities, various redistribution profiles and various baffle configurations

\((\rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg,0}} = 1.0 \text{ cm}, (L/S)_0 = 0.31)\)

<table>
<thead>
<tr>
<th>(\rho_{\text{Agg}}(\text{kg/m}^3))</th>
<th>(D_{\text{Agg}}(\text{cm}))</th>
<th>((L/S)_0)</th>
<th>1 Baf @ B (No FT)</th>
<th>1 Baf @ B (FT)</th>
</tr>
</thead>
<tbody>
<tr>
<td>(5 \text{ Banks})</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1100</td>
<td>1</td>
<td>0.31</td>
<td>1.05</td>
<td>1.08</td>
</tr>
<tr>
<td>900</td>
<td>1</td>
<td>0.31</td>
<td>1.04</td>
<td>1.10</td>
</tr>
<tr>
<td>(4 \text{ Banks}) (A off)</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1100</td>
<td>1</td>
<td>0.31</td>
<td>1.13</td>
<td>1.26</td>
</tr>
<tr>
<td>900</td>
<td>1</td>
<td>0.31</td>
<td>1.01</td>
<td>1.12</td>
</tr>
</tbody>
</table>
Table M-15. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, relative to various nozzle penetration & for various bubble diameter $d_b$, various redistribution profiles and various baffle configurations

$(\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{Agg,0} = 1.0 \text{ cm}, (L/S)_0 = 0.31)$

<table>
<thead>
<tr>
<th>$D_{Agg} \text{ (cm)}$</th>
<th>$(L/S)_0$</th>
<th>0 → 1</th>
<th>0 → 2</th>
<th>0 → 1</th>
<th>0 → 2</th>
<th>0 → 1</th>
<th>0 → 2</th>
<th>0 → 1</th>
<th>0 → 2</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 Baf @ B (No FT)</td>
<td>1 Baf @ B (FT)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1 Baf @ B (No FT)</td>
<td>1 Baf @ B (FT)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$D_{Agg}$ (cm) (L/S)$_0$</td>
<td>0 → 1</td>
<td>0 → 2</td>
<td>0 → 1</td>
<td>0 → 2</td>
<td>0 → 1</td>
<td>0 → 2</td>
<td>0 → 1</td>
<td>0 → 2</td>
<td></td>
</tr>
<tr>
<td>1 Banks</td>
<td>1.05</td>
<td>1.08</td>
<td>1.20</td>
<td>1.25</td>
<td>1.18</td>
<td>1.21</td>
<td>1.37</td>
<td>1.42</td>
<td></td>
</tr>
<tr>
<td>0.5</td>
<td>0.68</td>
<td>0.77</td>
<td>0.97</td>
<td>0.98</td>
<td>0.78</td>
<td>0.90</td>
<td>0.90</td>
<td>0.98</td>
<td></td>
</tr>
<tr>
<td>1.5</td>
<td>1.00</td>
<td>0.95</td>
<td>0.69</td>
<td>0.69</td>
<td>1.12</td>
<td>1.06</td>
<td>0.79</td>
<td>0.79</td>
<td></td>
</tr>
<tr>
<td>5 Banks</td>
<td>1.13</td>
<td>1.26</td>
<td>1.25</td>
<td>1.44</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>0.5</td>
<td>1.06</td>
<td>1.17</td>
<td>1.19</td>
<td>1.32</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1.5</td>
<td>1.10</td>
<td>1.19</td>
<td>1.19</td>
<td>1.31</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>4 Banks (A off)</td>
<td>1.10</td>
<td>1.19</td>
<td>1.19</td>
<td>1.31</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>0.5</td>
<td>1.06</td>
<td>1.17</td>
<td>1.19</td>
<td>1.32</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>1.5</td>
<td>1.10</td>
<td>1.19</td>
<td>1.19</td>
<td>1.31</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

3.4 Adding more than one baffle

Table M-16. Evolution of the amount of liquid carried-under to the bottom of the Fluid Coker, when more than one baffle is added (at banks B, C, D, & E, no flux-tubes) relative the addition of a single baffle at bank B (No flux-tubes) & for various initial agglomerate properties

$(\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}, L_{IN} = 0 \text{ cm})$

<table>
<thead>
<tr>
<th>$D_{Agg}$ (cm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>$(L/S)_0$</td>
</tr>
<tr>
<td>5 banks</td>
</tr>
<tr>
<td>0.10</td>
</tr>
<tr>
<td>0.31</td>
</tr>
<tr>
<td>0.62</td>
</tr>
<tr>
<td>4 banks (A off)</td>
</tr>
<tr>
<td>0.10</td>
</tr>
<tr>
<td>0.31</td>
</tr>
<tr>
<td>0.62</td>
</tr>
</tbody>
</table>

Table M-17. Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, when more than one baffle is added (at banks B, C, D, & E, no flux-tubes) relative the addition of a single baffle at bank B (No flux-tubes) & for various
agglomerate densities

\( \left( D_{\text{Agg},0} = 1.0 \text{ cm}, \ (L/S)_0 = 0.31, \ L_{\text{IN}} = 0 \text{ cm} \right) \)

\[
\begin{array}{c|c|c}
\rho_{\text{Agg}} \text{ (kg·m}^{-3}\text{)} & 900 & 1100 \\
\hline
D_{\text{Agg}} = 1.0 \text{ cm} & \begin{array}{c|c|c}
5 \text{ banks} & 0.63 & 0.91 \\
4 \text{ banks} \ (A \text{ off}) & 0.40 & 0.65 \\
\end{array} \\
(L/S)_0 = 0.31 & \\
\end{array}
\]

**Table M-18.** Evolution of the amount of liquid carried under to the bottom of the Fluid Coker, when more than one baffle is added (at banks B, C, D, & E, no flux-tubes) relative the addition of a single baffle at bank B (No flux-tubes) & for various bubble diameter \(d_b\)

\( (\rho_{\text{Agg},0} = 1100 \text{ kg·m}^{-3}, \ D_{\text{Agg},0} = 1.0 \text{ cm}, \ (L/S)_0 = 0.31, \ L_{\text{IN}} = 0 \text{ cm}) \)

\[
\begin{array}{c|c|c}
d_b \text{ (cm)} & 4.1 & 5 \\
\hline
D_{\text{Agg}} = 0.5 \text{ cm} & \begin{array}{c|c|c}
5 \text{ banks} & 0.83 & 0.94 \\
4 \text{ banks} \ (A \text{ off}) & 0.62 & 0.70 \\
\end{array} \\
(L/S)_0 = 0.10 & \\
\hline
D_{\text{Agg}} = 1.0 \text{ cm} & \begin{array}{c|c|c}
5 \text{ banks} & 0.91 & 1.02 \\
4 \text{ banks} \ (A \text{ off}) & 0.65 & 0.73 \\
\end{array} \\
(L/S)_0 = 0.31 & \\
\hline
D_{\text{Agg}} = 1.5 \text{ cm} & \begin{array}{c|c|c}
5 \text{ banks} & 0.82 & 0.92 \\
4 \text{ banks} \ (A \text{ off}) & 1.13 & 1.15 \\
\end{array} \\
(L/S)_0 = 0.62 & \\
\end{array}
\]
Appendix N: Statistical analysis of the model results

1) Test of hypotheses and significance for problem involving small samples

The differences of Means is defined by Spiegel and Stephens (2017) as follow.

Suppose that two random samples of sizes \(N_1\) and \(N_2\) are drawn from normal populations whose standard deviations are equal (\(\sigma_1 = \sigma_2 = \sigma\)). Suppose further that these two samples have means given by \(\bar{X}_1\) and \(\bar{X}_2\) and standard deviations given by \(s_1\) and \(s_2\), respectively. To test the hypothesis \(H_0\) that the two samples come from the same population (i.e., \(\mu_1 = \mu_2\) as well as \(\sigma_1 = \sigma_2\)); we use the t score given by:

\[
t = \frac{\bar{X}_1 - \bar{X}_2}{\sigma \cdot \sqrt{\frac{1}{N_1} + \frac{1}{N_2}}}
\]

Where \(\sigma = \sqrt{\frac{N_1 s_1^2 + N_2 s_2^2}{N_1 + N_2 - 2}}\)

The distribution t used is the Student’s distribution \(t(\nu, p)\), with \(\nu = N_1 + N_2 - 2\) degrees of freedom. A two-tailed test is conducted with a selected level of confidence \(p\) (\(p = 99\%\)).

2) Results of the statistical analysis for the liquid losses predicted by the model

a. Lateral injection redistribution

<p>| Table N-1. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ((\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{Agg,0} = 1.0 \text{ cm}, (L/S)<em>0 = 0.31, C</em>{Rewet} = 1, d_b = 4.1\text{ cm}, M = 7)) |
|---|---|---|---|---|---|
| (L_{IN} = 0\text{ cm}) | <strong>REFERENCE</strong> | <strong>Comparison</strong> | **| | |
| | 5 Banks | 4 Banks (A off) | 4 Banks (B off) | 4 Banks (C off) | 4 Banks (D off) | 4 Banks (E off) |
| Number of samples | 5 | 6 | 2 | 2 | 3 | 2 |
| Average | 1.16E-03 | 1.59E-04 | 9.80E-04 | 1.04E-03 | 8.43E-04 | 1.03E-03 |
| Standard Deviation | 2.69E-04 | 3.17E-05 | 3.47E-06 | 2.21E-04 | 9.03E-05 | 4.91E-04 |
| (|t|) | 2.02E-04 | 2.69E-04 | 3.03E-04 | 2.54E-04 | 4.11E-04 |
| (t\text{STUDENT(99%)}) | 3.25 | 4.03 | 4.03 | 3.71 | 4.03 |</p>
<table>
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<tr>
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<th>TRUE</th>
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<th>FALSE</th>
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</thead>
</table>

<table>
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<th>L(_{IN}) = 1 cm</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
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<tbody>
<tr>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>3</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>1.28E-03</td>
<td>1.79E-04</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.88E-04</td>
<td>1.26E-07</td>
</tr>
<tr>
<td>(\sigma)</td>
<td></td>
<td>1.88E-04</td>
</tr>
<tr>
<td>(</td>
<td>t</td>
<td>)</td>
</tr>
<tr>
<td>(t_{\text{STUDENT}(99%)})</td>
<td>5.84</td>
<td></td>
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| Statistically different? | TRUE |

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<th>Comparison</th>
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<tbody>
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<td>5 Banks</td>
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<td>4 Banks (B off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>2</td>
<td>3</td>
</tr>
<tr>
<td>Average</td>
<td>1.41E-03</td>
<td>7.48E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.84E-04</td>
<td>4.15E-06</td>
</tr>
<tr>
<td>(\sigma)</td>
<td>1.51E-04</td>
<td>5.35E-04</td>
</tr>
<tr>
<td>(</td>
<td>t</td>
<td>)</td>
</tr>
<tr>
<td>(t_{\text{STUDENT}(99%)})</td>
<td>5.84</td>
<td>9.92</td>
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</tbody>
</table>

| Statistically different? | TRUE | FALSE | FALSE | FALSE | FALSE |

<table>
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<tr>
<th>L(_{IN}) = 3 cm</th>
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<th>Comparison</th>
</tr>
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<tbody>
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<td>5 Banks</td>
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<td></td>
</tr>
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<td>Number of samples</td>
<td>2</td>
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<td>1.21E-05</td>
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<td>Standard Deviation</td>
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<td>(\sigma)</td>
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<td>(</td>
<td>t</td>
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| Statistically different? | TRUE |
Table N-2. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg},0} = 0.5 \text{ cm}$, $\text{L/S}_0 = 0.31$, $\text{C}_{\text{Rewet}} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

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<th>Comparison</th>
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</thead>
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<td>4 Banks (A off)</td>
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<td>5</td>
</tr>
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<td>2.18E-04</td>
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<td>t</td>
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<td>4.03</td>
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$L_{\text{IN}} = 1 \text{ cm}$

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<tbody>
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<td>4 Banks (A off)</td>
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<td>Number of samples</td>
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</tr>
<tr>
<td>Average</td>
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<tr>
<td>Standard Deviation</td>
<td>1.51E-04</td>
</tr>
<tr>
<td>$\sigma$</td>
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</tr>
<tr>
<td>$</td>
<td>t</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
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$L_{\text{IN}} = 2 \text{ cm}$

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<tbody>
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<td>Number of samples</td>
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</tr>
<tr>
<td>Average</td>
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<tr>
<td>Standard Deviation</td>
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### Table N-3. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{Agg,0} = 1100 \text{ kg}\cdot\text{m}^{-3}$, $D_{Agg,0} = 1.5 \text{ cm}$, $(L/S)_0 = 0.31$, $C_{Rewet} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

<table>
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<th>Comparison</th>
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</thead>
<tbody>
<tr>
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<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>5</td>
<td>6</td>
</tr>
<tr>
<td>Average</td>
<td>1.55E-04</td>
<td>2.19E-05</td>
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<td>3.71E-05</td>
<td>1.11E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
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<td></td>
</tr>
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<td>t</td>
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<th>Comparison</th>
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<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>1.71E-04</td>
<td>3.35E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
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<td>1.19E-07</td>
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<tr>
<td>$\sigma$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT(99%)}}$</td>
<td>9.92</td>
<td></td>
</tr>
<tr>
<td>$L_{IN}$ = 2 cm</td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>-----------------</td>
<td>-----------</td>
<td>------------</td>
</tr>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
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<td>3</td>
</tr>
<tr>
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<td>7.31E-05</td>
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</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{STUDENT(99%)}$</td>
<td></td>
<td></td>
</tr>
<tr>
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<td>FALSE</td>
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<table>
<thead>
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<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
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<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
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<td>3</td>
</tr>
<tr>
<td>Average</td>
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<td>$</td>
<td>t</td>
<td>$</td>
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<tr>
<td>$t_{STUDENT(99%)}$</td>
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<td></td>
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<tr>
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<td></td>
</tr>
</tbody>
</table>

**Table N-4.** Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{Agg,0}$ = 1100 kg·m$^{-3}$, $D_{Agg,0}$ = 1.0 cm, ($L/S$)$_0$ = 0.10, $C_{Rewet}$ = 1, $d_b$ = 4.1 cm, $M$ = 7)

<table>
<thead>
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</tr>
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<tbody>
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<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
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<tr>
<td>Number of samples</td>
<td>5</td>
<td>6</td>
</tr>
<tr>
<td>Average</td>
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<tr>
<td>Standard Deviation</td>
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<td>9.67E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
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<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
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</table>

This table shows the statistical analysis of differences of means for various nozzles penetrations, comparing a reference with different setups. The results indicate whether the differences are statistically significant.
<table>
<thead>
<tr>
<th>$t_{\text{STUDENT}}(99%)$</th>
<th>3.25</th>
<th>4.03</th>
<th>4.03</th>
<th>3.71</th>
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<table>
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<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
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<tr>
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<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}}(99%)$</td>
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</tr>
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<td>FALSE</td>
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</table>

<table>
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<th>Comparison</th>
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<tbody>
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<td>5 Banks</td>
<td>4 Banks (A off)</td>
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<td>Number of samples</td>
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<td>3</td>
</tr>
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<td>2.33E-03</td>
<td>2.43E-04</td>
</tr>
<tr>
<td>Standard Deviation</td>
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<td>1.21E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>1.17E-04</td>
<td>3.33E-04</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}}(99%)$</td>
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<td>9.92</td>
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<tr>
<td>Statistically different?</td>
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<td>FALSE</td>
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<table>
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<th>Comparison</th>
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</thead>
<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>1.58E-03</td>
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<td>$\sigma$</td>
<td>1.44E-04</td>
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<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}}(99%)$</td>
<td>9.92</td>
<td></td>
</tr>
<tr>
<td>Statistically different?</td>
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</table>
Table N-5. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{\text{Agg},0} = 1100$ kg·m$^{-3}$, $D_{\text{Agg},0} = 1.0$ cm, $(L/S)_0 = 0.62$, $C_{\text{Rewet}} = 1$, $d_b = 4.1$ cm, $M = 7$)

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<th>Comparison</th>
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<tbody>
<tr>
<td>$L_{IN} = 0$ cm</td>
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<td>Number of samples</td>
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<td>1.12E-03</td>
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<td>Standard Deviation</td>
<td>1.33E-04</td>
<td>8.76E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>1.22E-04</td>
<td>1.33E-04</td>
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<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td>3.25</td>
<td>4.03</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>TRUE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

| $L_{IN} = 1$ cm | REFERENCE | Comparison |
| Number of samples | 2 | 2 |
| Average | 1.71E-03 | 2.36E-04 |
| Standard Deviation | 2.35E-05 | 8.35E-07 |
| $\sigma$ | 2.35E-05 |
| $|t|$ | 39.54 |
| $t_{\text{STUDENT}(99\%)}$ | 9.92 |
| Statistically different? | TRUE |

| $L_{IN} = 2$ cm | REFERENCE | Comparison |
| Number of samples | 2 | 3 | 2 | 3 | 2 | 2 |
| Average | 1.66E-03 | 3.50E-04 | 1.72E-03 | 1.79E-03 | 1.50E-03 | 2.32E-03 |
| Standard Deviation | 1.92E-04 | 1.80E-04 | 1.32E-03 | 1.06E-03 | 2.06E-04 | 8.93E-04 |
| $\sigma$ | 2.39E-04 | 1.33E-03 | 1.07E-03 | 2.82E-04 | 9.13E-04 |
| $|t|$ | 6.00 | 0.05 | 0.14 | 0.55 | 0.73 |
| $t_{\text{STUDENT}(99\%)}$ | 5.84 | 9.92 | 5.84 | 9.92 | 9.92 |
| Statistically different? | TRUE | FALSE | FALSE | FALSE | FALSE |
### Table N-6. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{\text{Agg,0}} = 1100$ kg·m$^{-3}$, $D_{\text{Agg,0}} = 0.5$ cm, $(L/S)_0 = 0.10$, $C_{\text{Rewet}} = 1$, $d_b = 4.1$ cm, $M = 7$)

<table>
<thead>
<tr>
<th>$L_{\text{IN}}$</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>$L_{\text{IN}} = 0$ cm</td>
<td>3</td>
<td>3</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>1.29E-03</td>
<td>2.97E-04</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>1.40E-04</td>
<td>9.65E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>1.48E-04</td>
<td>1.41E-04</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT(99%)}}$</td>
<td>4.60</td>
<td>5.84</td>
</tr>
<tr>
<td><strong>Statistically different?</strong></td>
<td>TRUE</td>
<td>FALSE</td>
</tr>
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### L$_{\text{IN}} = 1$ cm

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<th>Comparison</th>
</tr>
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<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>1.04E-03</td>
<td>2.06E-04</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>4.66E-05</td>
<td>7.30E-07</td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
<td>4.66E-05</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT(99%)}}$</td>
<td>9.92</td>
<td></td>
</tr>
</tbody>
</table>
**Statistically different?**

<table>
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<th><strong>Comparison</strong></th>
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</thead>
<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
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<tr>
<td>Number of samples</td>
<td>2</td>
<td>3</td>
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<td>1.82E-03</td>
<td>2.62E-04</td>
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<td>Standard Deviation</td>
<td>2.30E-04</td>
<td>2.67E-05</td>
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<tr>
<td>(\sigma)</td>
<td></td>
<td>1.90E-04</td>
</tr>
<tr>
<td>(</td>
<td>t</td>
<td>)</td>
</tr>
<tr>
<td>(t\text{STUDENT}(99%))</td>
<td></td>
<td>5.84</td>
</tr>
<tr>
<td><strong>Statistically different?</strong></td>
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<td>FALSE</td>
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</thead>
<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
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<tr>
<td>Number of samples</td>
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<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>1.86E-03</td>
<td>7.43E-05</td>
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<td>2.16E-06</td>
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<tr>
<td>(\sigma)</td>
<td></td>
<td>1.68E-04</td>
</tr>
<tr>
<td>(</td>
<td>t</td>
<td>)</td>
</tr>
<tr>
<td>(t\text{STUDENT}(99%))</td>
<td></td>
<td>9.92</td>
</tr>
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<td><strong>Statistically different?</strong></td>
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**Table N-7.** Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations \(\rho\text{Agg,0} = 1100 \text{ kg·m}^{-3}\), \(D\text{Agg,0} = 1.5\text{ cm}\), \((L/S)_0 = 0.62\), \(C\text{Rewet} = 1\), \(d_b = 4.1\text{ cm}\), \(M = 7\)

<table>
<thead>
<tr>
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<th><strong>Comparison</strong></th>
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</thead>
<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>5</td>
<td>5</td>
</tr>
<tr>
<td>Average</td>
<td>7.54E-05</td>
<td>1.54E-06</td>
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<tr>
<td>Standard Deviation</td>
<td>1.77E-05</td>
<td>5.22E-07</td>
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<tr>
<td>(\sigma)</td>
<td></td>
<td>1.40E-05</td>
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<tr>
<td>(</td>
<td>t</td>
<td>)</td>
</tr>
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### L\(_{\text{IN}} = 1 \text{ cm}\)

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<tbody>
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<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
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</tr>
<tr>
<td>Average</td>
<td>7.27E-05</td>
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<tr>
<td>Standard Deviation</td>
<td>1.63E-06</td>
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<td>(\sigma)</td>
<td>1.63E-06</td>
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<td>(</td>
<td>t</td>
</tr>
<tr>
<td>(t) (_{\text{STUDENT(99%)}})</td>
<td>9.92</td>
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**Statistically different?**

TRUE

### L\(_{\text{IN}} = 2 \text{ cm}\)

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<tbody>
<tr>
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<td>4 Banks (A off)</td>
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<tr>
<td>Number of samples</td>
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</tr>
<tr>
<td>Average</td>
<td>3.54E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.90E-06</td>
</tr>
<tr>
<td>(\sigma)</td>
<td>2.37E-06</td>
</tr>
<tr>
<td>(</td>
<td>t</td>
</tr>
<tr>
<td>(t) (_{\text{STUDENT(99%)}})</td>
<td>5.84</td>
</tr>
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</table>

**Statistically different?**

TRUE FALSE FALSE FALSE FALSE

### L\(_{\text{IN}} = 3 \text{ cm}\)

<table>
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<tbody>
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<td>4 Banks (A off)</td>
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<td>Number of samples</td>
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</tr>
<tr>
<td>Average</td>
<td>2.46E-05</td>
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<tr>
<td>Standard Deviation</td>
<td>2.26E-06</td>
</tr>
<tr>
<td>(\sigma)</td>
<td>1.85E-06</td>
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<tr>
<td>(</td>
<td>t</td>
</tr>
<tr>
<td>(t) (_{\text{STUDENT(99%)}})</td>
<td>5.84</td>
</tr>
</tbody>
</table>

**Statistically different?**

TRUE
Table N-8. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{\text{Agg,0}} = 900 \text{ kg·m}^{-3}$, $D_{\text{Agg,0}} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

<table>
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<tbody>
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<td></td>
<td>5 Banks</td>
<td>4 Banks</td>
</tr>
<tr>
<td>Number of samples</td>
<td>3</td>
<td>4</td>
</tr>
<tr>
<td>Average</td>
<td>8.38E-04</td>
<td>3.76E-05</td>
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<tr>
<td>Standard Deviation</td>
<td>2.36E-04</td>
<td>1.95E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
<td>1.84E-04</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT(99%)}}$</td>
<td>4.03</td>
<td>5.84</td>
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<td></td>
<td>5 Banks</td>
<td>4 Banks</td>
</tr>
<tr>
<td>Number of samples</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>6.88E-04</td>
<td>6.56E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.07E-05</td>
<td>2.33E-07</td>
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<td>$\sigma$</td>
<td></td>
<td>2.07E-05</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT(99%)}}$</td>
<td>9.92</td>
<td></td>
</tr>
<tr>
<td>Statistically different?</td>
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<table>
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<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks</td>
</tr>
<tr>
<td>Number of samples</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>1.42E-03</td>
<td>1.55E-04</td>
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<tr>
<td>Standard Deviation</td>
<td>7.02E-05</td>
<td>3.38E-05</td>
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<td>$\sigma$</td>
<td></td>
<td>7.79E-05</td>
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<tr>
<td>$</td>
<td>t</td>
<td>$</td>
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</table>
## Table N-9. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations \( (p_{\text{Agg},0} = 1100 \text{ kg·m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31, C_{\text{Rewet}} = 1.5, d_b = 4.1 \text{ cm}, M = 7) \)

<table>
<thead>
<tr>
<th>( L_{\text{IN}} )</th>
<th>( \text{REFERENCE} )</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td>( L_{\text{IN}} = 3 \text{ cm} )</td>
<td><strong>REFERENCE</strong></td>
<td><strong>Comparison</strong></td>
</tr>
<tr>
<td></td>
<td><strong>5 Banks</strong></td>
<td><strong>4 Banks (A off)</strong></td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>1.42E-03</td>
<td>6.03E-06</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>8.57E-05</td>
<td>2.47E-07</td>
</tr>
<tr>
<td><strong>( \sigma )</strong></td>
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<td>8.57E-05</td>
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<td>**</td>
<td>16.52</td>
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<th>( \text{REFERENCE} )</th>
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<td></td>
<td><strong>5 Banks</strong></td>
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<tr>
<td></td>
<td><strong>4 Banks (B off)</strong></td>
<td><strong>4 Banks (C off)</strong></td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
<td>5</td>
<td>6</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>2.55E-03</td>
<td>3.51E-04</td>
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<tr>
<td><strong>Standard Deviation</strong></td>
<td>5.92E-04</td>
<td>6.98E-05</td>
</tr>
<tr>
<td><strong>( \sigma )</strong></td>
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<td>8.17</td>
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<td><strong>( t_{\text{STUDENT(99%)}} )</strong></td>
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<td>2</td>
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Table N-10. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{Agg,0} = 1100$ kg·m$^{-3}$, $D_{Agg,0} = 1.0$ cm, $(L/S)_0 = 0.31,$ $C_{Rewet} = 1$, $d_b = 5.0$ cm, $M = 7$)

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<td>---</td>
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<td>4.03</td>
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<td>2</td>
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<td></td>
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<td>2</td>
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<td>t</td>
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<td>9.92</td>
<td>5.84</td>
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<td>2</td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>4.95E-03</td>
<td>5.36E-04</td>
<td></td>
</tr>
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<td>1.01E-05</td>
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</tr>
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<td>$\sigma$</td>
<td>4.17E-04</td>
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<td></td>
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<tr>
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<td>t</td>
<td>$</td>
<td>10.57</td>
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</table>
Table N-11. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{\text{Agg},0} = 1100 \text{ kg·m}^{-3}$, $D_{\text{Agg},0} = 0.5 \text{ cm}$, $(L/S)_0 = 0.10$, $C_{\text{Rewet}} = 1$, $d_b = 5.0 \text{ cm}$, $M = 7$)

<table>
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<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>5</td>
<td>5</td>
</tr>
<tr>
<td>Average</td>
<td>1.09E-03</td>
<td>1.66E-04</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.00E-04</td>
<td>3.08E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td></td>
<td></td>
</tr>
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<td>Statistically different?</td>
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<th>Comparison</th>
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<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>2</td>
<td>2</td>
</tr>
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<td>Average</td>
<td>1.06E-03</td>
<td>2.10E-04</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>7.30E-05</td>
<td>7.45E-07</td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td></td>
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<th>Comparison</th>
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<tbody>
<tr>
<td></td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>2</td>
<td>3</td>
</tr>
<tr>
<td>Average</td>
<td>2.27E-03</td>
<td>3.22E-04</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>3.42E-04</td>
<td>9.00E-05</td>
</tr>
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<td>$\sigma$</td>
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<td></td>
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<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

| Number of samples              | 5.84      | 9.92        | 5.84          | 9.92            | 9.92            | 9.92            |
### Table N-12. Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{\text{Agg.0}} = 1100 \text{ kg·m}^{-3}$, $D_{\text{Agg.0}} = 1.5 \text{ cm}$, $(L/S)_0 = 0.62$, $C_{\text{Rewet}} = 1$, $d_b = 5.0 \text{ cm}$, $M = 7$)

<table>
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<th>$\text{REFERENCE}$</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td>0 cm</td>
<td>5 Banks</td>
<td>4 Banks</td>
</tr>
<tr>
<td></td>
<td>$\text{(A off)}$</td>
<td>$\text{(E off)}$</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>_{\text{STUDENT}}(99%)$</td>
</tr>
<tr>
<td>$\text{Statistically different?}$</td>
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<td>FALSE</td>
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</table>

L$_{\text{IN}} = 0$ cm

<table>
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<th>$\text{REFERENCE}$</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 cm</td>
<td>5 Banks</td>
<td>4 Banks</td>
</tr>
<tr>
<td></td>
<td>$\text{(A off)}$</td>
<td>$\text{(E off)}$</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>_{\text{STUDENT}}(99%)$</td>
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<tr>
<td>$\text{Statistically different?}$</td>
<td>TRUE</td>
<td>FALSE</td>
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</table>
**Table N-13.** Statistical analysis (Differences of means) – Effect of the redistribution for various nozzles penetrations ($\rho_{\text{Agg},0} = 1100 \text{ kg m}^{-3}$, $D_{\text{Agg},0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1 \text{ cm}$, $M = 6$)

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<th>REFERENCE</th>
<th>Comparison</th>
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<tbody>
<tr>
<td>2 cm</td>
<td>5 Banks</td>
<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>6.34E-05</td>
<td>5.30E-06</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.38E-06</td>
<td>4.24E-07</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td>9.92</td>
<td>9.92</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>TRUE</td>
<td>FALSE</td>
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<table>
<thead>
<tr>
<th>3 cm</th>
<th>5 Banks</th>
<th>4 Banks (A off)</th>
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</thead>
<tbody>
<tr>
<td>Number of samples</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>6.23E-05</td>
<td>7.00E-07</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>3.01E-06</td>
<td>4.24E-07</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td>9.92</td>
<td></td>
</tr>
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<td>1.22E-03</td>
<td>1.67E-04</td>
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<td>3.33E-05</td>
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<td>$\sigma$</td>
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<td>t</td>
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**Statistically different?**

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<td>FALSE</td>
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<td>FALSE</td>
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### L\text{\textsubscript{IN}} = 1 cm

<table>
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<tbody>
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<td>4 Banks (A off)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>3</td>
</tr>
<tr>
<td>Average</td>
<td>1.34E-03</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.03E-04</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>2.03E-04</td>
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<td>$</td>
<td>t</td>
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<td>$t_{\text{STUDENT}(99%)}$</td>
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**Statistically different?**

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### L\text{\textsubscript{IN}} = 2 cm

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<tr>
<td>Average</td>
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<td>Standard Deviation</td>
<td>1.94E-04</td>
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<td>$\sigma$</td>
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<td>t</td>
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<td>$t_{\text{STUDENT}(99%)}$</td>
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**Statistically different?**

<table>
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### L\text{\textsubscript{IN}} = 3 cm

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<tbody>
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</tr>
<tr>
<td>Number of samples</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>2.05E-03</td>
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<td>Standard Deviation</td>
<td>4.70E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>4.72E-05</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
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**Statistically different?**

<table>
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</table>
b. Nozzle penetration

**Table N-14.** Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) \((\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31, C_{\text{Rewet}} = 1, d_b = 4.1 \text{ cm}, M = 7)\)

<table>
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<th><strong>REFERENCE</strong></th>
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<tr>
<td>Number of samples</td>
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<tr>
<td>Average</td>
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<td>2.65E-04</td>
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<td>(\sigma)</td>
<td>3.17E-04</td>
<td>2.93E-04</td>
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<td>(</td>
<td>t</td>
<td>)</td>
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<tr>
<td>(t_{\text{STUDENT}(99%)})</td>
<td>4.03</td>
<td>4.03</td>
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<th>Comparison</th>
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<tr>
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<td>0 cm</td>
<td>1 cm</td>
</tr>
<tr>
<td>Number of samples</td>
<td>6</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>1.59E-04</td>
<td>1.78E-04</td>
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<tr>
<td>Standard Deviation</td>
<td>3.17E-05</td>
<td>6.32E-07</td>
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<td>(\sigma)</td>
<td>3.17E-05</td>
<td>2.95E-05</td>
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<td>(</td>
<td>t</td>
<td>)</td>
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<tr>
<td>(t_{\text{STUDENT}(99%)})</td>
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**Table N-15.** Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) \((\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, D_{\text{Agg},0} = 0.5 \text{ cm}, (L/S)_0 = 0.31, C_{\text{Rewet}} = 1, d_b = 4.1 \text{ cm}, M = 7)\)

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<th>Comparison</th>
</tr>
</thead>
<tbody>
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<td>0 cm</td>
<td>1 cm</td>
</tr>
<tr>
<td>Number of samples</td>
<td>5</td>
<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>2.70E-03</td>
<td>3.09E-03</td>
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<tr>
<td>Standard Deviation</td>
<td>1.08E-03</td>
<td>1.51E-04</td>
</tr>
<tr>
<td>(\sigma)</td>
<td>1.08E-03</td>
<td>1.19E-03</td>
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<td>$</td>
<td>t</td>
<td>$</td>
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<tr>
<td>----------------------------------</td>
<td>-------</td>
<td>-------</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td>4.03</td>
<td>4.03</td>
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<table>
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<th><strong>Comparison</strong></th>
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<tr>
<td>Number of samples</td>
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<td>2</td>
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<tr>
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<td>8.50E-04</td>
<td>8.59E-04</td>
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<td>$t_{\text{STUDENT}(99%)}$</td>
<td>4.03</td>
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**Table N-16.** Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) ($\rho_{\text{Agg,0}} = 1100$ kg·m$^{-3}$, $D_{\text{Agg,0}} = 1.5$ cm, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1$ cm, $M = 7$)

<table>
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<th><strong>REFERENCE</strong></th>
<th><strong>Comparison</strong></th>
</tr>
</thead>
<tbody>
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<td>2</td>
</tr>
<tr>
<td>Average</td>
<td>1.55E-04</td>
<td>1.71E-04</td>
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<tr>
<td>Standard Deviation</td>
<td>3.71E-05</td>
<td>1.17E-05</td>
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<td>$\sigma$</td>
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<td>3.72E-05</td>
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<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td>4.03</td>
<td>4.03</td>
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<th><strong>Comparison</strong></th>
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</thead>
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<td>2</td>
</tr>
<tr>
<td>Average</td>
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</table>
### Table N-17. Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) ($\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{Agg,0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.10$, $C_{Rewet} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

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<tr>
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### Table N-18. Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) ($\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{Agg,0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.62$, $C_{Rewet} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

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<tr>
<td><strong>Standard Deviation</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$\sigma$</td>
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<td></td>
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<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td><strong>t STUDENT(99%)</strong></td>
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<td></td>
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### Table N-19. Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) (ρ_{Agg,0} = 1100 kg·m⁻³, D_{Agg,0} = 0.5 cm, (L/S)_0 = 0.10, C_{Rewet} = 1, d_b = 4.1 cm, M = 7)

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<td>1.92E-04</td>
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<td>8.53E-04</td>
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<tr>
<td>σ</td>
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<td>1.80E-04</td>
<td>1.94E-04</td>
<td>5.55E-04</td>
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<td>4.03</td>
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<tr>
<td>σ</td>
<td>5.62E-05</td>
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<td>σ</td>
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Table N-20. Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) ($\rho_{Agg,0} = 1100$ kg·m$^{-3}$, $D_{Agg,0} = 1.5$ cm, $(L/S)_0 = 0.62$, $C_{Rewet} = 1$, $d_b = 4.1$ cm, $M = 7$)

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<td>7.27E-05</td>
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<td>Standard Deviation</td>
<td>1.77E-05</td>
<td>1.63E-06</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>1.77E-05</td>
<td>1.78E-05</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
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<td>4.03</td>
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Table N-21. Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) ($\rho_{\text{Agg},0} = 900$ kg·m$^{-3}$, $D_{\text{Agg},0} = 1.0$ cm, $\text{(L/S)}_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1$ cm, $M = 7$)

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Table N-22. Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) ($\rho_{\text{Agg},0} = 1100$ kg·m$^{-3}$, $D_{\text{Agg},0} = 1.0$ cm, $\text{(L/S)}_0 = 0.31$, $C_{\text{Rewet}} = 1.5$, $d_b = 4.1$ cm, $M = 7$)

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<tr>
<td>Average</td>
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<td>$</td>
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<td>3.71</td>
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<td>4.03</td>
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<td>$</td>
<td>t</td>
<td>,$</td>
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<td>3.50</td>
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<td>3.29E-04</td>
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<td>t</td>
<td>,$</td>
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Table N-23. Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) ($\rho_{\text{Agg},0} = 1100$ kg m$^{-3}$, $D_{\text{Agg},0} = 1.0$ cm, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 5.0$ cm, $M = 7$)
**Table N-24.** Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) \((\rho_{Agg,0} = 1100 \text{ kg·m}^{-3}, D_{Agg,0} = 0.5 \text{ cm}, (L/S)_0 = 0.10, C_{Rewet} = 1, d_b = 5.0 \text{ cm}, M = 7)\)

<table>
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**Table N-25.** Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) \((\rho_{Agg,0} = 1100 \text{ kg·m}^{-3}, D_{Agg,0} = 1.5 \text{ cm}, (L/S)_0 = 0.62, C_{Rewet} = 1, d_b = 5.0 \text{ cm}, M = 7)\)

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<th>0-2 cm</th>
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<td>2</td>
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<td>t</td>
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Table N-26. Statistical analysis (Differences of means) – Effect of nozzle penetration with 5 banks & 4 banks (A off and redistributed) \( (\rho_{\text{Agg},0} = 1100 \text{ kg m}^{-3}, D_{\text{Agg},0} = 1.0 \text{ cm}, (L/S)_0 = 0.31, C_{\text{Rewet}} = 1, d_b = 4.1 \text{ cm}, M = 6) \)

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<td>6.23E-05</td>
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<td>3.99E-05</td>
<td>4.25E-05</td>
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<td>FALSE</td>
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<th>Comparison</th>
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<td>1.03E-05</td>
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<td>t</td>
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<td>4.03</td>
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<td>FALSE</td>
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<tr>
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</tr>
<tr>
<td>Average</td>
<td>1.22E-03</td>
<td>1.34E-03</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.82E-04</td>
<td>2.86E-04</td>
</tr>
<tr>
<td>( \sigma )</td>
<td>3.36E-04</td>
<td>3.08E-04</td>
</tr>
<tr>
<td>(</td>
<td>t</td>
<td>)</td>
</tr>
<tr>
<td>( t_{\text{STUDENT(99%)}} )</td>
<td>4.03</td>
<td>4.03</td>
</tr>
<tr>
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<td>FALSE</td>
</tr>
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</table>

<table>
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<th>Comparison</th>
</tr>
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<tr>
<td>Number of samples</td>
<td>6</td>
<td>2</td>
</tr>
</tbody>
</table>
### Table N-27. Statistical analysis (Differences of means) – Effect of single baffle addition
($\rho_{\text{Agg},0} = 1100$ kg·m$^{-3}$, $D_{\text{Agg},0} = 1.0$ cm, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1$ cm, $M = 7$)

<table>
<thead>
<tr>
<th></th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE Comparison</td>
<td>REFERENCE Comparison</td>
</tr>
<tr>
<td>L$_{IN}$ = 0 cm</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>6</td>
<td>10</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}}(99%)$</td>
<td>2.98</td>
<td>3.05</td>
</tr>
<tr>
<td>Average</td>
<td>9.71E-04</td>
<td>4.59E-04</td>
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<tr>
<td>Standard Deviation</td>
<td>2.15E-04</td>
<td>8.41E-05</td>
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<td>$\sigma$</td>
<td>1.58E-04</td>
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</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
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<tr>
<td>Statistically different?</td>
<td>TRUE</td>
<td>TRUE</td>
</tr>
</tbody>
</table>

### Table N-28. Statistical analysis (Differences of means) – Effect of single baffle addition
($\rho_{\text{Agg},0} = 1100$ kg·m$^{-3}$, $D_{\text{Agg},0} = 0.5$ cm, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1$ cm, $M = 7$)

<table>
<thead>
<tr>
<th></th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE Comparison</td>
<td>REFERENCE Comparison</td>
</tr>
<tr>
<td>L$_{IN}$ = 0 cm</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>4</td>
<td>9</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}}(99%)$</td>
<td>3.11</td>
<td>3.50</td>
</tr>
<tr>
<td></td>
<td>5 banks</td>
<td>4 banks (A off)</td>
</tr>
<tr>
<td>--------------------------</td>
<td>--------------------------------------</td>
<td>-----------------------------------------</td>
</tr>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td></td>
<td>No Baffle</td>
<td>I Baffle (at B - No FT)</td>
</tr>
<tr>
<td>L_{IN} = 0 cm</td>
<td>6</td>
<td>10</td>
</tr>
<tr>
<td>Number of samples</td>
<td></td>
<td></td>
</tr>
<tr>
<td>\text{ t}_{\text{student}(99%)}</td>
<td>2.98</td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>6.67E-05</td>
<td>6.14E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.39E-05</td>
<td>1.72E-05</td>
</tr>
<tr>
<td>\sigma</td>
<td>1.72E-05</td>
<td></td>
</tr>
<tr>
<td></td>
<td>t</td>
<td></td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

Table N-29. Statistical analysis (Differences of means) – Effect of single baffle addition
($\rho_{\text{Agg,0}} = 1100 \text{ kg·m}^{-3}, D_{\text{Agg,0}} = 1.5 \text{ cm}, (L/S)_0 = 0.31, C_{\text{Rewet}} = 1, d_b = 4.1 \text{ cm}, M = 7$)

<table>
<thead>
<tr>
<th></th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td></td>
<td>No Baffle</td>
<td>I Baffle (at B - No FT)</td>
</tr>
<tr>
<td>L_{IN} = 0 cm</td>
<td>6</td>
<td>10</td>
</tr>
<tr>
<td>Number of samples</td>
<td></td>
<td></td>
</tr>
<tr>
<td>\text{ t}_{\text{student}(99%)}</td>
<td>2.98</td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>9.25E-04</td>
<td>4.40E-04</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.05E-04</td>
<td>8.01E-05</td>
</tr>
</tbody>
</table>

Table N-30. Statistical analysis (Differences of means) – Effect of single baffle addition
($\rho_{\text{Agg,0}} = 1100 \text{ kg·m}^{-3}, D_{\text{Agg,0}} = 1.0 \text{ cm}, (L/S)_0 = 0.10, C_{\text{Rewet}} = 1, d_b = 4.1 \text{ cm}, M = 7$)
Table N-31. Statistical analysis (Differences of means) – Effect of single baffle addition
($\rho_{Agg,0} = 1100$ kg·m$^{-3}$, $D_{Agg,0} = 1.0$ cm, $(L/S)_0 = 0.62$, $C_{Rewet} = 1$, $d_b = 4.1$ cm, $M = 7$)

<table>
<thead>
<tr>
<th>L$_{IN} = 0$ cm</th>
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<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>No Baffle</td>
<td>5.07E-04</td>
<td>3.97E-05</td>
</tr>
<tr>
<td>1 Baffle (at B - No FT)</td>
<td>1.52E-04</td>
<td>3.99E-05</td>
</tr>
<tr>
<td>Number of samples</td>
<td>6</td>
<td>10</td>
</tr>
<tr>
<td>$t_{STUDENT(99%)}$</td>
<td>2.98</td>
<td>3.05</td>
</tr>
<tr>
<td>Average</td>
<td>1.07E-03</td>
<td>5.07E-04</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.38E-04</td>
<td>9.27E-05</td>
</tr>
</tbody>
</table>

Table N-32. Statistical analysis (Differences of means) – Effect of single baffle addition
($\rho_{Agg,0} = 1100$ kg·m$^{-3}$, $D_{Agg,0} = 0.5$ cm, $(L/S)_0 = 0.10$, $C_{Rewet} = 1$, $d_b = 4.1$ cm, $M = 7$)

<table>
<thead>
<tr>
<th>L$_{IN} = 0$ cm</th>
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<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>No Baffle</td>
<td>7.60E-04</td>
<td>9.65E-05</td>
</tr>
<tr>
<td>1 Baffle (at B - No FT)</td>
<td>1.29E-04</td>
<td>2.97E-04</td>
</tr>
<tr>
<td>Number of samples</td>
<td>4</td>
<td>9</td>
</tr>
<tr>
<td>$t_{STUDENT(99%)}$</td>
<td>3.11</td>
<td>3.50</td>
</tr>
<tr>
<td>Average</td>
<td>1.29E-03</td>
<td>7.60E-04</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.15E-04</td>
<td>1.29E-04</td>
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<tr>
<td>$\sigma$</td>
<td>1.36E-04</td>
<td>6.32E-05</td>
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<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
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<td>TRUE</td>
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<tr>
<td>-------------------------</td>
<td>------</td>
<td>------</td>
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</tbody>
</table>

Table N-33. Statistical analysis (Differences of means) – Effect of single baffle addition

\( \rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}, \ D_{\text{Agg},0} = 1.5 \text{ cm}, \ (L/S)_0 = 0.62, \ C_{\text{Rewet}} = 1, \ d_b = 4.1 \text{ cm}, \ M = 7 \)

<table>
<thead>
<tr>
<th>( L_{IN} = 0 \text{ cm} )</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE Comparison</td>
<td>REFERENCE Comparison</td>
</tr>
<tr>
<td>No Baffle</td>
<td>1 Baffle (at B - No FT)</td>
<td>No Baffle</td>
</tr>
<tr>
<td>Number of samples</td>
<td>6</td>
<td>10</td>
</tr>
<tr>
<td>\text{t}_{\text{STUDENT}}(99%)</td>
<td></td>
<td>3.05</td>
</tr>
<tr>
<td>Average</td>
<td>7.54E-05</td>
<td>6.94E-05</td>
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<tr>
<td>Standard Deviation</td>
<td>1.58E-05</td>
<td>1.96E-05</td>
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Table N-34. Statistical analysis (Differences of means) – Effect of single baffle addition

\( \rho_{\text{Agg},0} = 900 \text{ kg} \cdot \text{m}^{-3}, \ D_{\text{Agg},0} = 1.0 \text{ cm}, \ (L/S)_0 = 0.31, \ C_{\text{Rewet}} = 1, \ d_b = 4.1 \text{ cm}, \ M = 7 \)

<table>
<thead>
<tr>
<th>( L_{IN} = 0 \text{ cm} )</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE Comparison</td>
<td>REFERENCE Comparison</td>
</tr>
<tr>
<td>No Baffle</td>
<td>1 Baffle (at B - No FT)</td>
<td>No Baffle</td>
</tr>
<tr>
<td>Number of samples</td>
<td>6</td>
<td>6</td>
</tr>
<tr>
<td>\text{t}_{\text{STUDENT}}(99%)</td>
<td></td>
<td>3.25</td>
</tr>
<tr>
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<td>7.28E-04</td>
<td>3.53E-04</td>
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<tr>
<td>Standard Deviation</td>
<td>1.61E-04</td>
<td>4.03E-05</td>
</tr>
<tr>
<td>\sigma</td>
<td></td>
<td>1.29E-04</td>
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<tr>
<td>\text{t}</td>
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</table>
Table N-35. Statistical analysis (Differences of means) – Effect of single baffle addition
\((\rho_{\text{Agg,0}} = 1100 \text{ kg·m}^{-3}, D_{\text{Agg,0}} = 1.0 \text{ cm}, (L/S)_0 = 0.31, C_{\text{Rewet}} = 1, d_b = 5.0 \text{ cm}, M = 7)\)

<table>
<thead>
<tr>
<th>L_IN = 0 cm</th>
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<th>REFERENCE</th>
<th>Comparison</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td>No Baffle</td>
<td>1 Baffle (at B - No FT)</td>
<td>No Baffle</td>
<td>1 Baffle (at B - No FT)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>6</td>
<td>10</td>
<td>6</td>
<td>8</td>
<td></td>
<td></td>
</tr>
<tr>
<td>t_{\text{STUDENT}}(99%)</td>
<td>2.98</td>
<td></td>
<td>3.05</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>1.07E-03</td>
<td>5.09E-04</td>
<td>1.53E-04</td>
<td>3.97E-05</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.38E-04</td>
<td>9.32E-05</td>
<td>5.58E-05</td>
<td>5.56E-06</td>
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<td></td>
<td></td>
</tr>
</tbody>
</table>

Table N-36. Statistical analysis (Differences of means) – Effect of single baffle addition
\((\rho_{\text{Agg,0}} = 1100 \text{ kg·m}^{-3}, D_{\text{Agg,0}} = 0.5 \text{ cm}, (L/S)_0 = 0.10, C_{\text{Rewet}} = 1, d_b = 5.0 \text{ cm}, M = 7)\)

<table>
<thead>
<tr>
<th>L_IN = 0 cm</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
<th>REFERENCE</th>
<th>Comparison</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td>No Baffle</td>
<td>1 Baffle (at B - No FT)</td>
<td>No Baffle</td>
<td>1 Baffle (at B - No FT)</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>4</td>
<td>9</td>
<td>3</td>
<td>6</td>
<td></td>
<td></td>
</tr>
<tr>
<td>t_{\text{STUDENT}}(99%)</td>
<td>3.11</td>
<td></td>
<td>3.50</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>1.43E-03</td>
<td>8.38E-04</td>
<td>3.29E-04</td>
<td>1.53E-05</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.27E-04</td>
<td>1.43E-04</td>
<td>1.06E-04</td>
<td>2.52E-06</td>
<td></td>
<td></td>
</tr>
<tr>
<td>(\sigma)</td>
<td>1.43E-04</td>
<td></td>
<td>5.73E-05</td>
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<td></td>
</tr>
<tr>
<td></td>
<td>3.83</td>
<td></td>
<td>6.25</td>
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<td></td>
<td></td>
</tr>
<tr>
<td>Statistically different?</td>
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<td>TRUE</td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Table N-37. Statistical analysis (Differences of means) – Effect of single baffle addition
\((\rho_{\text{Agg,0}} = 1100 \text{ kg·m}^{-3}, D_{\text{Agg,0}} = 1.5 \text{ cm}, (L/S)_0 = 0.62, C_{\text{Rewet}} = 1, d_b = 5.0 \text{ cm}, M = 7)\)
ii. Presence/absence of flux tubes

**Table N-38.** Statistical analysis (Differences of means) – Effect of presence or absence of flux-tubes on a single baffle ($\rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg,0}} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

<table>
<thead>
<tr>
<th>$L_{IN} = 0 \text{ cm}$</th>
<th><strong>5 banks</strong></th>
<th><strong>4 banks (A off)</strong></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td><strong>No Baffle</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>1 Baffle</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>(at $B$ - No FT)</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
<td>6</td>
<td>10</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}}(99%)$</td>
<td>2.98</td>
<td></td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>8.30E-05</td>
<td>7.65E-05</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>1.74E-05</td>
<td>2.17E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>2.05E-05</td>
<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td><strong>Statistically different?</strong></td>
<td>FALSE</td>
<td></td>
</tr>
</tbody>
</table>

**Table N-39.** Statistical analysis (Differences of means) – Effect of presence or absence of flux-tubes on a single baffle ($\rho_{\text{Agg,0}} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg,0}} = 0.5 \text{ cm}$, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)
### Table N-40. Statistical analysis (Differences of means) – Effect of presence or absence of flux-tubes on a single baffle ($\rho_{\text{Agg}},0 = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg}},0 = 1.5 \text{ cm}$, $(L/S)_0 = 0.62$, $C_{\text{Rewet}} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

| Number of samples | $t_{\text{STUDENT}}(99\%)$ | Average | Standard Deviation | $\sigma$ | $|t|$ | Statistically different? |
|-------------------|-------------------------------|---------|--------------------|---------|-----|------------------------|
| $L_{\text{IN}} = 0 \text{ cm}$ | ![5 banks](REFERENCE) | ![5 banks](Comparison) | ![4 banks (A off)](REFERENCE) | ![4 banks (A off)](Comparison) |
| ![1 Baffle (at B - No FT)](1 Baffle) | ![1 Baffle (at B - FT)](1 Baffle) | ![1 Baffle (at B - No FT)](1 Baffle) | ![1 Baffle (at B - FT)](1 Baffle) |
| 9 | 3.17 | 7.60E-04 | 1.29E-04 | 1.69E-04 | 0.83 | FALSE |
| 6 | 3.36 | 6.66E-04 | 2.14E-04 | 1.99E-06 | 1.85 | FALSE |

### Table N-41. Statistical analysis (Differences of means) – Effect of presence or absence of flux-tubes on a single baffle ($\rho_{\text{Agg}},0 = 900 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg}},0 = 0.5 \text{ cm}$, $(L/S)_0 = 0.10$, $C_{\text{Rewet}} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

| Number of samples | $t_{\text{STUDENT}}(99\%)$ | Average | Standard Deviation | $\sigma$ | $|t|$ | Statistically different? |
|-------------------|-------------------------------|---------|--------------------|---------|-----|------------------------|
| $L_{\text{IN}} = 0 \text{ cm}$ | ![5 banks](REFERENCE) | ![5 banks](Comparison) | ![4 banks (A off)](REFERENCE) | ![4 banks (A off)](Comparison) |
| ![1 Baffle (at B - No FT)](1 Baffle) | ![1 Baffle (at B - FT)](1 Baffle) | ![1 Baffle (at B - No FT)](1 Baffle) | ![1 Baffle (at B - FT)](1 Baffle) |
| 10 | 3.05 | 6.94E-05 | 1.96E-05 | 1.91E-05 | 2.37 | FALSE |
| 4 | 3.11 | 4.27E-05 | 1.17E-05 | 8.65E-07 | 0.93 | FALSE |
Table N-42. Statistical analysis (Differences of means) – Effect of presence or absence of flux-tubes on a single baffle ($\rho_{Agg,0} = 1100$ kg·m$^{-3}$, $D_{Agg,0} = 1.0$ cm, $(L/S)_0 = 0.31$, $C_{Rewet} = 1$, $d_b = 5.0$ cm, $M = 7$)

<table>
<thead>
<tr>
<th>L$_{IN}$ = 0 cm</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>1 Baffle</td>
<td>1 Baffle (at B - No FT)</td>
<td>1 Baffle (at B - FT)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>6</td>
<td>3</td>
</tr>
<tr>
<td>$t_{STUDENT(99%)}$</td>
<td></td>
<td>3.50</td>
</tr>
<tr>
<td>Average</td>
<td>3.53E-04</td>
<td>3.28E-04</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>4.03E-05</td>
<td>5.15E-05</td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
<td>5.03E-05</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td></td>
<td>FALSE</td>
</tr>
</tbody>
</table>

Table N-43. Statistical analysis (Differences of means) – Effect of presence or absence of flux-tubes on a single baffle ($\rho_{Agg,0} = 1100$ kg·m$^{-3}$, $D_{Agg,0} = 0.5$ cm, $(L/S)_0 = 0.10$, $C_{Rewet} = 1$, $d_b = 5.0$ cm, $M = 7$)

<table>
<thead>
<tr>
<th>L$_{IN}$ = 0 cm</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>1 Baffle</td>
<td>1 Baffle (at B - No FT)</td>
<td>1 Baffle (at B - FT)</td>
</tr>
<tr>
<td>Number of samples</td>
<td>10</td>
<td>4</td>
</tr>
<tr>
<td>$t_{STUDENT(99%)}$</td>
<td></td>
<td>3.05</td>
</tr>
<tr>
<td>Average</td>
<td>5.08E-04</td>
<td>4.80E-04</td>
</tr>
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<td>Standard Deviation</td>
<td>9.27E-05</td>
<td>1.52E-04</td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
<td>1.22E-04</td>
</tr>
<tr>
<td>$</td>
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<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td></td>
<td>FALSE</td>
</tr>
</tbody>
</table>
### Table N-44. Statistical analysis (Differences of means) – Effect of presence or absence of flux-tubes on a single baffle ($\rho_{\text{Agg},0} = 1100 \, \text{kg} \cdot \text{m}^{-3}$, $D_{\text{Agg},0} = 1.5 \, \text{cm}$, $(L/S)_0 = 0.62$, $C_{\text{Rewet}} = 1$, $d_b = 5.0 \, \text{cm}$, $M = 7$)

<table>
<thead>
<tr>
<th>$L_{\text{IN}} = 0 , \text{cm}$</th>
<th><strong>5 banks</strong></th>
<th><strong>4 banks (A off)</strong></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
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<td><strong>Comparison</strong></td>
</tr>
<tr>
<td></td>
<td>1 Baffle (at B - No FT)</td>
<td>1 Baffle (at B - FT)</td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
<td>9</td>
<td>3</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td></td>
<td>3.17</td>
</tr>
<tr>
<td><strong>Average</strong></td>
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<td>2.35E-04</td>
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<td>$\sigma$</td>
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<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td><strong>Statistically different?</strong></td>
<td>FALSE</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
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<th>$L_{\text{IN}} = 0 , \text{cm}$</th>
<th><strong>5 banks</strong></th>
<th><strong>4 banks (A off)</strong></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td><strong>REFERENCE</strong></td>
<td><strong>Comparison</strong></td>
</tr>
<tr>
<td></td>
<td>1 Baffle (at B - No FT)</td>
<td>1 Baffle (at B - FT)</td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
<td>10</td>
<td>4</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td></td>
<td>3.05</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>7.69E-05</td>
<td>4.72E-05</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>2.16E-05</td>
<td>1.30E-05</td>
</tr>
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<td>$\sigma$</td>
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</table>
iii. Nozzles penetration

**Table N-45.** Statistical analysis (Differences of means) – Effect of nozzle penetration with a single baffle (no flux-tubes), with 5 banks & 4 banks (A off and redistributed) ($\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{Agg,0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$, $C_{Rewet} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

<table>
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<tr>
<th>$L_{IN}$</th>
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<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0 cm</td>
<td>2</td>
<td>8</td>
</tr>
<tr>
<td>1 cm</td>
<td>4</td>
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<tr>
<td>2 cm</td>
<td>3.17</td>
<td>3.05</td>
</tr>
<tr>
<td>3.71</td>
<td>3.05</td>
<td>3.36</td>
</tr>
<tr>
<td>Average</td>
<td>4.59E-04</td>
<td>4.94E-04</td>
</tr>
<tr>
<td>4.84E-04</td>
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</tr>
<tr>
<td>$\sigma$</td>
<td>8.70E-05</td>
<td>1.75E-04</td>
</tr>
<tr>
<td>8.41E-05</td>
<td>4.98E-05</td>
<td>5.03E-06</td>
</tr>
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<td>5.30E-06</td>
<td>3.87E-04</td>
<td>4.79E-06</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>0.37</td>
<td>0.47</td>
<td>1.10</td>
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</table>

**Table N-46.** Statistical analysis (Differences of means) – Effect of nozzle penetration with a single baffle (no flux-tubes), with 5 banks & 4 banks (A off and redistributed) ($\rho_{Agg,0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{Agg,0} = 0.5 \text{ cm}$, $(L/S)_0 = 0.10$, $C_{Rewet} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)
### 1 baffle at bank B (NO flux-tubes)

<table>
<thead>
<tr>
<th>I Baffle (No FT)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
<th>REFERENCE</th>
<th>Comparison</th>
<th>REFERENCE</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
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<td></td>
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<td>$L_{IN} = 1,\text{cm}$</td>
<td>$L_{IN} = 2,\text{cm}$</td>
<td>$L_{IN} = 0,\text{cm}$</td>
</tr>
<tr>
<td>Number of samples</td>
<td>9</td>
<td>2</td>
<td>2</td>
<td>6</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td></td>
<td>3.25</td>
<td>3.25</td>
<td>3.71</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>7.60E-04</td>
<td>5.18E-04</td>
<td>5.89E-04</td>
<td>1.39E-05</td>
<td>1.47E-05</td>
<td></td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.29E-04</td>
<td>1.84E-06</td>
<td>2.34E-04</td>
<td>2.28E-06</td>
<td>6.65E-07</td>
<td></td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
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<td>1.70E-04</td>
<td>2.31E-06</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
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<td>1.29</td>
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<td>0.46</td>
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<td>FALSE</td>
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</tr>
</tbody>
</table>

### 1 baffle at bank B (Flux-tubes)

<table>
<thead>
<tr>
<th>I Baffle (No FT)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
<th>REFERENCE</th>
<th>Comparison</th>
<th>REFERENCE</th>
<th>Comparison</th>
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<tr>
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<td>$L_{IN} = 1,\text{cm}$</td>
<td>$L_{IN} = 2,\text{cm}$</td>
<td>$L_{IN} = 0,\text{cm}$</td>
</tr>
<tr>
<td>Number of samples</td>
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<td>2</td>
<td>2</td>
<td>3</td>
<td>2</td>
<td>2</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td></td>
<td>5.84</td>
<td>5.84</td>
<td>5.84</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>6.66E-04</td>
<td>6.45E-04</td>
<td>5.30E-04</td>
<td>1.15E-05</td>
<td>1.34E-05</td>
<td></td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.14E-04</td>
<td>2.87E-04</td>
<td>1.88E-06</td>
<td>3.48E-07</td>
<td>3.26E-06</td>
<td></td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
<td>3.17E-04</td>
<td>2.14E-04</td>
<td>2.69E-06</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
<td>0.07</td>
<td>0.70</td>
<td></td>
<td>0.78</td>
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</table>

### Table N-47. Statistical analysis (Differences of means) – Effect of nozzle penetration with a single baffle (no flux-tubes), with 5 banks & 4 banks (A off and redistributed) ($\rho_{agg,0} = 1100\,\text{kg}\cdot\text{m}^{-3}$, $D_{agg,0} = 1.5\,\text{cm}$, $(L/S)_0 = 0.62$, $C_{\text{Rewet}} = 1$, $d_b = 4.1\,\text{cm}$, $M = 7$)
<table>
<thead>
<tr>
<th>Number of samples</th>
<th>10</th>
<th>2</th>
<th>2</th>
<th>9</th>
<th>2</th>
</tr>
</thead>
<tbody>
<tr>
<td>$t_{\text{STUDENT}}$(99%)</td>
<td>3.17</td>
<td>3.17</td>
<td></td>
<td></td>
<td>3.25</td>
</tr>
<tr>
<td>Average</td>
<td>6.94E-05</td>
<td>6.91E-05</td>
<td>6.60E-05</td>
<td>1.41E-06</td>
<td>1.84E-06</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.96E-05</td>
<td>2.45E-07</td>
<td>1.34E-05</td>
<td>6.38E-07</td>
<td>8.31E-08</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>1.96E-05</td>
<td>2.05E-05</td>
<td></td>
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<td>6.40E-07</td>
</tr>
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<td>$</td>
<td>t</td>
<td>$</td>
<td>0.03</td>
<td>0.22</td>
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</tr>
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<td>Statistically different?</td>
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<td>FALSE</td>
<td></td>
<td></td>
<td>FALSE</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>1 Baffle (No FT)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td>$L_{\text{IN}}$</td>
<td>Comparison</td>
<td>$L_{\text{IN}}$</td>
</tr>
<tr>
<td>0 cm</td>
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<td></td>
<td>2 cm</td>
</tr>
<tr>
<td>Number of samples</td>
<td>4</td>
<td>2</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}}$(99%)</td>
<td>4.60</td>
<td>4.60</td>
</tr>
<tr>
<td>Average</td>
<td>4.27E-05</td>
<td>3.92E-05</td>
</tr>
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<td>4.99E-06</td>
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<td>1.22E-05</td>
<td>1.17E-05</td>
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<td>$</td>
<td>t</td>
<td>$</td>
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<tr>
<td>Statistically different?</td>
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<td>FALSE</td>
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</table>

Table N-48. Statistical analysis (Differences of means) – Effect of nozzle penetration with a single baffle (no flux-tubes), with 5 banks & 4 banks (A off and redistributed) ($\rho_{\text{Agg,0}} = 900$ kg·m$^{-3}$, $D_{\text{Agg,0}} = 1.0$ cm, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1$ cm, $M = 7$)
<table>
<thead>
<tr>
<th>Standard Deviation</th>
<th>4.03E-05</th>
<th>4.98E-05</th>
<th>1.75E-04</th>
<th>2.41E-06</th>
<th>3.38E-06</th>
</tr>
</thead>
<tbody>
<tr>
<td>σ</td>
<td>4.96E-05</td>
<td>1.29E-04</td>
<td>3.38E-06</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>t</td>
<td></td>
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<td>1.71</td>
<td>3.74</td>
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<td>FALSE</td>
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<td></td>
</tr>
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</table>

### 1 baffle at bank B (Flux-tubes)

<table>
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<th>1 Baffle (No FT)</th>
<th>Reference</th>
<th>Comparison</th>
<th>Reference</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
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<td>L_in = 1 cm</td>
<td>L_in = 2 cm</td>
<td>L_in = 0 cm</td>
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<td>Number of samples</td>
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<td>2</td>
<td>2</td>
<td>4</td>
</tr>
<tr>
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<td>5.84</td>
<td></td>
<td>4.60</td>
</tr>
<tr>
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<td>2.64E-04</td>
<td>2.73E-04</td>
<td>2.72E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
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<td>2.17E-04</td>
<td>2.24E-04</td>
<td>7.21E-06</td>
</tr>
<tr>
<td>σ</td>
<td>1.13E+00</td>
<td>1.07E+00</td>
<td></td>
<td>7.97E-06</td>
</tr>
<tr>
<td></td>
<td>t</td>
<td></td>
<td>1.13</td>
<td>1.07</td>
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<td>FALSE</td>
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</table>

### Table N-49. Statistical analysis (Differences of means) – Effect of nozzle penetration with a single baffle (no flux-tubes), with 5 banks & 4 banks (A off and redistributed) ($\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg},0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 5.0 \text{ cm}$, $M = 7$)

<table>
<thead>
<tr>
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<th>Reference</th>
<th>Comparison</th>
<th>Reference</th>
<th>Comparison</th>
</tr>
</thead>
<tbody>
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<td>L_in = 1 cm</td>
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<td>L_in = 0 cm</td>
</tr>
<tr>
<td>Number of samples</td>
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<td>2</td>
<td>4</td>
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</tr>
<tr>
<td>t_student(99%)</td>
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<td>3.05</td>
<td></td>
<td>3.36</td>
</tr>
<tr>
<td>Average</td>
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<td>5.36E-04</td>
<td>5.44E-04</td>
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</tr>
<tr>
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<td>5.51E-05</td>
<td>1.94E-04</td>
<td>5.56E-06</td>
</tr>
<tr>
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<td>1.40E-04</td>
<td></td>
<td>5.87E-06</td>
</tr>
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<td>t</td>
<td></td>
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<td>0.45</td>
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<td>FALSE</td>
<td>FALSE</td>
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</tr>
<tr>
<td>-------------------------</td>
<td>-------</td>
<td>-------</td>
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<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>1 baffle at bank B (Flux-tubes)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 Baffle (No FT)</td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
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<td>$L_{IN} = 0$ cm</td>
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</tr>
<tr>
<td>$L_{IN} = 1$ cm</td>
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</tr>
<tr>
<td>$L_{IN} = 2$ cm</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>4</td>
<td>2</td>
</tr>
<tr>
<td>t$_{STUDENT(99%)}$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Standard Deviation</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

Table N-50. Statistical analysis (Differences of means) – Effect of nozzle penetration with a single baffle (no flux-tubes), with 5 banks & 4 banks (A off and redistributed) ($\rho_{Agg,0} = 1100$ kg·m$^{-3}$, $D_{Agg,0} = 0.5$ cm, $(L/S)_0 = 0.10$, $C_{Rewet} = 1$, $d_b = 5.0$ cm, $M = 7$)

<table>
<thead>
<tr>
<th>1 baffle at bank B (NO flux-tubes)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 Baffle (No FT)</td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>$L_{IN} = 0$ cm</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$L_{IN} = 1$ cm</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$L_{IN} = 2$ cm</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>9</td>
<td>2</td>
</tr>
<tr>
<td>$t_{STUDENT(99%)}$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Standard Deviation</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$\sigma$</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>1 baffle at bank B (Flux-tubes)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
Table N-51. Statistical analysis (Differences of means) – Effect of nozzle penetration with a single baffle (no flux-tubes), with 5 banks & 4 banks (A off and redistributed) ($\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg},0} = 1.5 \text{ cm}$, $\text{(L/S)}_0 = 0.62$, $C_{\text{Rewet}} = 1$, $d_b = 5.0 \text{ cm}$, $M = 7$)

<table>
<thead>
<tr>
<th>1 Baffle (No FT)</th>
<th>REFERENCE</th>
<th>Comparison</th>
<th>L&lt;sub&gt;IN&lt;/sub&gt; = 0 cm</th>
<th>L&lt;sub&gt;IN&lt;/sub&gt; = 1 cm</th>
<th>L&lt;sub&gt;IN&lt;/sub&gt; = 2 cm</th>
</tr>
</thead>
<tbody>
<tr>
<td>Number of samples</td>
<td>3</td>
<td>2</td>
<td>2</td>
<td>3</td>
<td>2</td>
</tr>
<tr>
<td>$t_{\text{Student}}(99%)$</td>
<td>5.84</td>
<td>5.84</td>
<td>5.84</td>
<td>5.84</td>
<td></td>
</tr>
<tr>
<td>Average</td>
<td>7.39E-04</td>
<td>7.13E-04</td>
<td>5.84E-04</td>
<td>1.27E-05</td>
<td>1.49E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.35E-04</td>
<td>3.15E-04</td>
<td>2.08E-06</td>
<td>3.84E-07</td>
<td>3.61E-06</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>3.49E-04</td>
<td>2.35E-04</td>
<td>2.08E-06</td>
<td>2.97E-06</td>
<td></td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
<td>0.08</td>
<td>0.72</td>
<td>0.80</td>
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<tr>
<td>Statistically different?</td>
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<td>FALSE</td>
<td>FALSE</td>
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<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>1 baffle at bank B (NO flux-tubes)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 Baffle (No FT)</td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>Number of samples</td>
<td>10</td>
<td>2</td>
</tr>
<tr>
<td>$t_{\text{Student}}(99%)$</td>
<td>3.17</td>
<td>3.17</td>
</tr>
<tr>
<td>Average</td>
<td>7.67E-05</td>
<td>7.66E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>2.15E-05</td>
<td>2.70E-07</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>2.15E-05</td>
<td>2.25E-05</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>1 baffle at bank B (Flux-tubes)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 Baffle (No FT)</td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>Number of samples</td>
<td>4</td>
<td>2</td>
</tr>
</tbody>
</table>
### Table N-52. Statistical analysis (Differences of means) – Effect of addition of one baffle vs. more than one baffle ($\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg},0} = 1.0 \text{ cm}$, $(L/S)_0 = 0.31$, $C_{\text{Rewet}} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

<table>
<thead>
<tr>
<th></th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>L$_{\text{IN}} = 0$ cm</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>10</td>
<td>3</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td>4.60</td>
<td>4.60</td>
</tr>
<tr>
<td>Average</td>
<td>4.70E-05</td>
<td>4.35E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.30E-05</td>
<td>5.51E-06</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>1.35E-05</td>
<td>1.30E-05</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>

### Table N-53. Statistical analysis (Differences of means) – Effect of addition of one baffle vs. more than one baffle ($\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}$, $D_{\text{Agg},0} = 0.5 \text{ cm}$, $(L/S)_0 = 0.10$, $C_{\text{Rewet}} = 1$, $d_b = 4.1 \text{ cm}$, $M = 7$)

<table>
<thead>
<tr>
<th></th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>L$_{\text{IN}} = 0$ cm</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Number of samples</td>
<td>9</td>
<td>3</td>
</tr>
<tr>
<td>$t_{\text{STUDENT}(99%)}$</td>
<td>4.60</td>
<td>4.60</td>
</tr>
<tr>
<td>Average</td>
<td>4.70E-05</td>
<td>4.35E-05</td>
</tr>
<tr>
<td>Standard Deviation</td>
<td>1.30E-05</td>
<td>5.51E-06</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>1.35E-05</td>
<td>1.30E-05</td>
</tr>
<tr>
<td>$</td>
<td>t</td>
<td>$</td>
</tr>
<tr>
<td>Statistically different?</td>
<td>FALSE</td>
<td>FALSE</td>
</tr>
</tbody>
</table>
Table N-54. Statistical analysis (Differences of means) – Effect of addition of one baffle vs. more than one baffle ($\rho_{Agg,0} = 1100$ kg·m$^{-3}$, $D_{Agg,0} = 1.5$ cm, (L/S)$_0$ = 0.62, C$_{Rewet}$ = 1, $d_b = 4.1$ cm, M = 7)

<table>
<thead>
<tr>
<th></th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>L$_{IN}$ = 0 cm</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>REFERENCE</td>
<td>Comparison</td>
<td>REFERENCE</td>
</tr>
<tr>
<td>1 Baffle</td>
<td>4 Baffles</td>
<td>1 Baffle</td>
</tr>
<tr>
<td>(at B - No FT)</td>
<td>(at B, C, D, E - No FT)</td>
<td>(at B - No FT)</td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
<td>10</td>
<td>3</td>
</tr>
<tr>
<td><strong>t</strong>$_{STUDENT(99%)}$</td>
<td>3.17</td>
<td>3.11</td>
</tr>
<tr>
<td><strong>Average</strong></td>
<td>7.60E-04</td>
<td>6.31E-04</td>
</tr>
<tr>
<td><strong>Standard Deviation</strong></td>
<td>1.29E-04</td>
<td>1.58E-06</td>
</tr>
<tr>
<td><strong>σ</strong></td>
<td>1.23E-04</td>
<td>1.98E-06</td>
</tr>
<tr>
<td>**</td>
<td>t</td>
<td>**</td>
</tr>
<tr>
<td><strong>Statistically different?</strong></td>
<td>FALSE</td>
<td>TRUE</td>
</tr>
</tbody>
</table>

Table N-55. Statistical analysis (Differences of means) – Effect of addition of one baffle vs. more than one baffle ($\rho_{Agg,0} = 900$ kg·m$^{-3}$, $D_{Agg,0} = 1.0$ cm, (L/S)$_0$ = 0.31, C$_{Rewet}$ = 1, $d_b = 4.1$ cm, M = 7)

<table>
<thead>
<tr>
<th></th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>L$_{IN}$ = 0 cm</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>REFERENCE</td>
<td>Comparison</td>
<td>REFERENCE</td>
</tr>
<tr>
<td>1 Baffle</td>
<td>4 Baffles</td>
<td>1 Baffle</td>
</tr>
<tr>
<td>(at B - No FT)</td>
<td>(at B, C, D, E - No FT)</td>
<td>(at B - No FT)</td>
</tr>
<tr>
<td><strong>Number of samples</strong></td>
<td>6</td>
<td>3</td>
</tr>
<tr>
<td>Student (99%)</td>
<td>Average</td>
<td>Standard Deviation</td>
</tr>
<tr>
<td>---------------</td>
<td>---------</td>
<td>--------------------</td>
</tr>
<tr>
<td></td>
<td>3.50</td>
<td>4.034E-05</td>
</tr>
<tr>
<td></td>
<td>3.71</td>
<td>4.46E-05</td>
</tr>
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</table>

Table N-56. Statistical analysis (Differences of means) – Effect of addition of one baffle vs. more than one baffle (\(\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}\), \(D_{\text{Agg},0} = 1.0 \text{ cm}\), \((L/S)_0 = 0.31\), \(C_{\text{Rewet}} = 1\), \(d_b = 5.0 \text{ cm}\), \(M = 7\))

<table>
<thead>
<tr>
<th>L_{IN} = 0 cm</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>(1 \text{ Baffle (at B - No FT)})</td>
<td>10</td>
<td>3</td>
<td>8</td>
<td>4</td>
</tr>
<tr>
<td>(4 \text{ Baffles (at B, C, D, E - No FT)})</td>
<td>3.11</td>
<td>3.17</td>
<td></td>
<td></td>
</tr>
<tr>
<td>(\text{Number of samples})</td>
<td>3.50E-04</td>
<td>4.60E-04</td>
<td>3.96E-05</td>
<td>2.56E-05</td>
</tr>
<tr>
<td>(\text{Standard Deviation})</td>
<td>9.29E-05</td>
<td>7.95E-07</td>
<td>5.54E-06</td>
<td>2.61E-06</td>
</tr>
<tr>
<td>(\sigma)</td>
<td>8.85E-05</td>
<td>5.22E-06</td>
<td></td>
<td></td>
</tr>
<tr>
<td>(</td>
<td>t</td>
<td>)</td>
<td>0.81</td>
<td>4.37</td>
</tr>
<tr>
<td>(\text{Statistically different?})</td>
<td>FALSE</td>
<td>TRUE</td>
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<td></td>
</tr>
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</table>

Table N-57. Statistical analysis (Differences of means) – Effect of addition of one baffle vs. more than one baffle (\(\rho_{\text{Agg},0} = 1100 \text{ kg} \cdot \text{m}^{-3}\), \(D_{\text{Agg},0} = 0.5 \text{ cm}\), \((L/S)_0 = 0.10\), \(C_{\text{Rewet}} = 1\), \(d_b = 5.0 \text{ cm}\), \(M = 7\))

<table>
<thead>
<tr>
<th>L_{IN} = 0 cm</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
<th>5 banks</th>
<th>4 banks (A off)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>REFERENCE</td>
<td>Comparison</td>
<td>REFERENCE</td>
<td>Comparison</td>
</tr>
<tr>
<td>(1 \text{ Baffle (at B - No FT)})</td>
<td>9</td>
<td>3</td>
<td>6</td>
<td>4</td>
</tr>
<tr>
<td>(4 \text{ Baffles (at B, C, D, E - No FT)})</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
### Table \textbf{N-58}. Statistical analysis (Differences of means) – Effect of addition of one baffle vs. more than one baffle ($\rho_{\text{Agg},0} = 1100$ kg·m$^{-3}$, $D_{\text{Agg},0} = 1.5$ cm, (L/S)$_0 = 0.62$, $C_{\text{Rewet}} = 1$, $d_b = 5.0$ cm, $M = 7$)

| $L_{\text{IN}} = 0$ cm | 5 banks | | 4 banks (A off) | |
|-------------------------|---------|-------------------|-------------------|
|                         | REFERENCE | Comparison | REFERENCE | Comparison |
|                         | 1 Baffle (at B - No FT) | 4 Baffles (at B, C, D, E - No FT) | 1 Baffle (at B - No FT) | 4 Baffles (at B, C, D, E - No FT) |
| Number of samples       | 10      | 3                | 9               | 4               |
| $t_{\text{STUDENT}(99\%)}$ | 3.11    | 3.11             | 3.11            | 3.11            |
| Average                 | 7.68E-05 | 6.25E-05         | 1.56E-06        | 2.08E-06        |
| Standard Deviation      | 2.16E-05 | 1.57E-07         | 7.02E-07        | 8.55E-09        |
| $\sigma$                | 2.06E-05 | 6.35E-07         | 1.05            | 1.36            |
| $|t|$                    | 1.05    | 1.05             | 1.05            | 1.05            |
| Statistically different?| FALSE   | FALSE            | FALSE           | FALSE           |

References:

# Curriculum Vitae

Name: Yohann Cochet

**Post-secondary Education and Degrees:**

<table>
<thead>
<tr>
<th>Institution</th>
<th>Location</th>
<th>Field</th>
<th>Years</th>
<th>Details</th>
</tr>
</thead>
<tbody>
<tr>
<td>The University of Western Ontario</td>
<td>London, Ontario, Canada</td>
<td></td>
<td>2016 - 2021, Ph.D. (started as M.E.Sc in 2016, transferred in 2018)</td>
<td></td>
</tr>
<tr>
<td>ENSIC (École Nationale Supérieure des Industries Chimiques)</td>
<td>Nancy, Meurthe-et-Moselle, France</td>
<td>Engineering school (Master)</td>
<td>2014 – 2017</td>
<td></td>
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</table>

**Honours and Awards:**

- Guntz Award for best academic achievement (ENSIC), 2017
- Student travel bursary (ENSIC Alumni), 2016

**Related Work Experience:**

Research and Teaching Assistant

<table>
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<th>Location</th>
<th>Field</th>
<th>Years</th>
<th>Details</th>
</tr>
</thead>
<tbody>
<tr>
<td>The University of Western Ontario</td>
<td>Canada</td>
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<td>2016 – 2020 (Part-time)</td>
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Engineering internship

<table>
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<th>Location</th>
<th>Field</th>
<th>Years</th>
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</thead>
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<tr>
<td>INEOS Polymers Sarralbe</td>
<td>France</td>
<td></td>
<td>2016</td>
<td></td>
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</table>

**Publications:**
