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**Intensification of interfacial mass, heat &
momentum transfer in high-G fluidized beds in
vortex chambers:
experimental & theoretical study of potential
applications.**

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Acronyms

CCD	Charge-Coupled Device
CFD	Computational Fluid Dynamics
CFB	Circulating Fluidized Bed
CRFB	Circulating Rotating Fluidized Bed
CSTR	Continuous Stirred-Tank Reactor
FB	Fluidized bed
FCC	Fluid Catalytic Cracking
1G	earth gravitational field
high-G	high Gravity
HDPE	High Density PolyEthylene
LDV	Laser Doppler Velocimetry
MAIC	Magnetically Assisted Impaction Coater
MSN	methysiloxane-nylon-6
PI	Process Intensification
PID	Porportional-Integral-Derivative (controller)
PSD	Particle Size Distribution
RFB	Rotating Fluidized Bed
RFB-SG	Rotating Fluidized Bed in a Static Geometry
RTD	Residence Time Distribution
SEM	Scanning Electron Microscopy
VOC	Volatile Organic Compound

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Executive summary

Process intensification today groups a large variety of research topics aiming at increasing the efficiency of one of the rate limiting steps in a process. In general the intrinsic reaction kinetics and/or transport phenomena can be rate limiting. Increasing the intrinsic reaction rates selectively is achieved by using a proper catalyst and enormous efforts were made over the last decades to develop catalysts and get an improved understanding of their functioning [1]. Different types of transport phenomena can be distinguished, in particular if different phases are involved. Next to one or multiple fluid phase(s), a granular solid phase can play a role in the process. The latter is often porous, as such offering a high surface area for interactions with the fluid phase. Interfacial transfer limitations, that is, between the bulk fluid phase and the outer surface of the solid phase, and intra-particle transport limitations, i.e. in the fluid inside the pores of the solid phase, can then be encountered. To intensify intra-particle diffusion, the use of designed and multi-scale pore structures is studied [2–4].

The potential rate limiting nature of interfacial and intra-particle mass and heat transport limitations is obvious. Interfacial momentum transfer can, however, also have a significant effect, in particular when the solid phase is in motion, like in gas-solid fluidized beds, the subject of this work. When increasing the gas-solid slip velocity beyond the point of minimum fluidization, the bed expands so that the average solids concentration that can be achieved reduces [5]. Furthermore, the gas-solid slip velocity cannot exceed the so-called terminal velocity of the particles, particles being entrained by the gas at higher velocities. Insufficient interfacial momentum transfer also leads to a non-uniform distribution of the gas and particles in the fluidized bed. Bubbling fluidized beds and the formation of clusters in the riser regime are well-known examples of the heterogeneity that is introduced [6, 7]. The non-uniform distribution of the different phases is detrimental for the contact between them and for interfacial mass and heat transfer. It also makes accurate simulation of fluidized beds extremely challenging [7, 8]. Different types of intelligently designed fluid distributors have been proposed that allow significantly improving the particle bed uniformity and reducing bub-

bling [4, 9]. Interfacial momentum transfer also puts constraints on the types of particles that can be fluidized. When dealing with small/light cohesive particles, van der Waals forces can become important and prevent fluidization. Different approaches to remove the effects of the van der Waals forces have been proposed [10–12]. Typically, modification of the particle surface is involved [12].

The interfacial transfer limitations encountered in fluidized beds are related to the use of earth gravity in their operation. As will be demonstrated in this work, high-G operation allows higher gas-solid slip velocities and intensified interfacial transfer of mass, momentum and heat. The use of vortex chambers is focused on as they allow generating a high-G field without rotating equipment. After introducing some basic characteristics of conventional fluidized bed reactors in Chapter 1, the hydrodynamics and most important design aspects of rotating fluidized beds in vortex chambers are explained in Chapter 2. An important characteristic is the flexibility with respect to the gas flow rate, the latter affecting the centrifugal force and the counteracting radial gas-solid drag force in similar ways. The gas inlet and outlet design are critical to avoid entrainment of particles into the chimney, in particular when dealing with fine particles. Because of the complex hydrodynamics and the lack of quantitative data, in particular on the influence of the operating conditions and vortex chamber design on the particle bed rotational speed, the process intensification that can be truly achieved by using a vortex chamber is difficult to predict. Furthermore, the process intensification that can be achieved will depend on other rate determining factors eventually present in the process considered, e.g. strong intra-particle diffusion limitations. With the focus on the intensification of interfacial transfer and on practical applications, the continuous drying of biomass was experimentally studied in Chapter 4. The performance of a conventional fluidized bed and a rotating fluidized bed in a vortex chamber are compared and the process intensification evaluated. About one order of process intensification can be achieved, thanks to not only the intensification of interfacial mass and heat transfer, but also thanks to the intensification of interfacial momentum transfer. The latter leads to dense and more uniform particle beds at high gas-solid slip velocities and remains an important advantage even in the regime where intra-particle diffusion limits the drying rate. The process intensification requires high-G operation. The particle bed rotation motion being

generated by the gas flow, relatively high air flow rates are required and the air is not optimally used. Furthermore, the fast particle bed rotation leads to intense mixing in the particle bed and, as a result, a non-uniform product quality, i.e. moisture content. Scale-up of the technology is, hence, not obvious. Chapter 5 deals with these two important scale-up issues by studying different scale-up designs numerically. First a simulation model is developed and the model parameters estimated by means of the experimental data from Chapter 4 and additional non-stationary batch data. The simulations of typical commercial scale units show that the solids product uniformity can be significantly improved by working multi-zone with different chambers operated in series for the particles. A compartmented vortex chamber can, for example be used. The air consumption can be significantly reduced and the air utilization improved by re-using the air leaving one chamber in another chamber, that is, by operation in series for the air. A dual-chimney design can, for example be used to recycle the air from one compartment to another.

The intensification of interfacial momentum transfer does not only allow dense and more uniform particle beds at high gas-solid slip velocities and as such smaller equipment. It also opens the way to fluidizing and processing particles that cannot be dealt with in conventional fluidized beds and for maintaining a stable fluidized bed in the presence of a dispersed liquid phase at relatively high concentrations. Focusing on practical applications and interfacial transfer, the coating of cohesive powders is experimentally studied in Chapter 7. It is shown that a stable rotating fluidized bed of cohesive (fine) powders can be generated provided that a vortex chamber with sufficiently small gas inlet slots is used. Furthermore, it is demonstrated that the bed can be maintained when introducing a disperse liquid phase via a centrally installed spray nozzle and this within a sufficiently wide liquid droplet concentration range. Coating of cohesive particles can be successfully carried out and a good quality coating is obtained with a very short coating time. The impact of some important operating conditions on agglomeration and on the coating efficiency is studied. The flexibility of rotating fluidized beds in vortex chambers with respect to the gas flow rate is used to control agglomeration. The latter can be significantly reduced by operating at higher gas flow rates. The coating efficiency, on the other hand, can be drastically improved by using a higher concentration of smaller droplets, preventing uncoated particles to be entrained into the chimney.

Chapter 8, finally goes one step further, to the production of particles from droplets, i.e. spray drying in vortex chambers. A significant reduction of the equipment size is aimed at without loss of product quality. A proof of concept study was carried out, including a first investigation of the implications of the absence of a dense particle bed. The hydrodynamics is very different than in the presence of a particle bed and a free vortex behavior results in very high gas rotational speeds. A spray nozzle located in the vortex chamber generates too much turbulence in such a flow field, resulting in too pronounced depositions of liquid material on the walls of the small vortex chamber type spray dryer. Spray nozzles installed in the cylindrical wall of the vortex chamber are required. With such a configuration, milk powder with acceptable characteristics was produced.

The experimental and numerical studies presented in Chapters 4, 5, 7 and 8 demonstrate the significant process intensification that can be achieved by using vortex chambers and the relation with the intensification of interfacial mass, heat and momentum transfer. Dense and more uniform particle beds at high gas-solid slip velocities can be generated. For some existing processes this allows significantly reducing the equipment size, as for example in the drying of biomass. But new processing routes become also possible by the intensified interfacial momentum transfer. Cohesive particles can, for example be fluidized and coated with a dispersed liquid phase introduced in relatively high concentrations. Scale-up of the complex vortex chamber technology is not straightforward. Accurate modeling and simulation is required along the lines developed in Chapter 5.

Introduction

In the chemical industry, process intensification is related to the attempts made to improve the efficiency and safety of the processes, reduce the size of the production plants and optimize the operating conditions. Process intensification requires to overcome either kinetics, transport or hydrodynamics limitations. This is achieved through the use of hybrid or multifunctional units, the development of new materials/catalysts, catalyst/reactor structuring or the use of microfluidic reactors. In processes carried out in fluidized beds, limitations related to interfacial mass, heat or momentum transfer can be encountered. Some fluidized bed processes, such as drying, are intrinsically governed by mass and heat transfer, but mass and/or heat transfer can also determine the reaction rates in some catalytic process, making the use of more active catalysts pointless. Finally, the impossibility of fluidizing small cohesive particles (Geldart C-type) in conventional fluidized bed is due to insufficient interfacial momentum transfer.

In a conventional fluidized bed (FB), the particles are suspended as a result of two counteracting forces, the gravity and the drag force caused by the upward flowing gas. The gas flow rate cannot exceed the value for which the drag force equals the gravity force to avoid particles entrainment. As a result, the gas-solid slip velocity and the related interfacial mass, heat and momentum transfer rates are limited by the value of gravity. Furthermore, increasing the gas flow rate in a conventional fluidized bed leads to bed expansion and related low bed densities and to bubbling, i.e. poor bed uniformity. Bed expansion also limits the width-to-height ratio in FB, preventing their operation with an extremely short gas residence time.

To deal with these issues, high-G fluidization can be considered. In rotating fluidized beds, the particle bed is rotating in a cylindrically-shaped vessel to generate a radially outward centrifugal force. The gas flows radially inward, producing a counteracting drag force on the particles. If the particles rotate sufficiently fast, high-G fluidization is obtained. Two ways of making the particle bed rotate have been considered in the literature. The simplest one is to use a motor to make the fluidiza-

tion chamber rotate around its axis of symmetry. The rotation of the vessel can, however, lead to mechanical vibrations or sealing problems that are detrimental for its use in industrial applications. The alternative way to obtain a rotating fluidized bed, the focus of this work, is to use a vortex chamber. In this case, the rotation of the particles is induced by the tangential injection of the gas through inlet slots on the outer cylindrical wall of the vessel. So-called rotating fluidized beds in a static geometry have no moving parts but do not allow the independent control of the gas flow rate and the particle bed rotational speed. As a result, the design of the vortex chamber depends on the characteristics of the particles to be fluidized. Nevertheless, vortex chambers offer a unique flexibility with respect to the gas flow rate as the latter affects both the centrifugal force and the counter-acting gas solid drag force in a similar way.

The general aim of this work is to demonstrate experimentally the intensification of the interfacial mass, heat and momentum transfer using vortex chambers instead of conventional fluidized beds. To this end, three processes have been experimentally studied: the drying of biomass particles, the coating of cohesive powder and the spray drying of milk powder. All of these processes involve water evaporation and are thus governed by mass and heat transfer. Momentum transfer is also a key issue as it governs the hydrodynamic characteristics in the vortex chamber and determines if a particle bed can be generated. Spray drying differs from the two other processes as no rotating fluidized bed is created in the vortex chamber.

The work is organized as follow. Chapter 1 gathers the basic concepts about (conventional) fluidized beds and Chapter 2 is a state-of-the-art on vortex chambers. Chapters 3, 4 and 5 focus on biomass drying: Chapter 3 presents the conventional biomass drying techniques, Chapter 4 reports the experimental study of continuous biomass drying in both a conventional fluidized bed and a vortex chamber, and Chapter 5 presents the development of a numerical model to simulate biomass drying in vortex chambers. Next, particle coating is addressed: Chapter 6 summarizes the existing techniques for particle coating and Chapter 7 reports an experimental study of low-temperature coating of cohesive powder in vortex chambers. Finally, Chapter 8 is an outlook on vortex chamber spray drying through an experimental proof of concept.

The conventional fluidized bed

1.1 Introduction

In fluidized bed reactors, particles are suspended by an upward flow of a fluid. Depending on the fluid nature, they can be classified as gas-solids, liquid-solids or gas-liquid-solids fluidized beds. In each case, the fluidized bed can be obtained only within a certain range of fluid velocities. On one hand, this velocity has to be higher than the minimum fluidization velocity, that is, the velocity from which the particle bed starts to expand. If the fluid velocity is below that minimum fluidization velocity, the relative position of the particles is conserved and the bed is categorized as a fixed bed. On the other hand, if the fluid velocity is higher than the terminal or free-fall velocity, the particles are entrained by the fluid through and out of the reactor. It should be remarked that particle entrainment is desired in some cases, e.g. if small residence times are required [13].

1.2 History

The fluidized bed appeared first in 1926. It was originally used by Winkler to develop a coal gasification process but without commercial success. Apart from that, the Standard Oil Company (now ExxonMobil) had been searching from the beginning of the 1900's an alternative process for the thermal cracking of the crude oil large molecules in order to convert these into smaller and more valuable ones. In 1936, the "Houdry process" was developed. This new process made use of a catalytic fixed bed to crack the crude oil molecules. The drawback of this process was the catalyst deactivation because of coke deposition. As a result, the process had to be periodically interrupted in order to burn the coke and

reactivate the catalyst. Two years later, Exxon Research together with a consortium of other oil companies (Catalytic Research Associates) looked at developing a new catalytic cracking process that would bypass Houdry's patent. The concept of a moving catalytic bed was considered during the discussions. The designation "fluidized bed" was introduced because the moving particle bed had some properties similar to those of a fluid. Although the process had still to be interrupted to regenerate the catalyst, this new process showed some advantages compared to Houdry's, among which an increase of the production capacity. The first fluidized bed industrial unit for catalytic cracking was launched in 1942 [5, 14, 15].

From the sixties, fluidized beds were also used for exothermic processes requiring a careful control of the temperature. Among them, the Sohio process for acrylonitrile production and the Union Carbide process for polyethylene production are two examples. The latter allowed the utilisation of polyethylene for new applications by decreasing its costs of production [5, 14, 15]. Fluidized beds were also considered for non-catalytic processes, such as particle drying or coating.

Recently, new catalysts have been developed with increased activity and improved selectivity. In some cases, these catalysts require a frequent regeneration. That led to the development of circulating fluidized beds, also known as risers. Nowadays, fluidized beds are widely applied in chemical, pharmaceutical and feed and food industries to carry out chemical or physical processes.

1.3 Comparison and differences between fixed and fluidized beds

As shown in figure 1.1, different types of reactors are used to contact a fluid with solid particles. Among them, one can distinguish between the fixed bed (a,b,c) and the fluidized bed (d,e,f) reactors. The "moving-bed" (g) positions itself between the fixed and fluidized beds, showing benefits and drawbacks of both. The main differences between the fixed and fluidized beds are listed below [13, 16]:

- From an hydrodynamic point of view, the fluid flow in a fixed bed generally approaches a plug flow behaviour. In a bubbling fluidized

1.3. Comparison and differences between fixed and fluidized beds

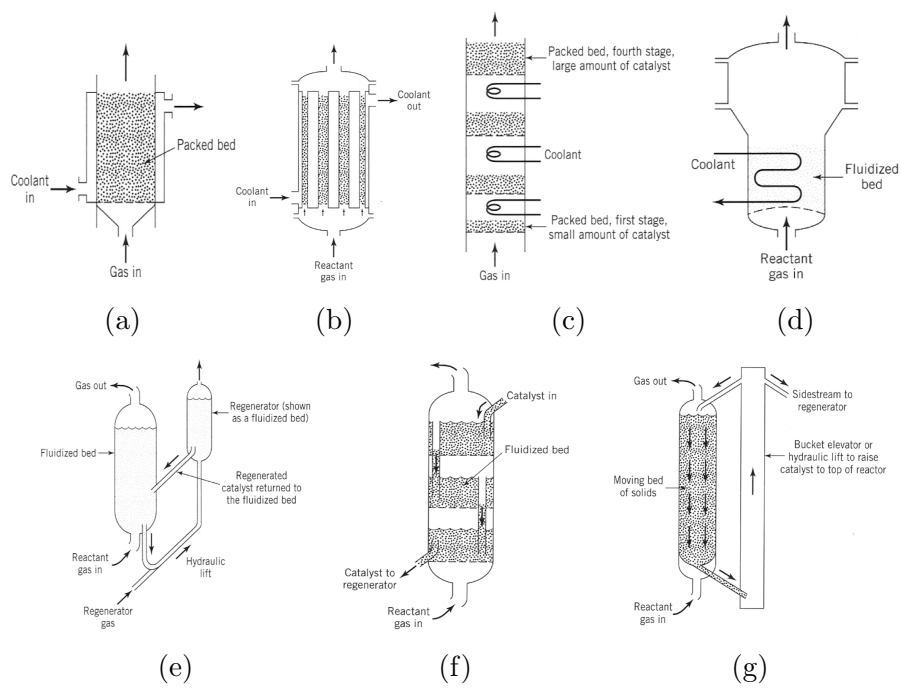


Figure 1.1: Reactors used to contact a fluid with solid particles: (a,b,c) fixed bed reactors, (d,e,f) fluidized bed reactors and (g) moving bed reactor [16].

bed, the flow is more complex and is still investigated. Moreover, some bypassing of the fluid (bubbles) are observed. This leads to poor fluid-solids contacts. As a result, the reactor volume and the quantity of catalyst required to reach a certain conversion are typically higher in a fluidized bed than in a fixed bed. The latter is thus preferred if the fluid-solids contacts efficiency is the key point of the process.

- A fixed bed leads to a low effective thermal conductivity. Consequently, controlling the temperature in large fixed beds is difficult. Moreover, thermal instabilities and temperature non-uniformities can appear in such reactors, leading to "hot spots" where the solids can be damaged. For exothermic polymerisation reactions, these hot spots can also cause polymer degradation. In fluidized beds, the efficient mixing of the particles allows a better control and uniformity of the temperature in the reactor. Fluidized beds will be preferred for processes in which the temperature control is crucial for explosion issues or for the product homogeneity. Furthermore, fluidized beds allow efficient heat exchange with heating or cooling coil.
- Small particles cannot be used in fixed beds because it would result in too high pressure drops. Fluidized beds allow dealing with small particles (if their fluidization is possible) with a reasonable pressure drop. Consequently, fluidized beds will be interesting for processes in which intra-particle diffusion limitations are a key issue.
- Fluidized beds will be preferred if the catalyst must be frequently regenerated because it allows particles circulation as well as continuous feeding and withdrawal in/out of the reactor.

1.4 Fluidization characteristics

Apart from the riser regime (see section 1.4.3), two main fluidization regimes can be distinguished between:

- The particulate fluidization is the regime for which the particle bed density is approximately the same everywhere in the particle bed. From a macroscopic point of view, the bed can be considered as uniform: there is no zone without solids and no bubbles. This

1.4. Fluidization characteristics

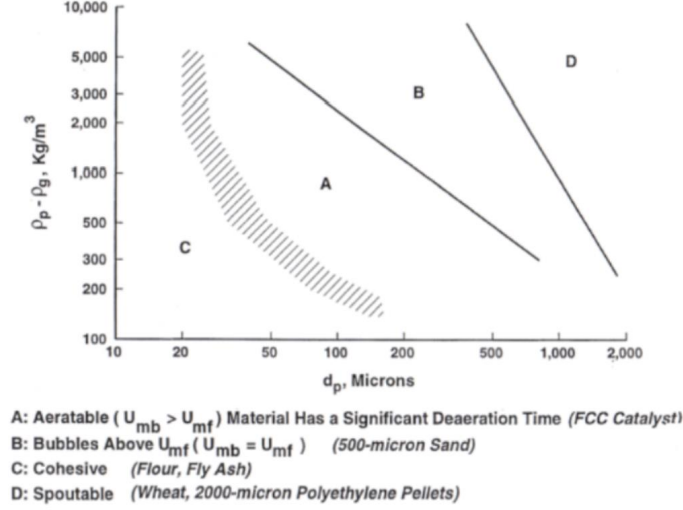


Figure 1.2: Geldart's classification [17].

regime is generally encountered for liquid-solid fluidization or for gas-solid fluidization with small/fine particles at low gas flow rates.

- The aggregative fluidization is the regime in which fluid bubbles containing no solids are present in the bed. This regime mainly occurs for gas-solid and gas-liquid-solid fluidization.

1.4.1 Feasibility of fluidization

Geldart (1973) [17] and Geldart and Abrahamsen (1978) [18] classified solid particles into 4 categories according to their fluidization behaviour. This classification depends on the apparent density ($\rho_s - \rho_f$) and equivalent mean (volume-based) diameter $\bar{d}_p$, of the particles. Figure 1.2 shows Geldart's classification and a description for each class [13, 16].

- A-class : particulate fluidization. The bed expands before bubbles appear.
- B-class : aggregative fluidization. Bubbles appear from the beginning of the fluidization
- C-class : for fine ($< 30 \mu\text{m}$) or light cohesive powders, the fluidization is almost impossible because of the pronounced inter-particle surface (Van Der Waals) forces between particles.

- D-class : the fluidization is possible but the particle bed is not stable. Fluidization can even become impossible for too large particles (>50 mm).

It should be noticed that Geldart's classification is valid for conventional fluidized beds without any external factors that could play a role on the fluidization behaviour. In particular, it is only valid in a 1G gravitational field fluidization and not valid for vibration-assisted fluidized beds [19].

1.4.2 Hydrodynamics of fluidized beds

Introduction

Four hydrodynamic quantities are important when studying the fluidization behaviour [5, 13, 16]:

1. the minimum superficial fluidization velocity u_{mf} .
2. the terminal or free-fall velocity u_t , which is the fluid superficial velocity above which particles are entrained.
3. The fluid volume fraction at the fluidization threshold ε_{mf} .
4. The bed expansion with the fluid superficial velocity u_{sf} .

The difference between the interstitial and superficial fluid velocity should be pointed out here to avoid later confusion.

- The interstitial velocity u_{ik} [m_r/s] is the physical velocity of the phase k in the reactor.
- The superficial velocity u_{sk} [$m_k^3/m_r^2.s$] of the phase k is defined as:

$$u_{sk} = \varepsilon_k \cdot u_{ik} \quad (1.1)$$

where ε_k [m_k^3/m_r^3] is the volumic fraction of phase k .

Before going further, it should be noticed that:

1.4. Fluidization characteristics

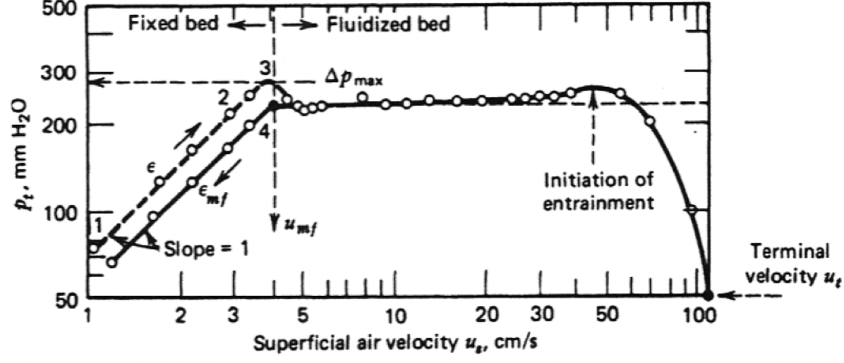


Figure 1.3: Pressure drop ΔP over the particle bed as a function of the superficial velocity of the gas u_{sg} for a bed composed of sand particles with uniform size [5].

1. The optimal usage of the reactor volume requires an operation with the minimum solids expansion for which an efficient mixing of the solid phase is ensured. Typically, a bed expansion of 10 to 25 % is used.
2. In a classical fluidized bed conception, the particles are supposed to be confined in the reactor and potential particles entrainment can be avoided using a filter at the top of the column. However, for specific processes like catalytic cracking (FCC), the circulation of the particles can be desired. So-called circulating fluidized beds (CFBs) or risers are operated above the terminal velocity leading to a co-current flow of the fluid and the solids.

Figure 1.3 shows the evolution of the pressure drop over the particle bed with the superficial velocity of the gas for a bed composed of sand particles with uniform size. As shown on the graph, the pressure drop over the bed increases with the superficial velocity at low superficial velocity (from point 1 to point 3). The pressure drop in this range of superficial velocity is due to the resistance to a fluid flow through a fixed bed granular medium and can be calculated using the Ergun equation [20]:

$$\frac{\Delta P}{L} = 150 \frac{\varepsilon_s^2}{(1 - \varepsilon_s)^3} \frac{\mu_f \cdot u_{sf}}{\bar{d}_p^2} + 1.75 \frac{\varepsilon_s}{(1 - \varepsilon_s)^3} \frac{\rho_f \cdot u_{sf}^2}{\bar{d}_p} \quad (1.2)$$

where

- $\varepsilon_s [m_s^3/m_r^3]$ is the solids volume fraction ($\varepsilon_s = 1 - \varepsilon_f$ where $\varepsilon_f [m_f^3/m_r^3]$ is the fluid volume fraction).
- $\mu_f [Pa.s]$ ($[kg/m.s]$) is the fluid viscosity.
- $\bar{d}_p [m]$ is the mean equivalent diameter of the particles.
- $\rho_f [kg_f/m_f^3]$ is the fluid density.

The force applied on the bed by friction of the fluid and turbulence generates the pressure drop. This force increases with the fluid flow rate until reaching the apparent weight of the bed when $u_{sg} = u_{mf}$. At that point, the particles start to be suspended and the bed loses its consistency. The small decrease of the pressure drop after point 3 is due to the re-arrangement of the particles.

Beyond the fluidization threshold, the particles move relative to each other and the bed is fluidized. The pressure drop over the particle bed is not dependent on the superficial velocity anymore as it remains constant at the value corresponding to the apparent weight of the particle bed. The bed expands with increasing superficial fluid velocity, leading to an increase of the fluid volume fraction which makes the interstitial velocity increasing less than proportionally with the gas flow rate. The superficial velocity and the fluid volume fraction at the minimum fluidization point u_{mf} and ε_{mf} can be easily determined from a pressure drop measurement.

The superficial velocity and the fluid volume fraction at the minimum fluidization point

If experimental data are available for u_{mf} , the fluid volume fraction at the minimum fluidization point can be determined from the exact balance between the pressure drop and the weight of the bed.

$$\frac{\Delta P}{L_{mf}} = (1 - \varepsilon_{mf}) \cdot (\rho_s - \rho_f) \cdot g \quad (1.3)$$

where

- $L_{mf} [m]$ is the bed height at the minimum fluidization point.

1.4. Fluidization characteristics

- ρ_s [kg_s/m_s^3] is the density of the particles.

If u_{mf} is not known, it can be calculated remembering that u_{mf} is the superficial velocity for which the force applied on the bed by friction of the fluid is equal to the apparent weight of the bed. Consequently, (1.2) and (1.3) are equalized:

$$\begin{aligned} \frac{\Delta P}{L_{mf}} &= 150 \frac{(1 - \varepsilon_{mf})^2}{\varepsilon_{mf}^3} \frac{\mu_f \cdot u_{mf}}{\bar{d}_p^2} + 1.75 \frac{(1 - \varepsilon_{mf})}{\varepsilon_{mf}^3} \frac{\rho_f \cdot u_{mf}^2}{\bar{d}_p} \\ &= (\rho_s - \rho_f) \cdot (1 - \varepsilon_{mf}) \cdot g \end{aligned} \quad (1.4)$$

An non-dimensional form of this equation can be obtained by multiplication with $\frac{\rho_f \cdot \bar{d}_p^3}{\mu_f^2 \cdot (1 - \varepsilon_{mf})}$

$$150 \frac{(1 - \varepsilon_{mf})}{\varepsilon_{mf}^3} Re_{p,mf} + 1.75 \frac{1}{\varepsilon_{mf}^3} Re_{p,mf}^2 = Ar \quad (1.5)$$

with

- $Re_{p,mf} = \frac{\bar{d}_p u_{mf} \rho_f}{\mu_f}$ [-] the Reynolds number at the minimum fluidization point.
- $Ar = \frac{\rho_f (\rho_s - \rho_f) \bar{d}_p^3 g}{\mu_f^2}$ [-] the Archimedes number.

The calculation of u_{mf} requires ε_{mf} . If not available, correlations can be used to determine u_{mf} . Leva [21] found a correlation with the particles mean diameter:

$$u_{mf} = 1.118 \cdot 10^{-13} \frac{\bar{d}_p^{1.82} \cdot (\rho_s - \rho_f)^{0.94}}{\rho_f^{0.06} \cdot \mu_f^{0.88}} \quad (1.6)$$

with $\bar{d}_p$ expressed in μm .

Alternatively, the terms including ε_{mf} in equation (1.5) can be approximated by [22]:

- $150 \frac{(1 - \varepsilon_{mf})^2}{\varepsilon_{mf}^3} \simeq 1650$
- $1.75 \frac{(1 - \varepsilon_{mf})}{\varepsilon_{mf}^3} \simeq 24.5$

Introducing these values in equation (1.5) leads to:

$$Re_{p,mf} = 33.7 \left[(1 + 3.6 \cdot 10^{-5} Ar)^{0.5} - 1 \right] \quad (1.7)$$

or

$$u_{mf} = 33.7 \frac{\mu_f}{\bar{d}_p \rho_f} \left[\left(1 + 3.6 \cdot 10^{-5} \frac{\rho_f (\rho_s - \rho_f) \bar{d}_p^3 g}{\mu_f^2} \right)^{0.5} - 1 \right] \quad (1.8)$$

Other correlations similar to equation (1.7) but with different numerical values have also been developed [23, 24]. It should be remarked that the equations presented above are not valid for tri-phase systems or for liquid-solid fluidization with particles smaller than $200 - 300 \mu m$. Indeed, the capillarity forces have to be accounted for in the force balance for these cases. The capillarity forces can even prevent fluidization if they become dominant.

The terminal or free-fall velocity

While further increasing the fluid velocity, the pressure drop decreases, indicating that particles start to be entrained by the fluid flow out of the reactor until the bed disappears when $u_{sf} = u_t$, the terminal or free-fall velocity. Particles entrainment is generally undesired in industrial operation although useful for some processes.

The terminal velocity is reached when the drag force F_D exerted on the particles by the fluid becomes higher than the apparent weight of these particles. The drag force is defined as:

$$F_D = \frac{1}{2} C_D \cdot \rho_f \cdot A \cdot u_{sg}^2 \quad (1.9)$$

where

- $C_D [-]$ is the drag coefficient.
- $A [m^2]$ is the cross-sectional area. For a spherical particle, $A = \frac{\pi \bar{d}_p^2}{4}$.

Consequently, when $u_{sg} = u_t$, the following equation can be written:

$$\frac{1}{2} C_D \cdot \rho_f \frac{\pi \bar{d}_p^2}{4} u_t^2 = \frac{\pi \bar{d}_p^3}{6} (\rho_s - \rho_f) \cdot g \quad (1.10)$$

1.4. Fluidization characteristics

so that:

$$u_t = \sqrt{\frac{4.g.\bar{d}_p.(\rho_s - \rho_f)}{3.\rho_f.C_D}} \quad (1.11)$$

The drag coefficient C_D depends on the Reynolds number $Re_p = \frac{\bar{d}_p.\rho_f.u_t}{\mu_f}$. For spherical particles and laminar flows ($Re_p < 0,4$), it is given by

$$C_D = \frac{24}{Re_p} \quad (1.12)$$

leading to the Stoke's law

$$u_t = \frac{(\rho_s - \rho_f).g.\bar{d}_p^2}{18.\mu_f} \quad (1.13)$$

For intermediate Reynolds numbers ($1 < Re_p < 10^3$), the following expression can be used [13]

$$C_D = -5.50 + \frac{69.43}{\ln Re_p + 7.99} \quad (1.14)$$

Finally, at high Reynolds numbers ($Re_p > 10^3$),

$$C_D = 0.44 \quad (1.15)$$

and the terminal velocity is given by the Newton's law:

$$u_t = \sqrt{\frac{3.1.\bar{d}_p.(\rho_s - \rho_f).g}{\rho_f}} \quad (1.16)$$

An alternative relationship valid for $Re_p < 10^3$ [25] is given by

$$C_D = \frac{24}{Re_p} (1 + 0.15.Re_p^{0.687}) \quad (1.17)$$

In each case, the terminal velocity increases with the particle mean diameter. That means that fine particles are more easily entrained.

As previously mentioned, a fluidized bed without particles entrainment will be obtained if $u_{mf} < u_{sf} < u_t$. The calculation of the minimum fluidization and terminal velocities is therefore crucial for the design of a fluidized bed. The ratio $\frac{u_t}{u_{mf}}$ is typically ranging between 10

for big particles ($Re_p > 10^3$) and 90 for small particles ($Re_p < 0,4$) [13]. Generally, fluidized beds are operated at $u_{sf} = 5 - 10 \times u_{mf}$ to ensure a good fluidization behaviour while limiting the risk of particles entrainment. Cyclone can be used downstream to remove the potentially entrained particles from the fluid.

Bed expansion

When the bed becomes fluidized, the bed pressure drop remains constant at the value corresponding to the apparent weight of the particle bed. From this point, the bed expands, meaning that its height as well as the fluid volume fraction ε_f increase. The expansion behaviour depends on the fluid used to fluidize the particles.

- **Liquid-solid fluidization**

In the case of liquid-solid fluidization, the bed remains uniform when increasing the liquid velocity. The bed shows a particulate fluidization behaviour and the solids volume fraction $\varepsilon_s = 1 - \varepsilon_l$ decreases with u_{sl} according to the empirical equation of Richardson et Zaki [26]

$$\frac{u_{sl}}{u_t} = (1 - \varepsilon_s)^n = \varepsilon_l^n \quad (1.18)$$

The exponent n is positive and depends on the Reynolds number $Re_p = \frac{u_t \cdot \rho_l \cdot \bar{d}_p}{\mu_l}$, the particles shape and the ratio $\bar{d}_p/d_t$ (with d_t the reactor diameter).

- **Gas-solid fluidization**

For gas-solid fluidization, increasing the gas superficial velocity u_{sg} above u_{mf} can lead to two different behaviours for the bed expansion:

1. If the fluidization is uniform (particulate fluidization), the Richardson and Zaki equation ((1.18)) can be used. However, this case is not common for gas-solid fluidization, except for small particles.
2. If the fluidization is heterogeneous with bubble appearance (aggregative fluidization), there is no equation available to correlate the solid volume fraction with u_{sg} .

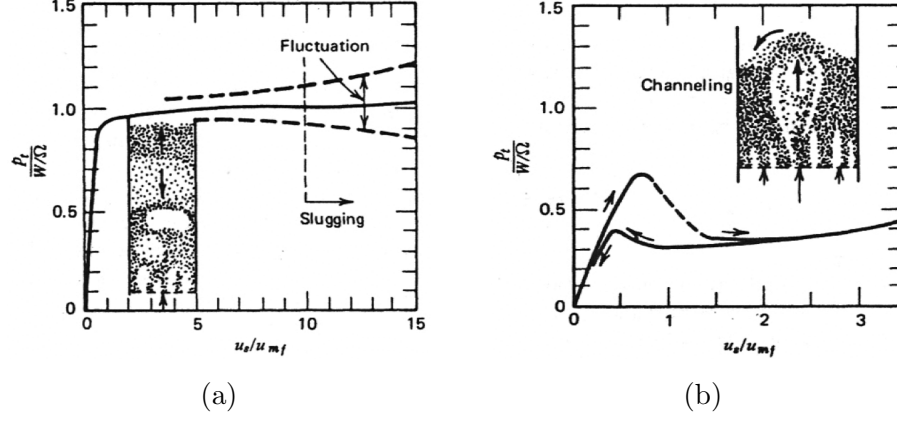


Figure 1.4: Bed pressure drop ΔP versus superficial gas velocity u_{sg} when (a) "slugging" or (b) "channelling" occurs [5].

"Slugging" and "Channelling"

Figure 1.3 presents the pressure drop evolution with the superficial fluid velocity for a normal fluidized bed behaviour. However, for certain operating conditions, undesired particle bed behaviour can occur. Figure 1.4 shows the pressure drop over the particle bed versus u_{sf} in case of "slugging" (a) or "channelling" (b).

- **Slugging** occurs if the bubbles diameter has the same order of magnitude as the reactor diameter. Slugging can be avoided by decreasing the ratio height/diameter of the reactor.
- **Channelling** occurs when the fluid flow has preferential paths to go through the bed. Typically, the major part of the fluid flows in the center of the reactor, leaving the major part of the solids near the reactor walls (core-annular flow). This phenomenon can be avoided by increasing the ratio height/diameter of the reactor or by a careful design of a fluid distributor at the bottom of the column.

1.4.3 Gas-solid fluidization regimes

Figure 1.5 shows the solids volume fraction ε_s in a fluidized bed as a function of the gas superficial velocity u_{sg} for both large (a) and small (b)

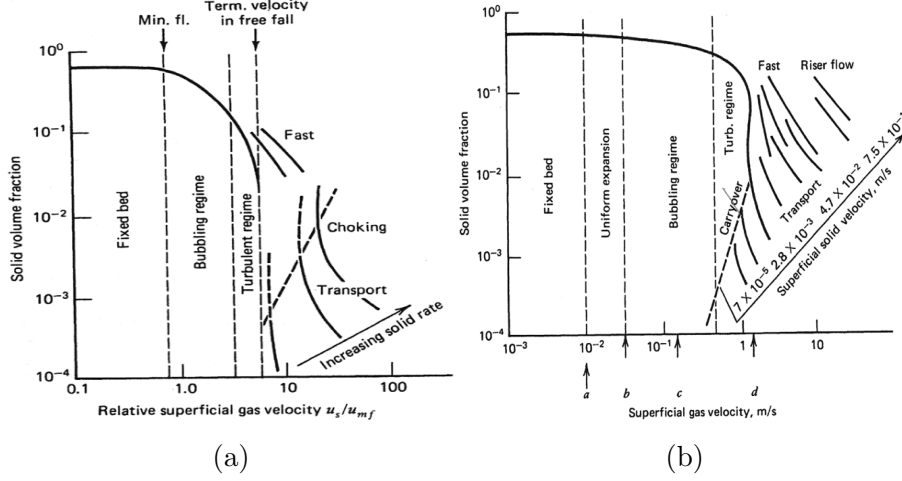


Figure 1.5: solids volume fraction ε_s in a fluidized bed as a function of the gas superficial velocity u_{sg} for both (a) large and (b) small particles [5]

particles. In each case, different fluidization regimes can be observed [5].

For large particles (Geldart B type), bubbling is observed as soon as the bed is fluidized, i.e. as soon as the superficial gas velocity reaches u_{mf} . When increasing further the superficial gas velocity, the flow becomes turbulent leading to a reduced bubble lifetime and, as a result, a more uniform fluidized bed. If u_{sg} is increased beyond the terminal velocity, a transport regime is observed: the solids volume fraction decreases sharply because of particles entrainment [5, 16].

For small particles (Geldart A type), a uniform expansion of the bed is first observed when the bed becomes fluidized (point a). This particulate fluidization regime is maintained in a certain range of gas velocities (between point a and b). The bed is more coherent as the particles are moving less individually compared to the bubble regime. When increasing further the gas superficial velocity, bubbling, turbulent and transport regimes occur as for Geldart group B particles. Figure 1.6 shows schematic views of the different fluidization behaviours described here-above [5, 16].

1.5. Interfacial heat and mass transfer in gas-solid fluidized beds

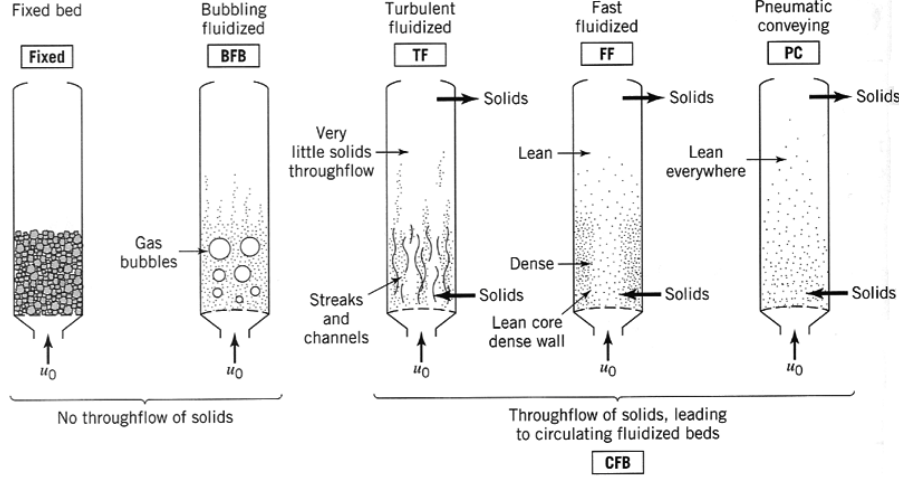


Figure 1.6: Schematic view of the different fluidization behaviours [16]

1.5 Interfacial heat and mass transfer in gas-solid fluidized beds

Because of the high gas-solid slip velocities and efficient gas-solid contacts, gas-solid interfacial heat and mass transfer in fluidized beds are more efficient than in other technologies, including fixed beds or rotating drums. Consequently, only when fast reactions are involved, heat and mass transfer limitations have to be accounted for in fluidized beds. The calculation of interfacial mass and heat transfer rates are also required when dealing with physical processes that are only limited by transfer phenomena [5, 13, 27].

The interfacial mass transfer flux j_{gs} is defined as the amount of A transferred from the gas phase to the solid phase per unit time and per unit of interfacial surface area. It can be described as the product of a mass transfer coefficient k_g and the difference between the interfacial and bulk concentrations of A in the gas phase:

$$j_{gs} = k_g (C_A - C_{Ai}) M_A [kg / (m_i^2 \cdot s)] \quad (1.19)$$

Similarly, the interfacial heat transfer rate Q_{gs} , the quantity of heat transferred from one phase to another per unit time and per unit of interfacial surface area can be described in terms of a heat transfer

coefficient h_f and the difference between the gas interfacial and bulk temperatures:

$$Q_{gs} = h_f (T - T_i) \quad [J / (m_i^2.s)] \quad (1.20)$$

In case the mass and heat transfers are not limiting for the process rate, the gas temperature and concentration of A at the external surface of the particles are equal to the bulk temperature and concentration.

The interfacial mass and heat transfer coefficients are usually correlated introducing respectively the Sherwood ($Sh = \frac{k_g d_p}{D_g}$), Reynolds and Schmidt ($Sc = \frac{\mu_g}{\rho_g D_g}$) numbers and the Nusselt ($Nu = \frac{h_f d_p}{\lambda_g}$), Reynolds and Prandtl ($Pr = \frac{\mu_g c_{p,g}}{\lambda_g}$) numbers [5]. Tables 1.1 and 1.2 summarize the different correlations used to calculate respectively the mass and heat transfer coefficients in fixed and fluidized beds. Most of them make use of the Colburn j-factors for mass and heat transfer.

$$\begin{aligned} Sh &= j_D Re_p Sc^{\frac{1}{3}} \\ \Leftrightarrow k_g &= j_D \frac{\left(\frac{\dot{m}_g}{\Omega}\right)}{\rho_g} Sc^{-2/3} \quad [m_g^3 / (m_i^2.s)] \end{aligned} \quad (1.21)$$

$$\begin{aligned} Nu &= j_D Re_p Pr^{\frac{1}{3}} \\ \Leftrightarrow h_f &= j_H c_{pg} \left(\frac{\dot{m}_g}{\Omega}\right) Pr^{-2/3} \quad [J / (m_i^2.K.s)] \end{aligned} \quad (1.22)$$

Depending on the model, j_D and j_H are correlated with the void fraction and either the Reynolds number $Re_p = \frac{d_p(\dot{m}_g/\Omega)}{\rho_g}$ or the modified Reynolds number $Re_{pm} = Re_p/(1 - \varepsilon)$. A correlation for j_H that includes the Archimedes number Ar [28] and more complex expressions for the calculation of k_g and h_f [29–33] have also been developed, some of them with the help of CFD [34].

These correlations are valid only for small driving forces, i.e. small concentration or temperature difference between the gas phase and the solid phase. Indeed, if the mass or heat transfer rate is sufficiently high, the velocity profile in the gas film where the transfer occurs can

1.5. Interfacial heat and mass transfer in gas-solid fluidized beds

Reference	Correlation	Applicability
Gamson et al. (1943)	$j_D = 16.8 Re_p^{-1}$ if $Re_p < 40$	Fixed beds
[35]	$j_D = 0.989 Re_p^{-0.41}$ if $Re_p > 350$	$20 < Re_p < 7000$
Wilke & Hougen (1945)	$j_D = 1.82 Re_p^{-0.51}$ if $Re_p < 40$	Fixed beds
[36]	$j_D = 0.989 Re_p^{-0.41}$ if $Re_p > 350$	$20 < Re_p < 7000$
de Acetis & Thodos (1960)	$j_D = \frac{0.723}{Re_p^{0.41-1.50}}$	Fixed beds
[37]		$15 < Re_p < 3000$
Bradshaw & Bennet (1961)	$j_D = 0.606 Re_p^{-0.309}$	Fixed and fluidized beds
[38]		$400 < Re_p < 10000$
Yoshida et al. (1962)	$j_D = 0.84 \left(\frac{Re_{pm}}{6} \right)^{-0.51}$ if $\frac{Re_{pm}}{6} < 50$	Fixed beds
[39]	$j_D = 0.57 \left(\frac{Re_{pm}}{6} \right)^{-0.41}$ if $\frac{Re_{pm}}{6} > 50$	$0.01 < \frac{Re_{pm}}{6} < 1000$
Gupta & Thodos (1962)	$\varepsilon j_D = \frac{0.300}{(Re_p \sqrt{\pi})^{0.35-1.90}}$	Fixed and fluidized beds
[40]		$50 < Re_p \sqrt{\pi} < 10000$
Gupta & Thodos (1962b)	$\varepsilon j_D = 0.010 + \frac{0.863}{Re_p^{0.58-0.483}}$	Fixed and fluidized beds
[41]		$1 < Re_p < 3000$
Bradshaw & Myers (1963)	$j_D = 2.25 Re_{pm}^{-0.501}$	Fixed and fluidized beds
[42]		Large particles $200 < Re_{pm} < 10000$
Gupta & Thodos (1963)	$\varepsilon j_D = 2.06 Re_p^{-0.575}$	Fixed and fluidized beds
[43]		$90 < Re_p < 4000$
Petrovic & Thodos (1968)	$\varepsilon j_D = 0.357 Re_p^{-0.359}$	Fixed beds
[44]		$3 < Re_p < 900$
Dwivedi & Upadhyay (1977)	$\varepsilon j_D = 0.765 Re_p^{-0.82} + 0.365 Re_p^{-0.386}$	Fixed and fluidized beds
[45]		$0.01 < Re_p < 15000$
Gunn (1978)	$Sh = (1 - 10\varepsilon + 5\varepsilon^2)(1 + 0.7 Re_p^{0.2} Sc^{1/3})$ $+ (1.33 - 2.4\varepsilon + 1.2\varepsilon^2) Re_p^{0.7} Sc^{1/3}$	Fixed and fluidized beds
[29]		$50 < Re_p < 8000$

Table 1.1: Correlations for calculating the mass transfer coefficient k_g in fixed and fluidized beds.

become distorted. In this case, the mass/heat transfer coefficient becomes dependent on the mass/heat transfer rate itself. Moreover, the correlations only describe the diffusive transport. When dealing with non-dilute systems, equations (1.21) and (1.22) have to be corrected by a "film pressure" factor to account for a convective contribution to the mass transfer [5, 43, 50].

A graphical representation of the j_D and j_H factor evolutions with the Reynolds number for the different correlations are shown in Figure 1.7 for a void fraction of 0.8 (fluidized bed) and in Figure 1.8 for a void fraction of 0.4 (packed bed). The dotted lines means that the correlation cannot be applied in this range of particle bed void fraction.

Although heat and mass transfers are efficient in conventional fluidized beds, they are limited by the fact that the gas-solid slip velocity cannot exceed the terminal velocity of the particles in the earth gravitational field [5].

Reference	Correlation	Applicability
Gamson et al. (1943) [35]	$j_H = 18.1 Re_p^{-1}$ if $Re_p < 40$ $j_H = 1.064 Re_p^{-0.41}$ if $Re_p > 350$	Fixed beds $20 < Re_p < 7000$
Glaser & Thodos (1958) [46]	$j_H = \frac{0.535}{(Re_{pm} \sqrt{\pi})^{0.30-1.6}}$	Fixed beds $100 < Re_{pm} \sqrt{\pi} < 9200$
de Acetis & Thodos (1960) [37]	$j_H = \frac{1.10}{Re_p^{0.41-1.50}}$	Fixed beds $15 < Re_p < 3000$
Yoshida et al. (1962) [39]	$j_H = 0.90 \left(\frac{Re_{pm}}{6} \right)^{-0.51}$ if $\frac{Re_{pm}}{6} < 50$ $j_H = 0.61 \left(\frac{Re_{pm}}{6} \right)^{-0.41}$ if $\frac{Re_{pm}}{6} > 50$	Fixed beds $0.01 < \frac{Re_{pm}}{6} < 1000$
Gupta & Thodos (1962) [40]	$\varepsilon j_H = \frac{0.322}{(Re_p \sqrt{\pi})^{0.35-1.90}}$	Fixed and fluidized beds $50 < Re_p \sqrt{\pi} < 10000$
Gupta & Thodos (1962b) [41]	$\varepsilon j_H = 0.0108 + \frac{0.929}{Re_p^{0.58-0.483}}$	Fixed and fluidized beds $20 < Re_p < 10000$
Bradshaw & Myers (1963) [42]	$j_H = 2.52 Re_{pm}^{-0.501}$	Fixed and fluidized beds Large particles $200 < Re_{pm} < 10000$
Gupta & Thodos (1964) [47]	$\varepsilon j_H = 2.06 Re_p^{-0.575}$ if $Re_p < 1900$ $\varepsilon j_H = 2.16 Re_p^{-0.575}$ if $Re_p > 1900$	Fixed and fluidized beds $90 < Re_p < 4000$
Handley & Heggs (1968) [48]	$\varepsilon j_H = 0.255 Re_p^{-0.335}$	Fixed beds $90 < Re_p < 4000$
Gupta et al. (1974) [49]	$\varepsilon j_H = 2.876 Re_p^{-1} + 0.3023 Re_p^{-0.35}$	Fixed and fluidized beds $10 < Re_p < 10000$
Balakrishnan & Pei (1977) [28]	$j_H = \left(\frac{Ar}{Re_{pm}^2} \right)^{0.25}$	Fluidized beds $0.02 < \frac{Ar}{Re_{pm}^2} < 10$
Gunn (1978) [29]	$Nu = (1 - 10\varepsilon + 5\varepsilon^2)(1 + 0.7 Re_p^{0.2} Pr^{1/3})$ $+ (1.33 - 2.4\varepsilon + 1.2\varepsilon^2) Re_p^{0.7} Pr^{1/3}$	Fixed and fluidized beds $50 < Re_p < 8000$

Table 1.2: Correlations for calculating the heat transfer coefficient h_f in fixed and fluidized beds.

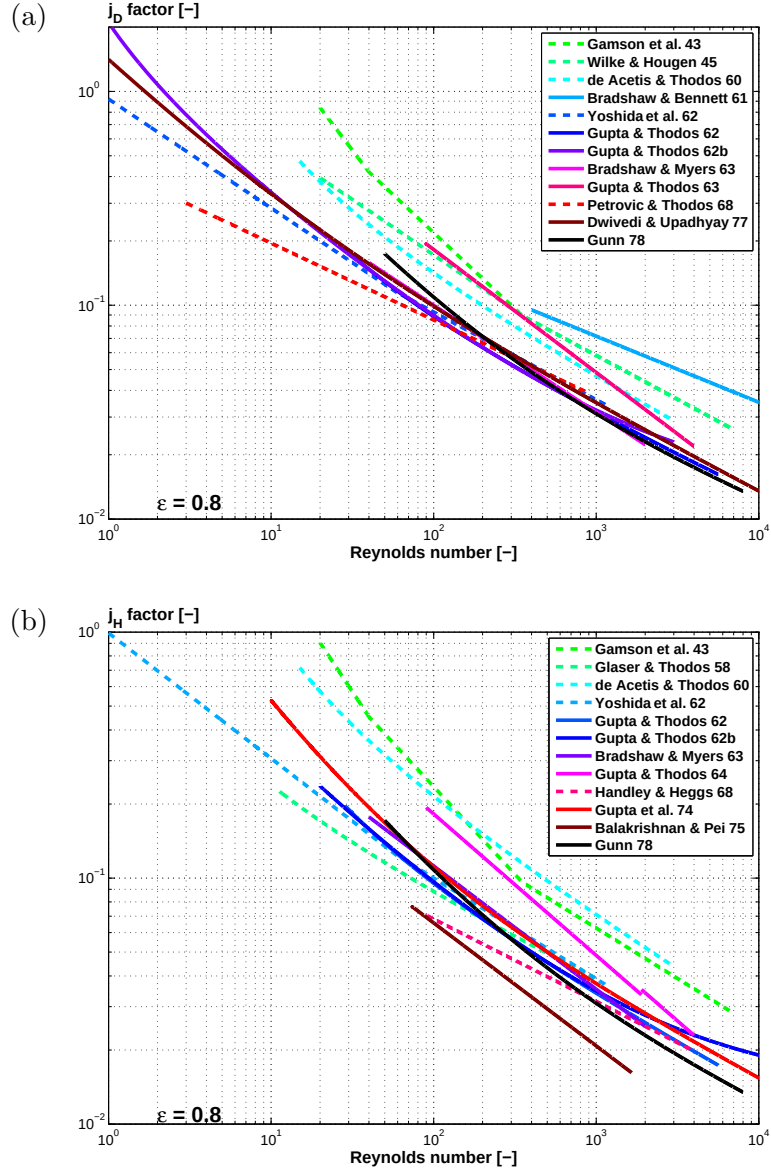


Figure 1.7: (a) j_D and (b) j_H factor evolutions with the Reynolds number for the different correlations in a fluidized bed ($\varepsilon = 0.8$).

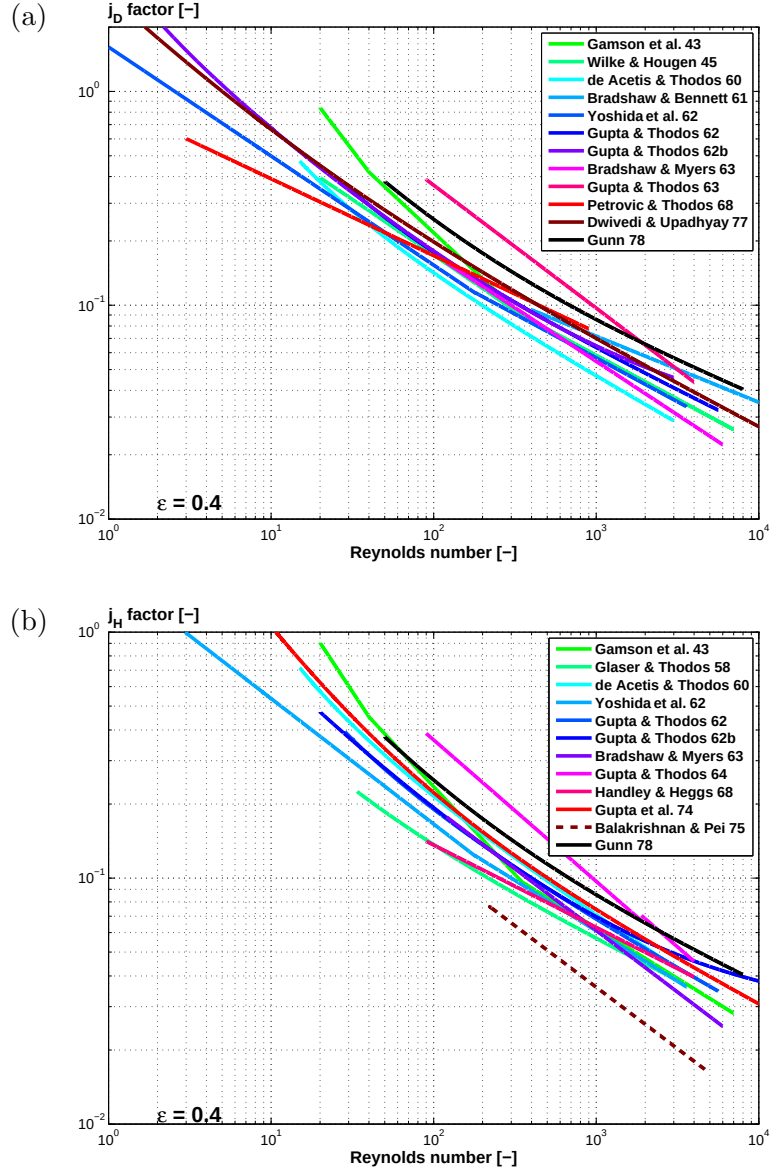


Figure 1.8: (a) j_D and (b) j_H factor evolutions with the Reynolds number for the different correlations in a packed bed ($\varepsilon = 0.4$).

Nomenclature

A	m^2	drag cross-sectional area
Ar	-	Archimedes number
C_A	mol/m_f^3	concentration of A in the fluid phase
C_D	-	drag coefficient
c_{pk}	$\text{J}/(\text{kg}\cdot\text{K})$	heat capacity of phase k
d_p	m	particle mean diameter
D_g	m^2/s	diffusivity in the gas phase
F_D	N	gas-solid drag force
g	m^2/s	earth's standard acceleration due to gravity
h_f	$\text{J}/(\text{K m}_i^2 \text{ s})$	gas-solid heat transfer coefficient
j_D	-	Colburn j-factor for mass transfer
j_H	-	Colburn j-factor for heat transfer
j_{gs}	$\text{kg}/(\text{m}_i^2 \cdot \text{s})$	mass transfer flux between gas and solid phase
k_g	$\text{m}_g^3/(\text{m}_i^2 \text{ s})$	gas-solid mass transfer coefficient
L	m_r	fluidized bed height
M_A	kg/mol	molecular mass of A
$\dot{m}_g$	kg/s	gas phase mass flow rate
Nu	-	Nusselt number
ΔP	Pa	pressure drop
Q_{gs}	$\text{J}/(\text{m}_i^2 \cdot \text{s})$	heat transfer flux between gas and solid phase
Re_p	-	particle-based Reynold number
Re_{pm}	-	modified particle-based Reynold number
Sc	-	Schmidt number
Sh	-	Sherwood number
T	K	temperature
u_{ik}	m_r/s	interstitial velocity of phase k
u_{sk}	$\text{m}_k^3/(\text{m}_r^2 \cdot \text{s})$	superficial velocity of phase k
u_{mf}	$\text{m}_f^3/(\text{m}_r^2 \cdot \text{s})$	fluid superficial velocity at the minimum fluidization point
u_t	m_r/s	particle terminal or free-fall velocity

Greek letters

ε	$\text{m}_f^3/\text{m}_r^3$	particle bed void fraction
ε_k	$\text{m}_k^3/\text{m}_r^3$	volume fraction of phase k
λ_k	$\text{W}/(\text{m.K})$	heat conductivity of phase k
μ_k	Pa.s	viscosity of phase k
ρ_k	kg/m_k^3	density of phase k
Ω	m^2	cross-sectional area

Subscripts

f	fluid phase
g	gas phase
i	gas-solid interface
mf	minimum fluidization
r	reactor
s	solid phase

Rotating fluidized beds in a vortex chamber

2.1 Introduction

In conventional, i.e. gravitational, fluidized beds (FB), the heat, mass and momentum transfer rates are limited because the gas-solid slip velocity cannot exceed the terminal velocity of the particles in the earth gravitational field. Moreover, even below the terminal velocity, increasing the gas flux to enhance mass and heat transfer rates leads to a dramatic decrease of the solids volume fraction in the particle bed. Finally, conventional fluidized beds do not allow fluidizing fine (below 30-50 μm) or light cohesive particles (1G-Geldart-C type) because of the prevailing Van der Waals forces.

To deal with these issues, rotating fluidized beds (RFBs) have been developed since the late sixties. As most of the pre-cited limitations are related to the gravitational force balancing the drag, rotating fluidized beds operate in a high-G field. The latter is achieved by the fast rotation of the particle bed in a cylindrically shaped vessel. The radial fluidization of the particle bed is ensured by balancing the radially outward centrifugal force and the gas-solid drag force resulting from radially inward flowing gas. As a result, the terminal velocity of the particles can be increased by increasing the bed rotational speed and higher gas-solid slip velocities can be achieved, leading to intensified interfacial heat, mass and momentum transfer rates. Furthermore, a rotating fluidized bed allows reaching high gas-solids slip velocities while maintaining a uniform bed with a high solids volume fraction, i.e. particle bed density. Finally, the centrifugal and drag forces being significantly higher

in RFBs than the gravitational and drag forces in conventional FBs, the Van der Waals forces between fine and light particles become negligible in RFBs, allowing their fluidization.

The rotation of the particle bed can be generated by different ways. In so-called rotating fluidized beds in a rotating geometry [51–55] (Figure 2.1a), a motor is used to make the fluidization chamber rotating around its axis of symmetry. By contrast, rotating fluidized beds in a static geometry (Figure 2.1b) can be generated in vortex or whirl chambers. In this case, the fluidization chamber is static and the fluidizing gas is injected tangentially, generating the particle bed rotation. The use of a motor presents a better operating flexibility because the particle bed rotational speed can be controlled independently of the gas flow rate. However, the rotating chamber can lead to mechanical vibrations and sealing problems, making the scale-up of RFBs in a rotating geometry challenging for industrial applications. Furthermore, the feeding and withdrawal of reagents or products are not straightforward, making difficult a continuous operation. Vortex chambers, having no moving part, suit better for the industry but do not allow the independent control of the particle bed rotational speed and the gas flow rate. The coupling factor depends on the vortex chamber design and on the type of particles to be fluidized. Consequently, a vortex chamber is commonly designed for a given type of particles and offers a limited flexibility around the characteristics of this type of particles. It should be remarked that the Geldart classification of particles based on their fluidization behaviour is only valid in a 1G field. Qian et al. [19] showed that in the presence of a high-G field, the classification is shifted. Consequently, the original Geldart classification will be referred as 1G-Geldart in what follows.

2.2 Generation of rotating fluidized beds in vortex chambers

In a vortex chamber, the process gas is injected tangentially via multiple gas inlet slots (Figure 2.2a) in the outer wall of the cylindrical vortex chamber. When particles are located in the vortex chamber, the tangential momentum of the gas is transferred to the particles, resulting in the formation of a rotating particle bed. As shown in Figure 2.2b, the gas is forced to move radially in order to leave the vortex chamber via its unique outlet, a centrally positioned chimney. Because of their inertia,

2.2. Generation of rotating fluidized beds in vortex chambers

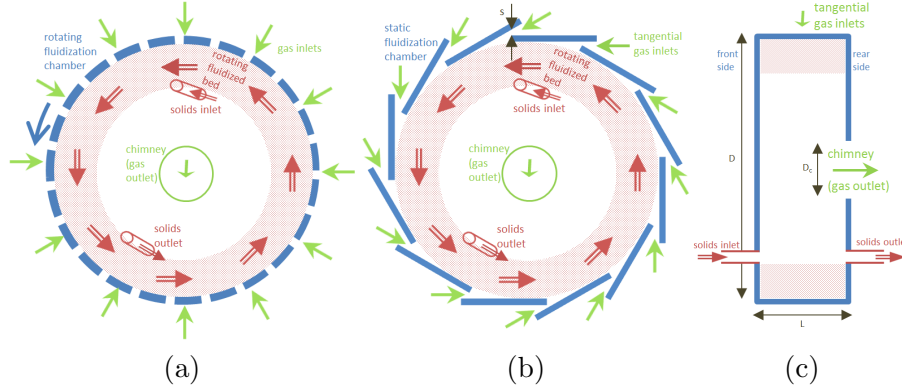


Figure 2.1: (a) Rotating fluidized bed in a rotating geometry. (b) Front view and (c) side view of a rotating fluidized bed in a static geometry (vortex chamber).

the particles do not follow the gas and remain between the outer wall of the cylindrical vessel and the freeboard region, the inner zone where no particle bed is observed. At a certain point, particles can be entrained by the gas in the freeboard region and leave the vortex chamber through the chimney. Particles entrainment occurs when the radially inward drag force exceeds the radially outward centrifugal force. In case of entrainment, the Coriolis effect may prevent the particles from going out the vortex chamber by converting its radially inward momentum to a tangential and next radially outward momentum. Balancing the drag and centrifugal forces requires a careful design of the vortex chamber, including the gas inlet slots and the chimney. As previously mentioned, the design depends on the characteristics of the particles to be fluidized. Radial fluidization is, however, not always required as tangential fluidization is ensured. Typically, the solids are injected while the gas flow is already established to avoid direct entrainment through the chimney. Finally, it should be noticed that an independent control of the gas and solids residence times requires a separate solids outlet.

Vortex chambers must not be confused with torbed reactors (Figure

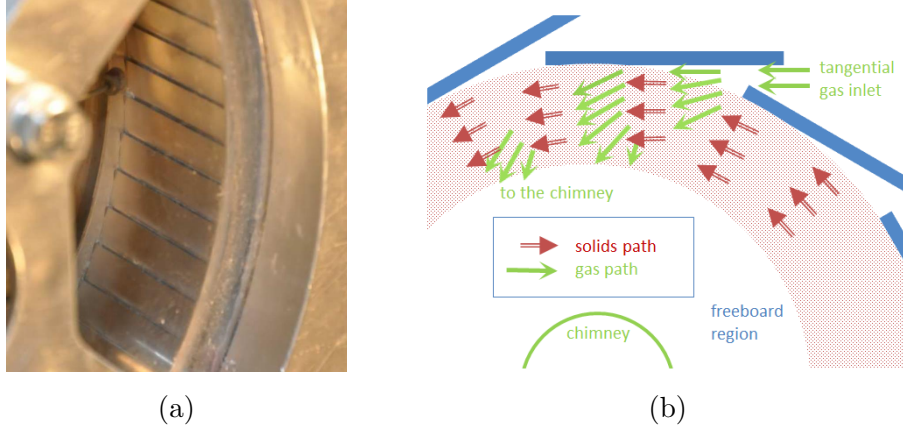


Figure 2.2: (a) Detail view of the gas inlet slots of a 72, 0.5 mm slots vortex chamber. (b) Gas and solids paths in a vortex chamber.

2.3). In the latter, a horizontal annular fluidized bed is suspended by a gas flow. The gas inlet slots are located at the bottom of the torbed and are inclined in such a way as both to fluidize the particles and to make the particle bed rotate [56]. The bed is in this case axially (vertically) and tangentially fluidized, in contrast with vortex chambers where the bed is radially and tangentially fluidized. Consequently, the axial fluidization is a gravitational fluidization comparable to that of conventional fluidized beds and, as a result, torbeds do not offer the advantages of high-G fluidization.

2.3 History

Gas-solid rotating fluidized beds in vortex chambers were first considered in the late sixties for drying solids material [57,58] and as nuclear reactor for rockets [59]. Because of the poor understanding of the technology, other applications were not considered until De Wilde and de Broqueville recently studied their hydrodynamic behaviour [60–62]. Gas-liquid [63, 64] and liquid-solid [65, 66] flows in vortex chambers have also been examined.

2.4. Hydrodynamic characteristics of rotating fluidized beds in vortex chambers

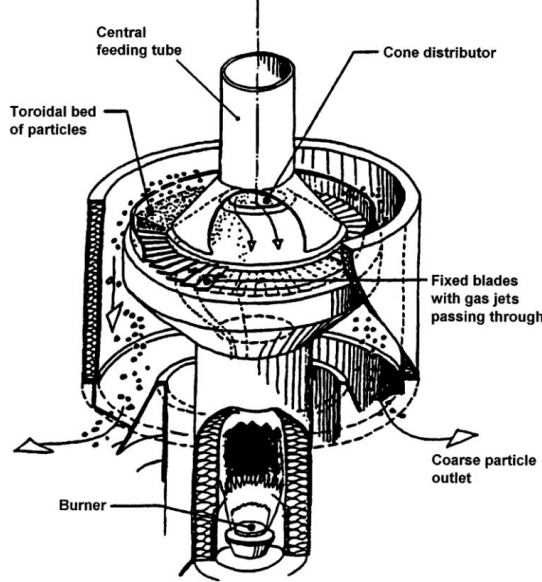


Figure 2.3: Scheme of a torbed reactor [56].

2.4 Hydrodynamic characteristics of rotating fluidized beds in vortex chambers

2.4.1 Radial and tangential fluidization in vortex chambers

Because the rotation of the particle bed is induced by the tangential fluidization of the gas, the particles are always tangentially fluidized in vortex chambers. Consequently, maintaining a particle bed in a vortex chamber does not necessarily require radial fluidization. Moreover, radial fluidization may become undesirable as it can lead to a reduction of the bed density and an increased presence of particles in the freeboard region and related losses through the chimney. Nevertheless, radial fluidization is required to avoid permanent contacts between particles in some applications, e.g. particles coating. Radial fluidization occurs when the radial gas-solid drag and particles inertial (centrifugal and Coriolis) forces are balanced. If the intensity of the inertial forces exceeds that of the radial gas-solid drag, the particle bed remains in the vortex chamber but no radial fluidization occurs. The relative intensity of the two counteracting forces depends on the particles characteristics

and on the vortex chamber design [60, 62], including the design of the gas inlet and outlet (chimney), the solids inlet and outlet and the chamber shape and dimensions (length L , diameter D , chimney diameter D_c , number of slots n , slot size s and fraction of the chamber circumference taken by gas inlet slots $\lambda = n \cdot s / (\pi D)$, see Figure 2.1).

Figure 2.4 presents the fluidization behaviours obtained with two different types of particles for the same chamber design and same gas flow rate. Radial fluidization is not observed with the 1G-Geldart-D type HDPE polymer pellets (Figure 2.4a). The particle bed is dense and is only tangentially fluidized, forming a so-called concentrated pseudo-fluidized layer. On the other hand, with the same vortex chamber and same operating conditions, the smaller 1G-Geldart-B alumina particles are radially fluidized, resulting in a less dense particle bed (Figure 2.4b). Whether radial fluidization occurs can be determined by measuring the evolution of the pressure drop over the particle bed with the gas flow rate (Figure 2.5). For HDPE polymer particles, no radial fluidization occurs, meaning that the pressure drop over the particle bed is insufficient to compensate for the weight of the particle bed in the high-G field. Consequently, the particle bed is radially packed and at least partially supported by the outer cylindrical wall of the vortex chamber. By analogy with the behaviour of the conventional fixed bed shown in Figure 1.3, the pressure drop of the particle bed is thus increasing with the gas flow rate (Figure 2.5a). By contrast, the pressure drop over the bed becomes independent of the gas flow rate when radial fluidization occurs (Figure 2.5a), as it is the case in conventional fluidized beds.

2.4.2 Flow pattern with and without solids loading

When air is injected in a vortex chamber in the absence of solids, a free vortex flow pattern is observed. That means that the circulation $\Gamma = u_{ig,t} \cdot r$ is constant everywhere in the vortex chamber except in the central region where the gas is evacuated through the chimney. The tangential velocity of the air is thus inversely proportional to its radial position. This is shown on figure 2.6a [68] where experimental data of the normalized circulation are plotted versus the normalized radial position in the vortex chamber. The same figure also shows a different flow pattern in the presence of 2 mm sand particles. In the presence of a solid phase, the free vortex is maintained in the freeboard region while in the outer region where the particle bed is located, the free vortex is

2.4. Hydrodynamic characteristics of rotating fluidized beds in vortex chambers

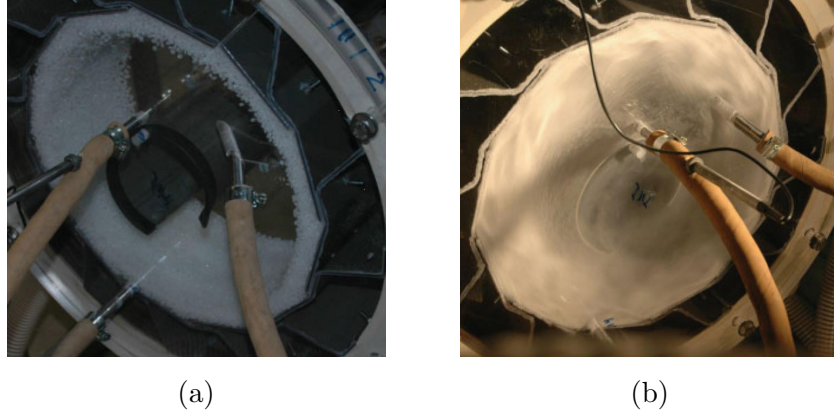


Figure 2.4: Fluidization behaviour in a vortex chamber of (a) HDPE polymer particles ($\rho_s = 950 \text{ kg/m}^3$, $d_p = 3.5 \text{ mm}$) and (b) alumina particles ($\rho_s = 2100 \text{ kg/m}^3$, $d_p = 350 \text{ }\mu\text{m}$). Vortex chamber: $D = 0.36 \text{ m}$, $D_c = 0.12 \text{ m}$, $L = 0.135 \text{ m}$, $n = 12$, $s = 4 \text{ mm}$, $\lambda = 0.042$. Gas flow rate of $0.2 \text{ m}^3/\text{s}$ and solids loading of 1-1.5 kg. Continuous operation with side solids inlet and outlet [60].

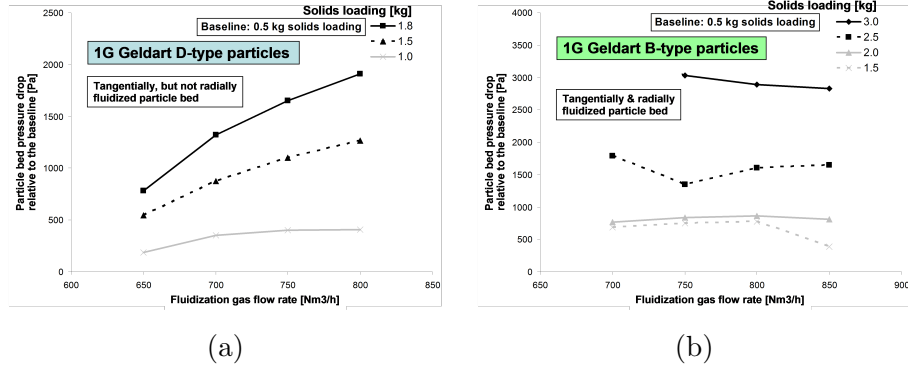


Figure 2.5: Evolution of the pressure drop over the particle bed with the gas flow rate while fluidizing in a vortex chamber (a) HDPE polymer particles ($\rho_s = 950 \text{ kg/m}^3$, $d_p = 3.5 \text{ mm}$) and (b) salt particles ($\rho_s = 2200 \text{ kg/m}^3$, $d_p = 350 \text{ }\mu\text{m}$). Vortex chamber: $D = 0.24 \text{ m}$, $D_c = 0.16 \text{ m}$, $L = 0.115 \text{ m}$, $n = 24$, $s = 2.3 \text{ mm}$, $\lambda = 0.073$. Continuous operation with side solids inlet and rotating chimney outlet [67].

destroyed and a flow pattern between free vortex and solid body type ($u_{ig,t}$ is proportional to r so that the rotational speed $\omega = u_{ig,t}/r$ is constant) is observed. The destruction of the free vortex by solids injection can be easily detected as it comes with the decrease of the pressure drop over the vortex chamber [67].

The real flow pattern in the particle bed and in the freeboard regions actually depends on the relative intensity of the radial gas-solid drag force and the particles inertial forces because it governs the radial fluidization and related bed density, the entrainment of particles in the freeboard region and the potential formation of meso-scale non-uniformities, i.e. bubbles. As previously mentioned, the relative intensity of the two forces is affected by the vortex chamber design and the characteristics of the particles to be fluidized. With large 1G-Geldart D-type particles (HDPE polymer particles used in Figure 2.4a), the dense, radially not fluidized bed rotates nearly like a solid body, even if fast mixing of particles is observed in the radial direction. The flow pattern is more complex with 1G-Geldart B-type particles (alumina particles used in Figure 2.4b) as bubbles are observed in the radially fluidized bed. With even smaller particles from 1G-Geldart group A, the radial fluidization is more pronounced resulting in a much less dense bed where particles are easily entrained in the freeboard region. As a result, there is no clear demarcation between the particle bed and the freeboard. The resulting flow pattern can be explained from Figure 2.6b [59] showing the rotational speed of a 20 μm talc particles bed at different vortex chamber radial positions as a function of the solids loading. At low solids loadings (between 0.2 and 0.5 kg), the bed rotational speed is seen to increase with decreasing radial position, indicating a flow pattern approaching that of a free vortex resulting from the low density particle bed. At higher solids loadings, the rotational speeds at different radial positions are seen to converge, especially near the outer cylindrical wall, indicating a progressive switch to a solid body type flow pattern. It should be noticed that the rotational speed logically decreases with the solids loading at each radial position, as already seen in Figure 2.6a.

The paths taken by the gas and solid phases in vortex chambers have been recently studied both numerically, with CFD (Figure 2.7a), and experimentally by De Wilde [67]. For the latter, smoke was injected through one of the gas inlet slots while keeping air injection through the

2.4. Hydrodynamic characteristics of rotating fluidized beds in vortex chambers

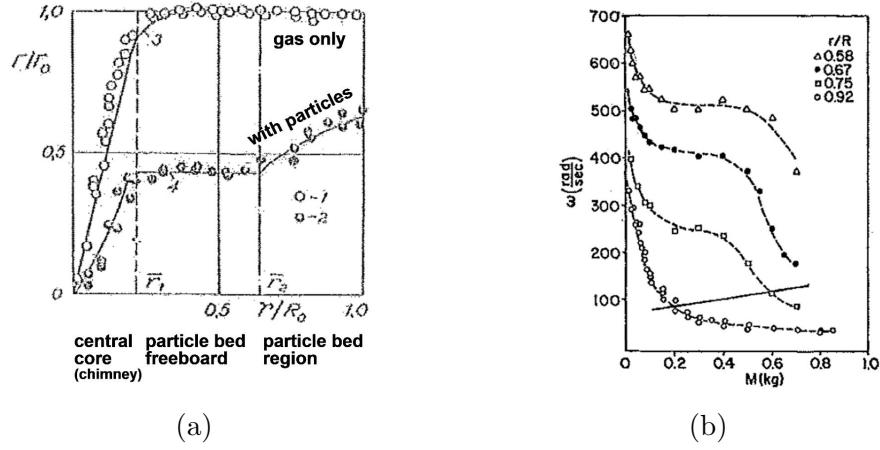


Figure 2.6: (a) Normalized circulation $\Gamma = u_{g,t} \cdot r$ versus the normalized radial position (subscript ⁰ indicates at the vortex chamber outer wall, for gas only) in the absence and presence of particles. Vortex chamber: $D = 0.36$ m, $D_c = 0.08$ m, $L = 0.06$ m, $n > 20$, $\lambda = 0.099$. Solids loading: 1 kg of sand particles ($\rho_s = 1900$ kg/m³, $d_p = 2$ mm) [68]. (b) Rotational speed of a talc particles ($\rho_s = 2700$ kg/m³, $d_p = 20$ μ m) bed at different vortex chamber radial positions as a function of the solids loading. Vortex chamber: $D = 0.305$ m, $D_c = 0.15 \rightarrow 0.08$ m (convergent), $L = 0.063$ m, $n = 12$, $s = 0.3$ mm, $\lambda = 0.00376$. [59].

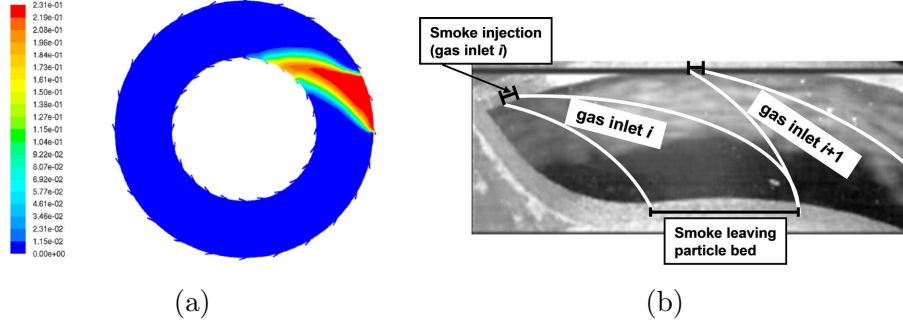


Figure 2.7: Tracer study of the gas phase flow path through a rotating bed of particles in a vortex chamber. (a) CFD simulation with a red tracer injected via two of the gas inlet slots [67]. (b) Experimental study with smoke injection via one of the gas inlet slots [69].

others and its path was visualized by means of a rapid camera (Figure 2.7b). These tracer studies demonstrated that the tangential velocity of the gas rapidly decreases upon contact with the particle bed, resulting in a radial deflection of the gas path towards the chimney. This deflection was shown to be accentuated when the gas approaches the next gas inlet slot in such a way that the gas injected through one slot is barely mixed with the gas injected through the next one. This observation can be exploited for multi-zone operation of vortex chambers (see section 2.7). Because of the inertial forces, the particles are unable to follow the gas path and form, as previously mentioned, a radially packed or fluidized bed depending on the particles characteristics and the vortex chamber design.

2.4.3 Operating flexibility with respect to the gas flow rate and the solids loading

A unique feature of rotating fluidized beds in vortex chamber is the operating flexibility with respect to the gas flow rate. This can be explained as the radial gas-solid drag and counteracting centrifugal forces are affected in a similar way by the gas flow rate. Indeed, while in-

2.4. Hydrodynamic characteristics of rotating fluidized beds in vortex chambers

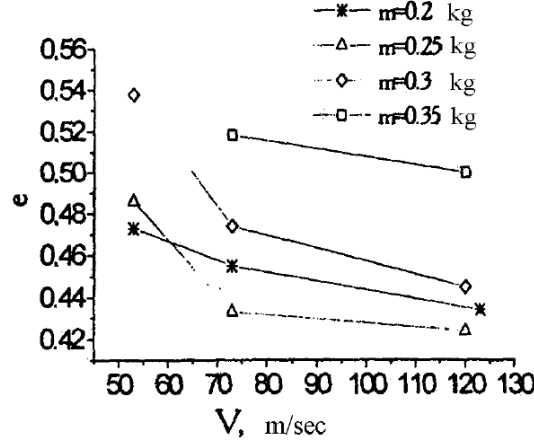


Figure 2.8: Evolution of the particle bed void fraction with the gas inlet velocity for different solids loadings in a vortex chamber used for grain drying [70].

creasing the gas-solid drag force, a higher gas flow rate also makes the particles rotating faster and, as a result, increases the centrifugal force. Depending on the chamber design and on the particles to be fluidized, the centrifugal force may be, however, more or less increased than the gas-solid drag force, resulting respectively in a slight bed contraction or expansion with increasing gas flow rate. Expansion with increasing gas flow rate comes generally at high gas flow rates because shear becomes prevailing and the bed rotational speed increase with the gas flow rate is less pronounced than at lower gas flow rates. Figure 2.8 shows the evolution of the particle bed void fraction with the gas inlet velocity for different solids loadings [70]. The latter demonstrates the operating flexibility with respect to the gas flow rate with only a limited contraction/densification of the bed with increasing gas flow rate for the given vortex chamber design and the range of operating conditions studied.

Sazhin and al. [71] theoretically demonstrated that the retention capacity of vortex chambers, i.e. the amount of solids remaining in the chamber after stopping the solids feeding, is independent of the gas flow rate if the gas flow rate is sufficiently high to form a stable rotating particle bed and to allow neglecting the effects of gravity. Oppositely, at lower gas flow rate for which the gravity cannot be neglected, the retention

capacity was shown to increase rapidly with increasing gas flow rate [71]. These results are consistent with previous experimental observations by Kochetov et al. [57, 58] and Anderson et al. [59]. The minor influence of the gas flow rate on the solids loading and particle bed stability in vortex chambers when operating at sufficiently high gas flow rate was also experimentally proven by De Wilde and de Broqueville [60, 62].

These authors also derived theoretical expressions of the centrifugal and gas-solid drag forces as well as the minimum and terminal velocities in vortex chambers as a function of the total volumetric gas flow rate Q_g [27]. The centrifugal force, expressed per unit volume fluidization chamber, is given by

$$f_c = \varepsilon_s \rho_s a_c \quad (2.1)$$

The centripetal acceleration a_c can be expressed as

$$a_c = \frac{u_{is,t}^2}{r} = \frac{(s_{gs} u_{ig,t})^2}{r} \quad (2.2)$$

where $u_{ig,t}$ and $u_{is,t}$ are the tangential interstitial velocities of respectively the gas and the solid phases and s_{gs} is the tangential gas-solid slip factor, assumed to be constant.

The tangential velocity of the gas is related to the total volumetric gas flow rate Q_g by

$$u_{ig,t} = \frac{Q_{g,t}}{S_t} = \frac{Q_g n_{rot}}{S_t} = \frac{Q_g n_{rot}}{\varepsilon_g ((D/2) - R_f) L} \quad (2.3)$$

with $Q_{g,t}$ the tangential volumetric gas flow rate, n_{rot} the number of rotation of the gas phase in the particle bed, S_t the surface area available for the tangential motion of the gas phase in the particle bed and R_f the radial position of the freeboard.

Combining equations (2.2) and (2.3) and averaging over the particle bed volume gives

2.4. Hydrodynamic characteristics of rotating fluidized beds in vortex chambers

$$\begin{aligned}
\bar{a}_c &= \frac{Q_g^2 \bar{n}_{rot}^2 s_{gs}^2}{\bar{\varepsilon}_g^2 ((D/2) - R_f)^2 L^2 \bar{r}} = \frac{Q_g^2 \bar{n}_{rot}^2 s_{gs}^2}{\bar{\varepsilon}_g^2 ((D/2) - R_f)^2 L^2 \bar{r}^2} \bar{r} \\
&= \frac{4Q_g^2 \bar{n}_{rot}^2 s_{gs}^2}{\bar{\varepsilon}_g^2 ((D/2) - R_f)^2 L^2 ((D/2) + R_f)^2} \bar{r} = \frac{4Q_g^2 \bar{n}_{rot}^2 s_{gs}^2}{\bar{\varepsilon}_g^2 \left((D/2)^2 - R_f^2 \right)^2 L^2} \bar{r} \\
&= \frac{4\pi^2 Q_g^2 \bar{n}_{rot}^2 s_{gs}^2}{\bar{\varepsilon}_g^2 V_{bed}^2} \bar{r} = \left(\frac{2\pi Q_g \bar{n}_{rot} s_{gs}}{\bar{\varepsilon}_g V_{bed}} \right)^2 \bar{r}
\end{aligned} \tag{2.4}$$

The centrifugal force is consequently given by

$$\bar{f}_c = \bar{\varepsilon}_s \rho_s \left(\frac{2\pi Q_g \bar{n}_{rot} s_{gs}}{\bar{\varepsilon}_g V_{bed}} \right)^2 \bar{r} \tag{2.5}$$

The radial drag force per unit volume fluidization chamber can be expressed as

$$f_{d,r} = \beta (u_{ig,r} - u_{is,r}) \tag{2.6}$$

where the drag coefficient β in fluidized beds may be given by Gidaspow [72], combining the Wen and Yu [73] drag model valid for diluted beds and the Ergun [20] equation relating the drag and the pressure drop in denser beds.

$$\begin{aligned}
\beta &= \frac{3}{4} C_D \frac{\varepsilon_g \varepsilon_s}{d_p} \rho_g |\vec{u}_{ig} - \vec{u}_{is}| \varepsilon_g^{-2.65} \text{ for } \varepsilon_s \leq 0.2 \\
\beta &= 150 \frac{\varepsilon_s^2 \mu_g}{\varepsilon_g d_p^2} + 1.75 \frac{\varepsilon_s \rho_g}{d_p} |\vec{u}_{ig} - \vec{u}_{is}| \text{ for } \varepsilon_s > 0.2
\end{aligned} \tag{2.7}$$

As previously indicated in section 1.4.2, the drag coefficient can be correlated with the particles Reynolds number $Re_p = \frac{|\vec{u}_{ig} - \vec{u}_{is}| \varepsilon_g \rho_g d_p}{\mu_g}$.

$$\begin{aligned}
C_D &= \frac{24}{Re_p} (1 + 0.15 Re_p^{0.687}) \text{ for } Re_p \leq 1000 \\
C_D &= 0.44 \text{ for } Re_p > 1000
\end{aligned} \tag{2.8}$$

Other drag laws are given by Syamlal and O'Brien [74] and by Parmentier et al. [75].

	$\text{Re}_p < 0.4$	$\text{Re}_p > 1000$
$\varepsilon_s \leq 0.2$	$f_d \propto Q_g^1$	$f_d \propto Q_g^2$
$\varepsilon_s > 0.2$	$f_d \propto Q_g^1$	$f_d \propto Q_g^{1 \rightarrow 2}$

Table 2.1: Dependency of the drag force f_d with the gas flow rate.

The radial interstitial velocity of the gas phase is related to the gas flow rate by the surface area available for the radial motion of the gas phase.

$$u_{ig,r} = \frac{Q_g}{S_r} = \frac{Q_g}{\varepsilon_g 2\pi r L} \quad (2.9)$$

Neglecting the radial interstitial velocity of the solids with respect to that of the gas phase in equation (2.6) and averaging over the particle bed volume, the radial drag force can be written as

$$f_{d,r} = \beta \bar{u}_{ig,r} = \beta \frac{Q_g}{\varepsilon_g 2\pi \bar{r} L} \quad (2.10)$$

The relationship between the drag force and the gas flow rate is thus depending on the solids volume fraction and the particles Reynolds number. Table 2.1 shows the four different cases, assuming that $|\vec{u}_{ig} - \vec{u}_{is}| \propto u_{ig,r}$ and that ε_s is not significantly affected by the fluidization gas velocity.

For dilute beds ($\varepsilon_s \leq 0.2$) and sufficiently high particles Reynolds number, the centrifugal and drag forces are seen to be both affected in the same way by the total gas flow rate, indicating that the bed density should not be affected by the gas flow rate. For dense beds ($\varepsilon_s > 0.2$) and/or low particles Reynolds number, the centrifugal force increases more rapidly than the drag force with the gas flow rate, resulting in a contraction/densification of the particle bed with increasing gas flow rate. In both cases, the particle bed is maintained while increasing the gas flow rate.

The terminal and minimum fluidization velocities can be calculated by extending the expressions presented in section 1.4.2 from a gravity field to a centrifugal field. The terminal velocity is consequently given by

2.4. Hydrodynamic characteristics of rotating fluidized beds in vortex chambers

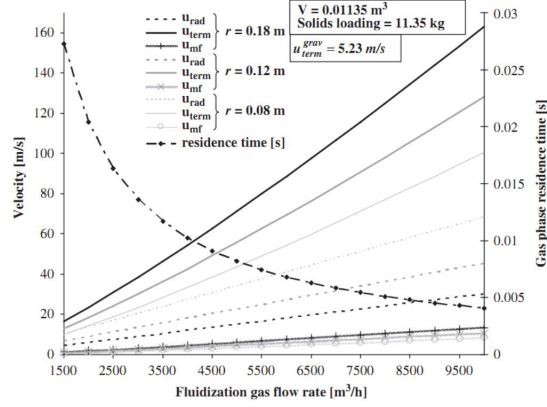


Figure 2.9: Illustration of the gas flow rate influence on the minimum fluidization velocity and on the terminal velocity of the particles in vortex chambers [27].

$$u_t = \sqrt{\frac{4\bar{a}_c d_p (\rho_s - \rho_g)}{3\rho_g C_D}} = \frac{2\pi Q_g \bar{n}_{rot} s_{gs} \sqrt{\bar{r}}}{\bar{\varepsilon}_g V_{bed}} \sqrt{\frac{4d_p (\rho_s - \rho_g)}{3\rho_g C_D}} \quad (2.11)$$

The minimum fluidization velocity can be calculated from [22]

$$u_{mf} = \frac{\mu_g}{d_p \rho_g} \left(33.7 \left[(1 + 3.6 \cdot 10^{-5} Ar)^{0.5} - 1 \right] \right) \quad (2.12)$$

with the Archimede number in the centrifugal field given by

$$Ar = \frac{\rho_g (\rho_s - \rho_g) d_p^3 \bar{a}_c}{\mu_g^2} = \frac{\rho_g (\rho_s - \rho_g) d_p^3 \bar{r}}{\mu_g^2} \left(\frac{2\pi Q_g \bar{n}_{rot} s_{gs}}{\bar{\varepsilon}_g V_{bed}} \right)^2 \quad (2.13)$$

Equations (2.11) and (2.12), which are plotted in Figure 2.9, show that both the terminal and the minimum fluidization velocities increase with the gas flow rate, demonstrating again the flexibility with respect to the gas flow rate in vortex chambers.

Rotating fluidized beds in vortex chambers also offer a certain flexibility with respect to the solids loading. Figure 2.6b shows that introducing particles into the vortex chamber reduces the rotational speed for all the different radial positions in the vortex chamber, and particularly in the dense particle bed region near the outer wall ($r/R = 0.92$).

Above a certain point and within a given range of solids loading, the rotational speed becomes independent from the solids loading. In this range, the particle bed characteristics are maintained, meaning that the vortex chamber exhibits a reasonable flexibility with respect to the solids loading. However, above a critical value of the solids loading, the rotational speed is seen to decrease sharply with increasing solids loading. Anderson et al. [59] explained this behaviour by the increasing momentum losses resulting from friction in the dense particle bed region. Moreover, Rosales et al. [76] demonstrated that the free vortex in the freeboard region accelerates and stabilizes the dense particle bed region. As a result, as the solids loading increases, more particles end up in the freeboard region, breaking the free vortex and its effects on the particle bed. Figure 2.8 [70] is another experimental work showing a relatively small effect of the solids loading on the particle bed density, confirming a certain flexibility of vortex chambers with respect to the solids loading. It should be noticed that, in this case, the solids volume fraction first slightly increases and then slightly decreases with the solids loading, confirming the increase of frictional losses and the reduction of the rotational speed when operating at high solids loading. De Wilde et al. [62], fluidizing continuously fed 1G-Geldart B and D type particles in a given vortex chamber, also observed a flexibility of the latter with respect to the solids loading. These authors also observed that hysteresis behaviours could be obtained when measuring the evolution of some flow characteristics with the solids loading [67].

2.4.4 Particle bed stability and large-scale non-uniformities

As previously discussed in section 1.4.2, large-scale non-uniformities like slugging or channelling are detrimental for the bed stability and for the gas-solids contact and related interfacial momentum, mass and heat transfer. As in conventional fluidized beds, these phenomena can occur with rotating fluidized beds in vortex chambers and are related to gas distribution non-uniformities. In vortex chambers, the latter can consist in difference of gas flow rate from one slot to another or within the axial direction of an individual slot [77]. Channelling in vortex chambers occurs when friction with the end walls results in a preferential path of the process gas through the axial center of the vortex chamber. As a result, the process gas is mainly entering the vortex chamber through the axial center of the inlet slots and bypasses the solids near the end walls. Slugging in vortex chambers corresponds to a tangentially non-uniform

2.4. Hydrodynamic characteristics of rotating fluidized beds in vortex chambers



Figure 2.10: Schematic representation of channelling and slugging in vortex chambers [62].

distribution of the particles which form a rotating slug in the vortex chamber. That means that the rotating particle bed does not occupy the whole circumference of the annulus. The process gas is entering mostly through the inlet slots that does not face the resistance of the particle bed and as a result, bypasses the solids [62]. While channelling can exhibit a stationary flow pattern, slugging evidently results in an unsteady flow pattern. Slugging or channelling can be detected through pressure measurements at different locations in the vortex chamber. Indeed, channelling will induce pressure differences along the axial direction whereas slugging will lead to pressure fluctuations in time [60]. A schematic representation of channelling and slugging in a vortex chamber is given in figure 2.10 [62].

Figure 2.11 presents an experimental map of the different flow regimes as a function of the solids loading and the gas flow rate for the fluidization of 1G-Geldart D polymer particles in a vortex chamber [62]. It can be seen that when increasing the solids loading, the flow regime evolves from channelling, slugging, to finally get a uniform and stable rotating particle bed. Consequently, the vortex chamber has to be designed to limit particles losses through the chimney in order to allow the build-up of a sufficiently high solids loading to obtain a stable particle bed. Figure 2.11 also reflects the flexibility of vortex chambers with respect to the gas flow rate as the latter has only a minor impact on the flow regime observed. Increasing the length of the vortex chamber can facilitate the occurrence of large scale non-uniformities in the axial direction (channelling). On the other hand, the axial uniformity can be improved by increasing the pressure drop over the gas inlet slots. For a given range of operating conditions, Kochetov et al. [57] found that

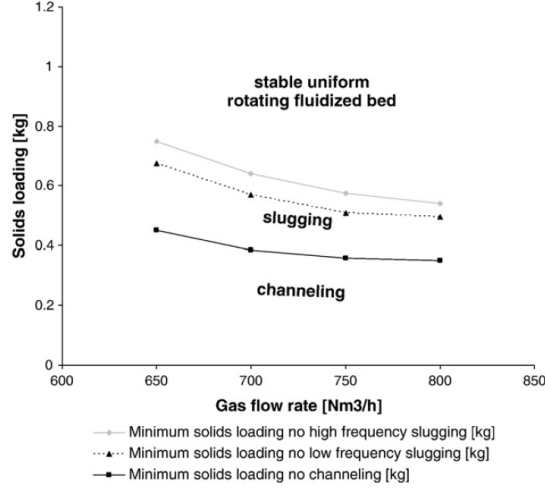


Figure 2.11: Flow regime map as a function of the solids loading and the gas flow rate obtained while fluidizing polymer particles ($\rho_s = 950$ kg/m³, $d_p = 3.5$ mm) in a vortex chamber with $D = 0.24$ m, $D_c = 0.15$ m, $L = 0.135$ m, $n = 24$, $s = 2.3$ mm, $\lambda = 0.073$ [62].

limiting the chamber length to the half of its diameter guarantees the axial uniformity of the particle bed. This ratio was confirmed by other experimental data [59, 60, 62, 67, 68, 76].

2.4.5 Meso-scale non-uniformities

In conventional fluidized beds, the presence of bubbles cannot be avoided, at least when fluidizing 1G-Geldart B and D type particles. As previously mentioned, the presence of bubbles in the particle bed results in poor gas-solids contact and related interfacial mass, heat and momentum transfer. High-G fluidization allows reducing or avoiding bubbles formation and improving the particle bed uniformity through the generation of a high pressure drop over the particle bed. The latter requires, however, both sufficiently high solids loading and particle bed rotational speed in the vortex chamber. Both the vortex chamber design and the particles characteristics can affect bubbles formation. Fluidization experiments carried out by De Wilde and de Broqueville [60] in a vortex chamber of given design revealed no bubbles formation with 1G-Geldart D particles, whereas bubbles significantly appeared with 1G-Geldart B

2.4. Hydrodynamic characteristics of rotating fluidized beds in vortex chambers

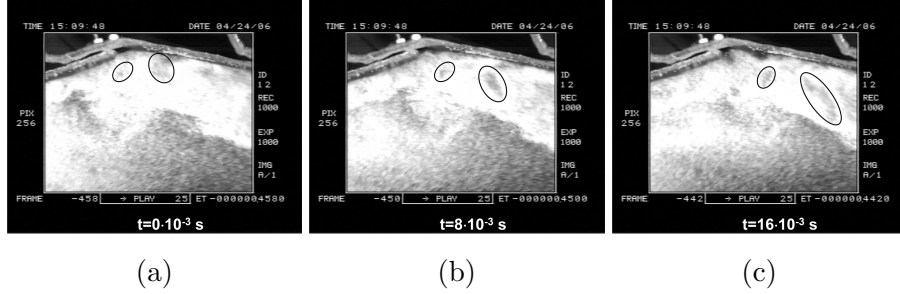


Figure 2.12: Visualization of bubbles formation and propagation in a particle bed of alumina particles ($\rho_s = 2100 \text{ kg/m}^3$, $d_p = 350 \text{ }\mu\text{m}$) through high speed CCD camera pictures at 1000 fps. Vortex chamber: $D = 0.36 \text{ m}$, $D_c = 0.12 \text{ m}$, $L = 0.135 \text{ m}$, $n = 12$, $s = 4 \text{ mm}$, $\lambda = 0.042$. Gas flow rate of $0.2 \text{ m}^3/\text{s}$ and solids loading of 1.1 kg [60].

particles. High speed CCD camera pictures (Figure 2.12) [60] showed that these bubbles were formed in the vicinity of the gas inlet slots and propagated both tangentially and radially. Rosales and De Wilde [76] demonstrated that bubbles formation could be avoided when fluidizing $70\text{--}85 \text{ }\mu\text{m}$ (1G-Geldart-A) FCC catalyst particles through a careful vortex chamber design.

2.4.6 Non-ideality of gas and solids phase distributions and flow patterns

As previously mentioned, the gas, injected tangentially through the inlet slots, is rapidly deflected in the radial direction in such a way that gases injected through adjacent slots are barely mixed. As a result, the gas phase motion can be described by the plug flow assumption if bubbles formation is limited or avoided. Oppositely, mixing is pronounced in the particle bed, both in the tangential direction because of the bed rotational motion and in the radial direction, especially for high intensities of shear and radial fluidization. Consequently, the particle bed can be considered well-mixed (CSTR). This was corroborated by experimental measurements of the response in terms of mass fraction of coloured particles in the particle bed to a step change in the colour of the 1G-Geldart D type particles fed to a vortex chamber of given design [69]. Indeed, Figure 2.13 demonstrates, for the given vortex chamber design and particles characteristics, that a CSTR behaviour of the response is

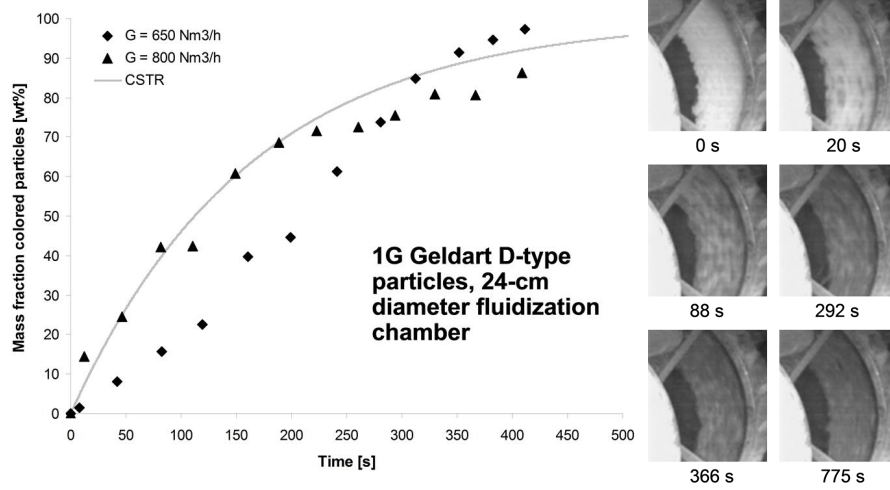


Figure 2.13: Step response study of solids mixing in a rotating fluidized bed in a vortex chamber using 1G-Geldart D type HDPE polymer particles ($\rho_s = 950 \text{ kg/m}^3$, $d_p = 3.5 \text{ mm}$): response in terms of mass fraction of coloured particles in the particle bed to a step change at $t = 0 \text{ s}$ in the colour of the particles fed for gas flow rates of 650 and 800 Nm^3/h . The solid line is the theoretical curve of the CSTR behaviour. Vortex chamber: $D = 0.24 \text{ m}$, $D_c = 0.16 \text{ m}$, $L = 0.115 \text{ m}$, $n = 24$, $s = 2.3 \text{ mm}$, $\lambda = 0.073$. Continuous operation with side solids inlet and rotating chimney outlet [69].

obtained if the gas flow rate and related particle bed rotational speed are sufficiently high to generate a dense and uniform particle bed. If reactions occur in the vortex chamber, a CSTR model can still be used if the time scale related to the mixing is significantly smaller than the time scale related to the reaction, that is, if the particles can make multiple rotations in the vortex chamber while reacting [5]. A model assuming plug flow for the gas phase and a CSTR behaviour for the solid phase was used by Staudt et al. [78] to study the potential of vortex chambers for the low-temperature pyrolysis and gasification of biomass particles.

However, a dense and uniform particle bed is not always obtained. When fluidizing small and/or light particles, the bed is radially fluidized and bubbling can occur, leading to time dependant solid volume fraction ε_s profiles in the particle bed in the tangential, radial and axial directions (see Figure 2.12). Anderson et al. [59] measured the particle bed density

2.4. Hydrodynamic characteristics of rotating fluidized beds in vortex chambers

$\rho_b = \varepsilon_s \cdot \rho_s$ by X-ray absorption at different locations while fluidizing talc particles ($\rho_s = 2700 \text{ kg/m}^3$, $d_p = 20 \text{ }\mu\text{m}$) in a vortex chamber. Figure 2.14b shows the tangentially-, axially- and time-averaged radial profiles of the particle bed density for different solids loadings while 2.14c shows an axially-averaged typical particle bed density contour. The particle bed density is seen to be strongly non-uniform, both in the radial and tangential directions. The gas injection design of the vortex chamber used in this study has a strong impact on the tangential profile of the particle bed density near the outer cylindrical wall. Indeed, lower densities are observed where the gas is injected while solids accumulation can be seen upstream of a gas inlet slot. This accumulation of particles results in the radial deflection of the gas injected through the previous slot, as previously discussed in section 2.4.2.

Axial non-uniformities can also occur. As previously mentioned in section 2.4.4, axial non-uniformities, of which channelling is the extreme case, are due to friction with the end plates of the vortex chamber resulting in a boundary flow at this location. Volchkov et al. [68] measured the axial profiles of the radial gas velocity in a 26 mm long vortex chamber with and without 3.5 mm diameter wheat grain particles. It can be observed from Figure 2.15 that the flow pattern is axially non-uniform in the absence and presence of particles. Moreover, the axial uniformity is improved in the presence of the particle bed for the given particles characteristics and vortex chamber design. The reduction of the bed rotational speed near the end walls can still result in the generation of a secondary, radially inward, flow of particles. The particles are consequently moving towards the chimney along the end plates of the vortex chamber, resulting in particles losses through the chimney. This phenomenon, which is seen to grow with decreasing particles size, can be reduced by a careful design of the vortex chamber (see section 2.5).

Finally, when working with particles which are poly-disperse in nature (size, density), segregation can occur in the vortex chamber because of the dependence of the drag and centrifugal forces on the particles characteristics. This phenomenon can be exploited in vortex chambers for certain processes such as grain cleaning, in which impurities such as seeds are separated from grain particles [70].

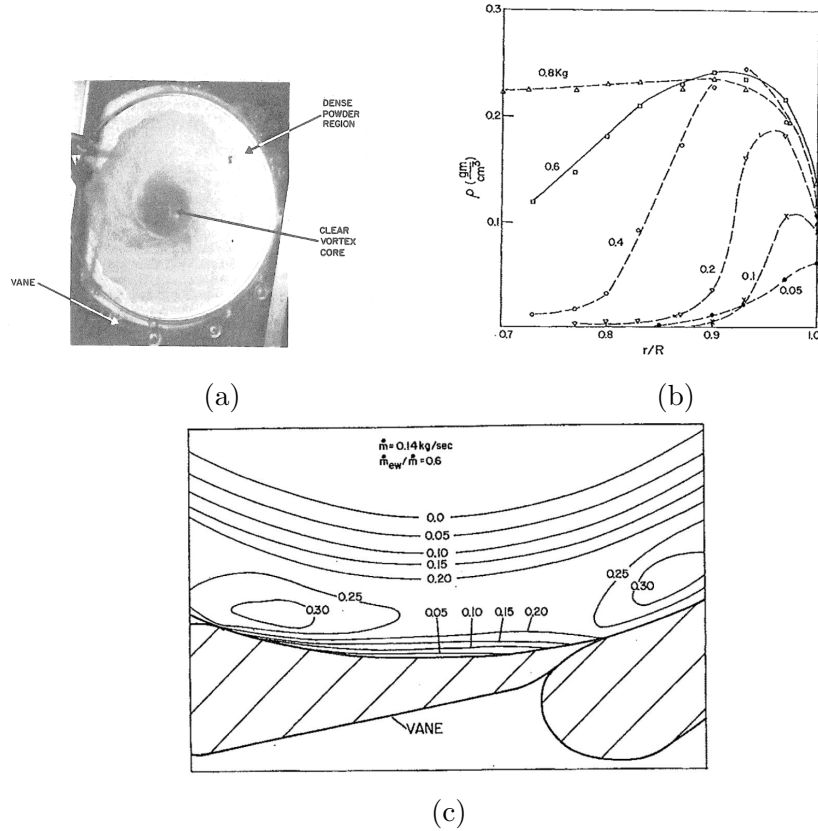


Figure 2.14: Particle bed density measurements by X-ray absorption (100 kV, 15 mA, typical exposure time = 1 s) at different locations while fluidizing talc particles ($\rho_s = 2700 \text{ kg/m}^3$, $d_p = 20 \text{ }\mu\text{m}$) in a vortex chamber. (a) Picture of the particle bed; (b) Tangentially-, axially- and time-averaged radial profiles of the particle bed density for different solids loadings; (c) Axially-averaged typical particle bed density contour for a solids loading of 0.4 kg. Vortex chamber: $D = 0.305 \text{ m}$, $D_c = 0.15 \rightarrow 0.08 \text{ m}$ (convergent), $L = 0.063 \text{ m}$, $n = 12$, $s = 0.3 \text{ mm}$, $\lambda = 0.00376$. Solids loadings between 0.05 and 0.8 kg and total gas flow rate of $0.11 \text{ Nm}^3/\text{s}$ [59].

2.5. Vortex chamber design

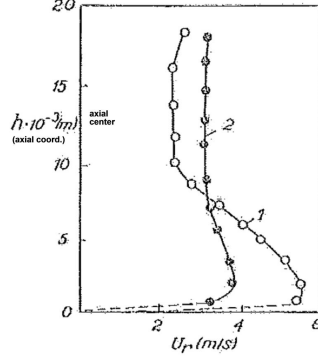


Figure 2.15: Experimental measurements of the axial profiles of the gas radial velocity in the (1) absence and (2) presence of wheat grain particles ($\rho_s = 1200 \text{ kg/m}^3$, $d_p = 3.5 \text{ mm}$) in a vortex chamber with $D = 0.20 \text{ m}$, $D_c = 0.10 \text{ m}$, $L = 0.026 \text{ m}$, $n > 20$, $\lambda = 0.0518$ [68].

2.5 Vortex chamber design

The previous section described the different hydrodynamic characteristics that can be observed in a vortex chamber depending on its design, on the particles characteristics and on the operating conditions. In particular, it was demonstrated that a careful vortex chamber design is crucial to obtain a dense, stable and uniform particle bed as well as to avoid too pronounced solid losses through the chimney. Table 2.2 summarises the vortex chamber designs and operating conditions used in different experimental studies of gas-solid fluidized beds in vortex chambers reported in the literature. This section presents guidelines to design the different parts of the vortex chamber based on the experimental observations made in these studies.

2.5.1 Gas inlet design

As briefly explained in section 2.4, the balance between the centrifugal and drag forces can be modified by the gas inlet slot design, i.e. the number of gas inlet slots n and their width s . Whereas the number of slots has to be sufficiently high to ensure a tangentially uniform gas and solid distribution, the fraction of the chamber circumference taken by gas inlet slots $\lambda = \frac{n \cdot s}{\pi \cdot D}$ determines the gas injection velocity in the vortex chamber for a given gas flow rate and the related ratio between the centrifugal and drag forces. Consequently, the generation of a stable

Chapter 2: Rotating fluidized beds in vortex chambers

Ref.	Vortex / Distr. chamber diameter (D_c / D^*)	Chimney diameter (D_c)	Vortex chamber length (L)	Number of gas inlet slots (n)	Gas inlet slot width (s)	$\lambda = n \cdot s / (\pi D)$	Gas flow rate (Q_g)	Gas phase res. time (τ_g)	Gas inlet velocity ($u_{ig, in}$)	Sup. radial gas velocity ($u_{sg, r}$)	Particle density (ρ_p)	Mean particle diameter (d_p)
	[m]	[m]	[m]		[m]		[m ³ /s]	[s]	[m/s]	[m ³ /(m ² s)]	[kg/m ³]	[m]
Rochetov et al., 1969	0.12* 0.12* 0.24* " 0.48* 0.82*	0.024-0.096 0.03 0.048-0.192 (0.06) 0.06 (0.11)	0.024-0.12 0.035 0.048-0.24 0.07 0.035 (0.128)	1, 5 (uneven distr.), 6 " " " (6)	" " " (0.0034)	0.016-0.0637 " " 0.0265	0.01-0.028 " " "	" 0.014-0.039 0.112-0.315 0.057-0.158	" " " "	0.76-2.14 0.19-0.53 0.38-1.06	MSN styrene copolymer ($\neq$ fractions): $\pm$ 1000 quartz sand: 1201 poppy seeds: $\pm$ 940 millet: $\pm$ 1000	Range: 100 E-6-2.5 E-3
Reported optimum: $D_c = 0.3 - 0.5 \times D$, chimney entry length: $0.1 - 0.2 \times L$ and $L < 0.5 \times D$. Optimum $n \uparrow$ as $d_p \downarrow$.												
Rochetov et al., 1969	0.12* 0.24* " "	(0.03) (0.06) " "	(0.0348) (0.035) " "	1, 5 (uneven distr.), 6 1 5 (u.d.) 6	/ 0.015 0.003 0.0025	/ 0.0637 0.0199 " "	0.01-0.028 " " " "	0.014-0.039 " 0.057-0.158 " "	" " " "	0.76-2.14 0.38-1.06 " "	MSN copo., Sulfanilamide, Melalite, Polystyrene: 5 fractions: (0.25-0.5) E-3 (0.5-1.0) E-3 (1.2-1.5) E-3 (1.5-2.0) E-3 (2.0-2.5) E-3	MSN copo., Sulfanilamide, Melalite, Polystyrene: 5 fractions: (0.25-0.5) E-3 (0.5-1.0) E-3 (1.2-1.5) E-3 (1.5-2.0) E-3 (2.0-2.5) E-3
Anderson et al., 1972	0.305	0.15 $\rightarrow$ 0.08 (convergent) (second. gas injection)	0.063	12	0.0003	0.0038	0.042-0.15	0.031-0.110	184-661	0.69-2.485	Zinc: 7000 Tungsten: 19100 Talc: 2700	Zinc: 10 E-6 Tungsten: 12 E-6 Talc: 20 E-6
Volchkov et al., 1993	0.2	0.1 (Flat/hyperbolic)	0.026	> 20	"	0.0518	0.046-0.073	0.011-0.018	"	2.8-4.5	Wheat grain: 1200	Wheat grain: 2.5 E-3
Volchkov et al., 1993	0.36 " "	0.08 " (Flat/hyperbolic)	0.026 0.06 "	> 20 " "	" " "	0.0991 " "	0.01-0.1 " "	0.03-0.27 0.06-0.61 "	" " "	0.34-3.4 0.15-1.47 "	Sand: 1900-2000 Wheat grain: 1200 Corn: 1200 Plastic: 1060-1100	Sand: (0.2-2) E-3 Wheat grain: 3.5 E-3 Corn: 8.0 E-3 Plastic: 3.0 E-3
Volchkov et al., 2003	0.2	"	0.036	"	"	0.0322	0.036-0.051	0.022-0.031	50-70	1.61-2.25	Wheat grain: 1200	Wheat grain: 3.5 E-3
De Wilde & de Broqueville, 2007	0.36	$\pm$ 0.12 (Slotted with slot size: 0.09 m)	0.135	12	0.004	0.0424	0.2	0.0687	31	1.31	HDPE: cyl.: 950 Alumina: 2100	HDPE: cyl.: d/1 = 5/2 E-3 Alumina: (300-400) E-6
De Wilde et al., 2007; De Wilde & de Broqueville, 2008	0.24 0.36	0.15 (Slot: 0.08 m, with/out grid) 0.12 (Slot: 0.09 m)	0.135 "	12/24 12	0.0023 0.004	0.0366/0.0732 0.0424	0.2 "	0.0305 0.0687	54/27 31	1.96 1.31	HDPE: cyl.: 950 Salt: 2160	HDPE: cyl.: d/1 = 5/2 E-3 Salt: 350 E-6
De Wilde & de Broqueville, 2008, 2010	0.24	0.16 (rotating) slot size:	0.115	24	0.0023	0.0732	0.18-0.236	0.022-0.029	28-37	2.08-2.72	HDPE: cyl.: 950 Salt: 2200	HDPE: cyl.: d/1 = 5/2 E-3 Salt: 350 E-6
Dvornikov & Belousov, 2011	(i) 0.26 (ii) 0.26 $\rightarrow$ 0.23 (iii) 0.23 $\rightarrow$ 0.2	(i) 0.16 "	(i) 0.032 (ii) 0.012 (iii) 0.083	(i) 54 (ii) 40 (iii) 40/80	0.0015	(i) 0.0992 (i) $\pm$ 0.081 (i)+(ii)+(iii) 0.283-0.316	(i) $\pm$ 0.021 (i) $\pm$ 0.021 (i) $\pm$ 0.021	(i) $\pm$ 31.2 (i) $\pm$ 31.2 (i) $\pm$ 31.2	(i) $\pm$ 3.1 (i) $\pm$ 3.1 (i) $\pm$ 3.1	1420 1420 1420	2.5 E-3	
Combined cylindrical-conical design: (i) upper cylindrical; (ii) transit conical; (iii) lower conical.												
Rikapture et al., 2011	0.54 " "	0.15 " "	0.10 " "	36 " "	0.002 0.006	0.042 0.127	0.33-1.0 "	0.023-0.069 "	46-139 15-46	1.95-5.90 "	HDPE: 950 FCC cat.: 1500	HDPE: 3 fractions: 9 E-4, 1.6 E-3, 2.4 E-3 FCC cat.: 70 E-6
Rosales & De Wilde, 2011, 2012	0.24 " " " "	0.10 " " " "	0.05 " " " "	24 72 36 72 36	0.003 0.0005 0.0239 0.0191 0.0096	0.355 0.0477 0.0239 0.0191 0.0096	0.061-0.167 " " " 0.061-0.097	0.014-0.037 " " " 0.023-0.037	17-46 34-93 68-185 85-232 170-270	1.62-4.42 " " 1.62-2.58	HDPE: 950 FCC cat.: 1500	HDPE: 1 E-3 2 E-3 300 E-6 FCC cat.: 74 E-6

Table 2.2: Vortex chamber designs and operating conditions used in experimental studies of gas-solid fluidized beds in vortex chambers.

2.5. Vortex chamber design

particle bed with uniform axial gas and solids distribution, which requires a sufficiently high gas injection velocity, necessitates a sufficiently small slot size.

Kochetov et al. [57, 58] tested three different vortex chambers, respectively with a single gas inlet slot, 5 gas inlet slots located in the lower part of the vortex chamber and 6 uniformly distributed gas inlet slots and keeping λ constant. As seen from figure 2.16b [58], the usage of a single gas inlet slot results in strong tangential non-uniformities in the particle bed. Figure 2.16c [57] presents the evolution of the retention capacity with the gas flow rate in the three different vortex chambers. The strong dependence of the retention capacity with the gas flow rate is explained by the fact that operation was not truly high-G, the gravity not being negligible with respect to the drag and centrifugal forces. This is also the reason why the authors tested a configuration with 5 gas inlet slots located in the lower part of the vortex chamber. Focusing on the highest flow rate, the retention capacity does not depend anymore on the gas flow rate, indicating that gravity is become negligible. In this range of operation, the retention capacity is seen to increase with the number of gas inlet slots, indicating that the particle bed tangential uniformity is improved when increasing n . Fluidizing and drying different granular materials in vortex chambers, Kochetov et al. [57] showed the importance of using multiple gas inlet slots, especially while fluidizing fine particles. This was corroborated by Sazhin et al. [71] who theoretically demonstrated that a minimum of 4 inlet slots was required to ensure an axisymmetric radial outflow from the vortex chamber. These authors also found out, still theoretically, that for relatively large particles ($d_p \geq 100 \mu\text{m}$), a further increase of the number of the inlet slots was not useful to improve the particle bed uniformity and to increase its rotational speed. On the other hand, Volchkov et al. [68] claimed that at least 20 gas inlet slots were required to optimize the particle bed uniformity and rotational speed for given gas flow rate, particles characteristics and λ . The relationship between the number of slots and the particle bed uniformity and rotational speed was also studied by Rosales et al. [71, 76]. By fluidizing different types of particles in vortex chambers with 24, 36 and 72 slots and given λ , they experimentally demonstrated that increasing the number of slots, at least up to a certain number, improves the particle bed density and increases the particle bed rotational speed. The optimal number of slots was also found to increase

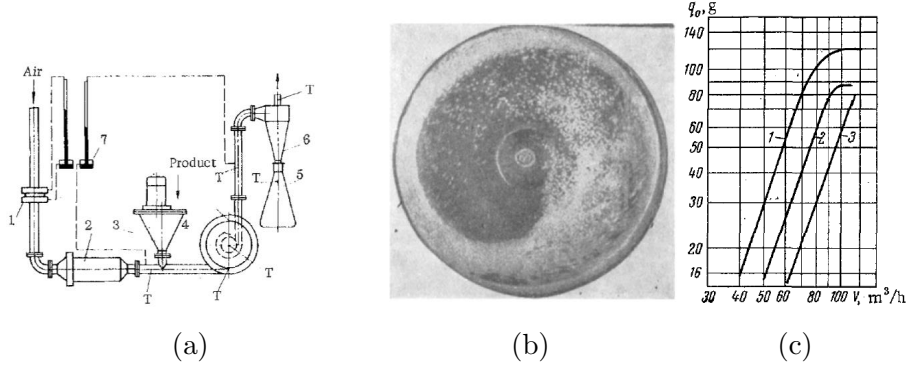


Figure 2.16: (a) Schematic representation and (b) picture of a vortex chamber with a single gas inlet slot through which the gas-solid mixture is tangentially injected. Vortex chamber with $D = 0.24$ m, $L = 0.035$ m, $n \cdot s = 15$ mm, $\lambda = 0.02$ [58]. (c) Solids retention capacity versus gas flow rate in a vortex chamber with given λ and with (1) 5 gas inlet slots located in the lower part of the vortex chamber, (2) a single gas inlet slot and (3) 6 uniformly distributed gas inlet slots. Vortex chamber with $D = 0.24$ m, $L = 0.035$ m, $n \cdot s = 20$ mm, $\lambda = 0.027$ [57].

with decreasing size and density of the particles to be fluidized.

Varying the fraction of the chamber circumference occupied by the gas inlet slots, λ allows modifying the ratio between the centrifugal and drag forces. As the intensity of these two forces are impacted by the particles characteristics, the optimal λ logically depends on the type of particles to be fluidized. Figure 2.17 focuses on the particle bed characteristics while fluidizing different types of particles in a vortex chamber of given λ . Figure 2.17a [58] shows the retention capacity of a given vortex chamber with $\lambda = 0.02$ versus the gas flow rate for polystyrene particles of different sizes while feeding no solids. At sufficiently high gas flow rate, i.e. when gravity is negligible, the retention capacity of the vortex chamber is seen to decrease significantly with decreasing particles size. This trend has been confirmed by Rosales et al. [76], fluidizing different

2.5. Vortex chamber design

types of particles in a vortex chamber with relatively large $\lambda = 0.096$. Figure 2.17b-d shows pictures of the particle beds obtained while continuously feeding solids for given solids and gas flow rates. The decrease of the retention capacity with decreasing particles size even leads to the impossibility to obtain a stable rotating fluidized bed for the smallest particles of FCC catalyst ($d_p = 70 - 85 \mu\text{m}$) in the vortex chamber with $\lambda = 0.096$. Kochetov et al. [57, 58] tested different vortex chambers with λ ranging between 0.016 and 0.064 for 1G-Geldart B and D type particles and concluded that the optimal λ was comprised between 0.02 and 0.038 depending on the particles size and density. However, it should be noticed that the operation was not fully high-G and that the gas inlet slots were not always uniformly distributed over the chamber circumference. Anderson et al. [59] carried out experiments with small but high density powders (talc with $d_p = 20 \mu\text{m}$ and $\rho_s = 2700 \text{ kg/m}^3$, zinc with $d_p = 10 \mu\text{m}$ and $\rho_s = 7000 \text{ kg/m}^3$ and tungsten with $d_p = 12 \mu\text{m}$ and $\rho_s = 19100 \text{ kg/m}^3$) in a vortex chamber with a much smaller $\lambda = 0.00376$. As seen from Figure 2.14a showing a rotating fluidized bed of talc particles, these authors obtained a uniform particle bed with almost no bubbles formation. De Wilde and de Broqueville [60, 61] confirmed that bubbling could be avoided when operating high-G at a sufficiently high solids loading.

A theoretical criterion, allowing to determine the required slot size and related lambda as a function of the characteristics of the particles to be fluidized was derived by Rosales and De Wilde [76]. This criterion confirms that the maximum allowable λ decreases with decreasing particles size and density. As a result, generating a rotating fluidized bed of fine and light powder in a vortex chamber may require very small slot widths s .

A revised version of this criterion is given below. To ensure that the particles are retained in the chamber, the centrifugal force F_c has to be higher than the counteracting gas-solid radial drag force $F_{d,r}$. The maximum allowable λ corresponds to the exact balance between these two forces. To determine a criterion for λ , the influence of the latter on F_c and $F_{d,r}$ has to be determined. The total centrifugal force over the particle bed is given by

$$F_c = \int_{V_{bed}} f_c \, dV = \int_{V_{bed}} \varepsilon_s \rho_s \frac{u_{is,t}^2}{r} \, dV \quad (2.14)$$

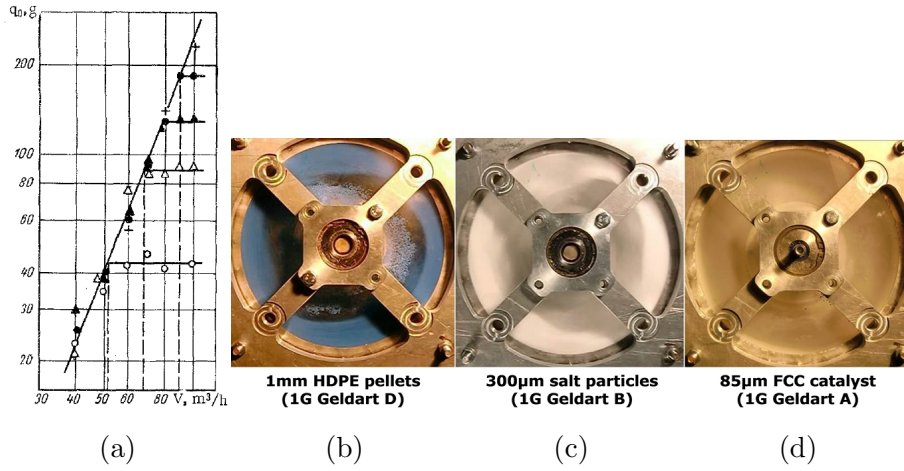


Figure 2.17: (a) Solids retention capacity versus the gas flow rate for polystyrene particles of different sizes ($\circ$: $d_p = 0.25 - 0.5$ mm; $\triangle$: $d_p = 0.5 - 1.0$ mm; $\blacktriangle$: $d_p = 1.2 - 1.5$ mm; $\bullet$: $d_p = 1.5 - 2.0$ mm; $+$: $d_p = 2.0 - 2.5$ mm) while feeding no solids in a vortex chamber with $D = 0.24$ m, $L = 0.035$ m, $n \cdot s = 15$ mm, $\lambda = 0.02$ [58]. (b-d) Pictures of the particle beds obtained while continuously feeding solids for given solids (8.35 g/s) and gas flow rates ($700 \text{ Nm}^3/\text{h}$) for different types of particles. Vortex chamber: $D = 0.24$ m, $D_c = 0.10$, $L = 0.05$ m, $n = 24$, $s = 3$ mm, $\lambda = 0.096$ [76].

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The latter depends on the rate of tangential momentum injected with the gas, which is transferred to the solids to make the particle bed rotate. Introducing $f_{t,m}$, the fraction on the tangential momentum transferred for the particle bed rotational motion, the solids tangential velocity $u_{is,t}$ can be calculated from

$$F_{d,t} = \int_{V_{bed}} f_{d,t} dV = \int_{V_{bed}} \beta (u_{ig,t} - u_{is,t}) dV = f_{t,m} \frac{\Delta p_{inj,t}}{\Delta t} \quad (2.15)$$

Assuming that the slots are purely tangential, the rate of injection of tangential momentum in the vortex chamber is given by

$$\frac{\Delta p_{inj,t}}{\Delta t} = \dot{m}_g u_{ig,inj} = \rho_g Q_g u_{ig,inj} = \rho_g n s L u_{ig,inj}^2 \quad (2.16)$$

where $u_{ig,inj}$ is the gas injection velocity. For turbulent and sufficiently dense gas-solid flows and assuming that the solid motion is purely tangential, the drag coefficient β can be calculated from [20]

$$\beta = 1.75 \frac{\varepsilon_s \cdot \rho_g}{d_p} |\vec{u}_{ig} - \vec{u}_{is}| = 1.75 \frac{\varepsilon_s \cdot \rho_g}{d_p} \sqrt{(u_{ig,t} - u_{is,t})^2 + u_{ig,r}^2} \quad (2.17)$$

Combining equations (2.15) and (2.17), setting down $u_{slip,t} = (u_{ig,t} - u_{is,t})$, leads to

$$F_{d,t} = \int_0^L \int_{R_f}^{D/2} \int_0^{2\pi} 1.75 \frac{\varepsilon_s \cdot \rho_g}{d_p} u_{slip,t} \sqrt{u_{slip,t}^2 + u_{ig,r}^2} d\phi r dr dl \quad (2.18)$$

Assuming that L is sufficiently small to ensure the axial uniformity and that the bed thickness h is sufficiently small ($h = D/2 - R_f \ll D/2$) to assume that $u_{ig,r}$ is constant across the particle bed, equation (2.18) can be simplified to

$$\begin{aligned} F_{d,t} &= L \frac{\left(\left(\frac{D}{2}\right)^2 - R_f^2\right)}{2} \int_0^{2\pi} 1.75 \frac{\varepsilon_s \cdot \rho_g}{d_p} u_{slip,t} \sqrt{u_{slip,t}^2 + u_{ig,r}^2} d\phi \\ &\simeq L h \frac{D}{2} \int_0^{2\pi} 1.75 \frac{\varepsilon_s \cdot \rho_g}{d_p} u_{slip,t} \sqrt{u_{slip,t}^2 + u_{ig,r}^2} d\phi \end{aligned} \quad (2.19)$$

Given the periodical domain, the integral in the tangential direction can be simplified to n times the integral between one gas inlet slot $\phi = 0$

and the angular position just before the next inlet slot $\phi = \varphi = \frac{2\pi}{n}$. Moreover, it is assumed that the solids volume fraction is tangentially uniform.

$$F_{d,t} = nLh \frac{D}{2} \frac{1.75\varepsilon_s \cdot \rho_g}{d_p} \int_0^\varphi u_{slip,t} \sqrt{u_{slip,t}^2 + u_{ig,r}^2} d\phi \quad (2.20)$$

Most of the transfer of tangential momentum to the particle bed occurs in the immediate vicinity of the gas inlet slot where the gas has a tangential velocity significantly higher than the particle bed tangential velocity. Setting down φ' such as most of the tangential momentum transfer occurs in the tangential location $0 \leq \phi \leq \varphi'$ and noticing that $u_{ig,r}$ is negligible compared to $u_{slip,t}$ in this zone because the gas has not been radially deflected yet, equation (2.20) becomes

$$F_{d,t} = nLh \frac{D}{2} \frac{1.75\varepsilon_s \cdot \rho_g}{d_p} \int_0^{\varphi'} u_{slip,t}^2 d\phi \quad (2.21)$$

In order to obtain the solution of equation (2.21), the tangential profile of $u_{slip,t}^2$ has to be determined. Assuming that $u_{ig,t}$ linearly decreases from $u_{ig,inj}$ at $\phi = 0$ to 0 at $\phi = \varphi$:

$$F_{d,t} = nLh \frac{D}{2} \frac{1.75\varepsilon_s \cdot \rho_g}{d_p} \int_0^{\varphi'} \left[u_{ig,inj} \left(1 - \frac{\phi}{\varphi} \right) - u_{is,t} \right]^2 d\phi \quad (2.22)$$

The resolution of equation (2.22) leads to

$$\begin{aligned} F_{d,t} &= \frac{\pi LhD}{\varphi} \frac{1.75\varepsilon_s \cdot \rho_g}{d_p} \varphi' \cdot \\ &\quad \left[u_{ig,inj}^2 \left(1 - \frac{\varphi'}{\varphi} + \frac{1}{3} \left(\frac{\varphi'}{\varphi} \right)^2 \right) - u_{ig,inj} u_{is,t} \left(2 - \frac{\varphi'}{\varphi} \right) + u_{is,t}^2 \right] \\ &= \pi LhD \frac{1.75\varepsilon_s \cdot \rho_g}{d_p} \frac{\varphi'}{\varphi} \left[A \left(\frac{\varphi'}{\varphi} \right) u_{ig,inj}^2 - B \left(\frac{\varphi'}{\varphi} \right) u_{ig,inj} u_{is,t} + u_{is,t}^2 \right] \end{aligned} \quad (2.23)$$

As $u_{is,t} \ll u_{ig,inj}$, the term in $u_{is,t}^2$ can be neglected. Equalizing equations (2.23) and (2.16) leads to a proportional relationship between the particle bed tangential velocity and the gas injection velocity:

2.5. Vortex chamber design

$$\begin{aligned}
f_{t,m}\rho_g n s L u_{ig,inj}^2 &= \pi L h D \frac{1.75 \varepsilon_s \cdot \rho_g}{d_p} \frac{\varphi'}{\varphi} [A u_{ig,inj}^2 - B u_{ig,inj} u_{is,t}] \\
&\Leftrightarrow f_{t,m} n s u_{ig,inj} = \pi h D \frac{1.75 \varepsilon_s}{d_p} \frac{\varphi'}{\varphi} [u_{ig,inj} - B u_{is,t}] \quad (2.24) \\
&\Leftrightarrow u_{is,t} = \frac{1}{B \left(\frac{\varphi'}{\varphi} \right)} \left[A \left(\frac{\varphi'}{\varphi} \right) - \frac{\varphi}{\varphi'} \frac{n s}{\pi h D} \frac{f_{t,m} d_p}{1.75 \varepsilon_s} \right] u_{ig,inj}
\end{aligned}$$

Consequently, the centrifugal force is given by

$$\begin{aligned}
F_c &= \int_{V_{bed}} \varepsilon_s \rho_s \frac{u_{is,t}^2}{r} dV = \int_0^L \int_{R_f}^{D/2} \int_0^{2\pi} \varepsilon_s \rho_s u_{is,t}^2 d\phi dr dl \\
&= 2\pi L h \varepsilon_s \rho_s \left(\frac{1}{B \left(\frac{\varphi'}{\varphi} \right)} \left[A \left(\frac{\varphi'}{\varphi} \right) - \frac{\varphi}{\varphi'} \frac{n s}{\pi h D} \frac{f_{t,m} d_p}{1.75 \varepsilon_s} \right] u_{ig,inj} \right)^2 \quad (2.25)
\end{aligned}$$

The radial gas-solid drag force over the particle bed can be calculated from

$$F_{d,r} = \int_{V_{bed}} f_{d,r} dV = \int_{V_{bed}} \beta (u_{ig,r} - u_{is,r}) dV \quad (2.26)$$

Introducing expression (2.17) in equation (2.26) leads to

$$\begin{aligned}
F_{d,r} &= \int_0^L \int_{R_f}^{D/2} \int_0^{2\pi} 1.75 \frac{\varepsilon_s \cdot \rho_g}{d_p} u_{ig,r} \sqrt{u_{slip,t}^2 + u_{ig,r}^2} d\phi r dr dl \\
&= 1.75 n L h \frac{D}{2} \frac{\varepsilon_s \cdot \rho_g}{d_p} \int_0^\varphi u_{ig,r} \sqrt{u_{slip,t}^2 + u_{ig,r}^2} d\phi \quad (2.27)
\end{aligned}$$

The radial drag force is significant mostly far from the gas inlet slot where the gas motion is essentially radial. In this zone ($\varphi'' \leq \phi \leq \varphi$), $u_{slip,t}$ can be neglected as compared to $u_{ig,r}$ leading to

$$F_{d,r} = 1.75 n L h \frac{D}{2} \frac{\varepsilon_s \cdot \rho_g}{d_p} \int_{\varphi''}^\varphi u_{ig,r}^2 d\phi \quad (2.28)$$

As $u_{ig,r}$ in the zone $0 \leq \phi \leq \varphi''$ can be neglected as compared to $u_{ig,r}$ in the zone $\varphi'' \leq \phi \leq \varphi$,

$$\int_{\varphi''}^{\varphi} u_{ig,r}^2 d\phi \simeq \int_0^{\varphi} u_{ig,r}^2 d\phi = \frac{2\pi}{n} \overline{u_{ig,r}^2} \quad (2.29)$$

Finally, the drag force can be expressed as

$$F_{d,r} = 1.75\pi LhD \frac{\varepsilon_s \cdot \rho_g \overline{u_{ig,r}^2}}{d_p} \quad (2.30)$$

To avoid particles entrainment, the mean centrifugal force ((2.25)) has to be higher than the mean radial gas-solid drag force ((2.30)), leading to the following criterion:

$$\begin{aligned} & 2\pi Lh\varepsilon_s\rho_s \left(\frac{1}{B\left(\frac{\varphi'}{\varphi}\right)} \left[A\left(\frac{\varphi'}{\varphi}\right) - \frac{\varphi}{\varphi'} \frac{ns}{\pi hD} \frac{f_{t,m}d_p}{1.75\varepsilon_s} \right] \right)^2 u_{ig,inj}^2 \\ & \geq 1.75\pi LhD \frac{\varepsilon_s \cdot \rho_g \overline{u_{ig,r}^2}}{d_p} \\ \Leftrightarrow & 2\rho_s \left(\frac{1}{B\left(\frac{\varphi'}{\varphi}\right)} \left[A\left(\frac{\varphi'}{\varphi}\right) - \frac{\varphi}{\varphi'} \frac{ns}{\pi hD} \frac{f_{t,m}d_p}{1.75\varepsilon_s} \right] \right)^2 u_{ig,inj}^2 \geq 1.75D \frac{\rho_g \overline{u_{ig,r}^2}}{d_p} \end{aligned} \quad (2.31)$$

Given that $u_{ig,inj} = \frac{Q_g}{nsL}$ and $\overline{u_{ig,r}^2} \simeq \frac{Q_g}{\varepsilon_g \pi DL}$ if h is sufficiently small, equation (2.31) becomes

$$\begin{aligned}
& 2\rho_s \left(\frac{1}{B\left(\frac{\varphi'}{\varphi}\right)} \left[A\left(\frac{\varphi'}{\varphi}\right) - \frac{\varphi}{\varphi'} \frac{ns}{\pi h D} \frac{f_{t,m} d_p}{1.75 \varepsilon_s} \right] \right)^2 \left(\frac{Q_g}{nsL} \right)^2 \\
& \geq 1.75 D \frac{\rho_g}{d_p} \left(\frac{Q_g}{\varepsilon_g \pi D L} \right)^2 \\
& \Leftrightarrow \pi \varepsilon_g \sqrt{\frac{2}{1.75} \frac{\rho_s}{\rho_g}} D d_p \left(\frac{1}{B\left(\frac{\varphi'}{\varphi}\right)} \left[A\left(\frac{\varphi'}{\varphi}\right) - \frac{\varphi}{\varphi'} \frac{ns}{\pi h D} \frac{f_{t,m} d_p}{1.75 \varepsilon_s} \right] \right) \geq (ns) \\
& \Leftrightarrow \varepsilon_g \sqrt{\frac{2}{1.75} \frac{\rho_s}{\rho_g} \frac{d_p}{D}} \left(\frac{1}{B\left(\frac{\varphi'}{\varphi}\right)} \left[A\left(\frac{\varphi'}{\varphi}\right) - \frac{\varphi}{\varphi'} \lambda \frac{f_{t,m} d_p}{1.75 h \varepsilon_s} \right] \right) \geq \lambda \\
& \Leftrightarrow A\left(\frac{\varphi'}{\varphi}\right) \left[\frac{B\left(\frac{\varphi'}{\varphi}\right)}{\varepsilon_g \sqrt{\frac{2}{1.75} \frac{\rho_s}{\rho_g} \frac{d_p}{D}}} + \frac{\varphi}{\varphi'} \frac{f_{t,m} d_p}{1.75 h \varepsilon_s} \right]^{-1} \geq \lambda
\end{aligned} \tag{2.32}$$

The maximum allowable slot size that satisfies this criterion versus the particles mean diameter is shown graphically in Figures 2.18a and 2.18b, respectively for different chamber diameters and for different solids densities. These graphs are valid for air as the gas phase and have been generated with realistic values of $\varepsilon_g = 0.5$, $h = D/6$, $\varphi' = 0.5\varphi$ and $f_{t,m} = 0.8$. As expected, fluidizing small and light particles requires a smaller slot size.

In summary, for a given gas flow rate and related gas-solid drag force, the amount of tangential momentum injected with the gas and the related centrifugal force can be increased by increasing $u_{ig,inj}$, that is, by decreasing $n \cdot s$ or λ . Reducing the slot size, however, results in an increased pressure drop over the gas inlet slots and related higher power requirements for the compressor.

It should be reminded here that equation (2.32) has been developed assuming high-G operation, i.e. that the gravity is negligible compared to the other forces involved. Equation (2.33) allows determining the minimum gas flow rate that ensure high-G.

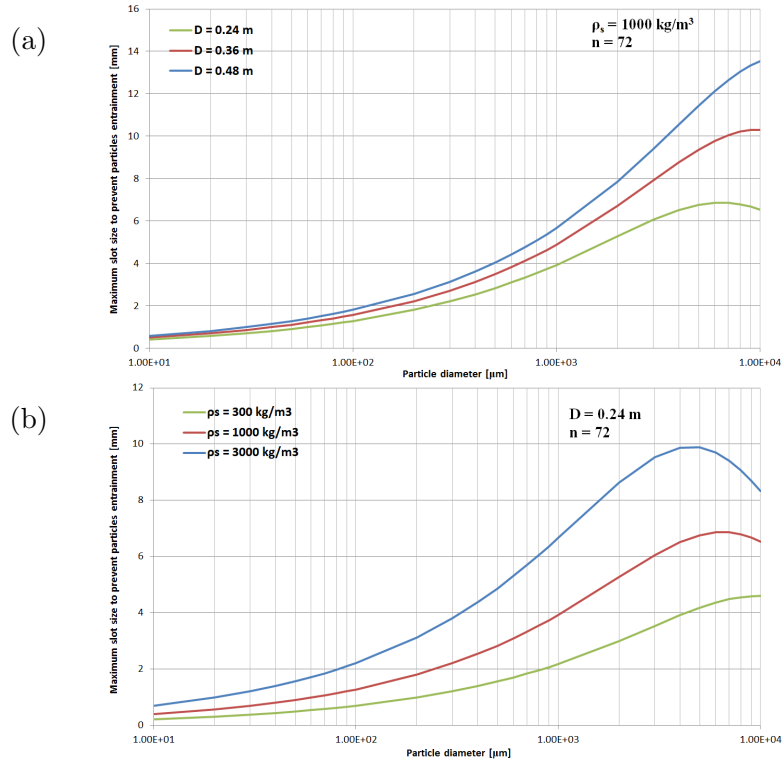


Figure 2.18: Maximum allowable slot size that satisfy equation (2.32) versus the particles mean diameter for (a) different chamber diameters and (b) different solids densities. Gas phase considered is air. Other numerical values: $\varepsilon_g = 0.5$, $h = D/6$, $\varphi' = 0.5\varphi$, $f_{t,m} = 0.8$ and $n = 72$.

$$\begin{aligned}
a_c &= \frac{u_{is,t}^2}{D/2} \gg g \\
\Leftrightarrow \left(\frac{1}{B\left(\frac{\varphi'}{\varphi}\right)} \left[A\left(\frac{\varphi'}{\varphi}\right) - \frac{\varphi}{\varphi'} \frac{ns}{\pi h D} \frac{f_{t,m} d_p}{1.75 \varepsilon_s} \right] \right)^2 \left(\frac{Q_g}{nsL} \right)^2 \frac{2}{D} &\gg g \quad (2.33) \\
\Leftrightarrow Q_g &\gg nsL \sqrt{\frac{gD}{2}} \left(\frac{1}{B\left(\frac{\varphi'}{\varphi}\right)} \left[A\left(\frac{\varphi'}{\varphi}\right) - \frac{\varphi}{\varphi'} \frac{ns}{\pi h D} \frac{f_{t,m} d_p}{1.75 \varepsilon_s} \right] \right)^{-1}
\end{aligned}$$

To validate their criterion, Rosales and De Wilde [76] carried out fluidization experiments with 70 μm FCC catalyst particles varying λ for given solids and gas flow rates. As seen from Figure 2.19, reducing λ allows reducing particles losses through the chimney and increasing the solids loading in the vortex chamber. Increasing the centrifugal force while keeping constant the gas-solid drag force also leads to an improved particle bed uniformity, i.e. by limiting bubbles formation. Ekatpure et al. [79] reported similar observations carrying out experiments with different types of particles in a vortex chamber with $\lambda = 0.127$ ($n = 36$, $s = 6$ mm). They obtained a stable rotating fluidized bed with relatively large polymer particles ($d_p = 0.9, 1.6$ and 2.4 mm, $\rho_s = 950$ kg/m³) but not with small FCC catalyst particles ($d_p = 70$ μm , $\rho_s = 1500$ kg/m³). The fluidization of FCC catalyst required smaller slot size of 2 mm.

As previously mentioned, increasing the pressure drop over the gas inlet slots can lead to an improved axial uniformity in the particle bed or, in the extreme case, avoid channelling. Axial non-uniformities can be measured through differences in the pressure or the rotational speed. Using Laser Doppler Velocimetry (LDV) and a rotating antenna, Dvornikov and Belousov [80] found out that the rotational speed of the particle bed was significantly lower near the end walls than in the axial center while drying relatively large particles ($d_p = 3.5$ mm, $\rho_s = 1420$ kg/m³) in a vortex chamber with $\lambda = 0.1$ (Figure 2.20b). The same observation was made by Rosales and De Wilde [76] while fluidizing FCC catalyst particles ($d_p = 70 - 85$ μm , $\rho_s = 1500$ kg/m³) in a vortex chamber with $\lambda = 0.048$ and 0.024 (Figure 2.20c). These authors also reported that the axial gradient in the particle bed rotational speed was seen to increase with the gas flow rate. Figure 2.20c also shows the impact of

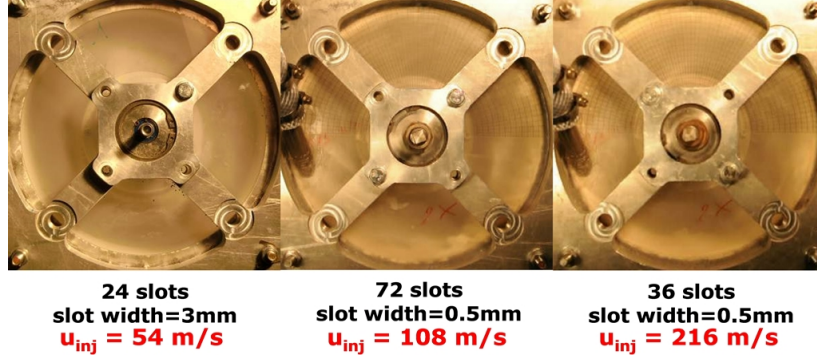


Figure 2.19: Pictures of the FCC catalyst ($d_p = 70 - 85 \text{ }\mu\text{m}$, $\rho_s = 1500 \text{ kg/m}^3$) particle beds obtained while continuously feeding solids for given solids (8.35 g/s) and gas flow rates ($700 \text{ Nm}^3/\text{h}$) in vortex chambers with different λ . Vortex chamber: $D = 0.24 \text{ m}$, $D_c = 0.10$, $L = 0.05$ [76].

an increased pressure drop over the gas inlet slots on the axial uniformity. Indeed, decreasing λ from 0.048 to 0.024 and as such increasing the pressure drop over the slots, reduces the axial gradient in the particle bed rotational speed and, as a result, improves the particle bed axial uniformity. Rosales and De Wilde [76] developed a theoretical criterion to ensure the axial uniformity in the particle bed. This criterion is based on the fact that the pressure drop over one slot has to be higher than the pressure drop over the particle bed to allow the gas repartition not being dependent on what occurs in the vortex chamber.

$$\Delta p_{slot} > \Delta p_{bed} \quad (2.34)$$

As for the tangential uniformity, the required λ to ensure axial uniformity was theoretically shown to decrease with decreasing particles size and density [76]. Depending on the vortex chamber design and operating conditions, criterion (2.32) or (2.34) can be more restrictive to design the size of the gas inlet slots.

2.5.2 Gas outlet design

The gas outlet, i.e. chimney, design influences the particles losses rate through the chimney. Solids losses occur when particles start to move radially inward, i.e. when the centrifugal force is not sufficiently high to compensate for the radial gas-solid drag force. As previously men-

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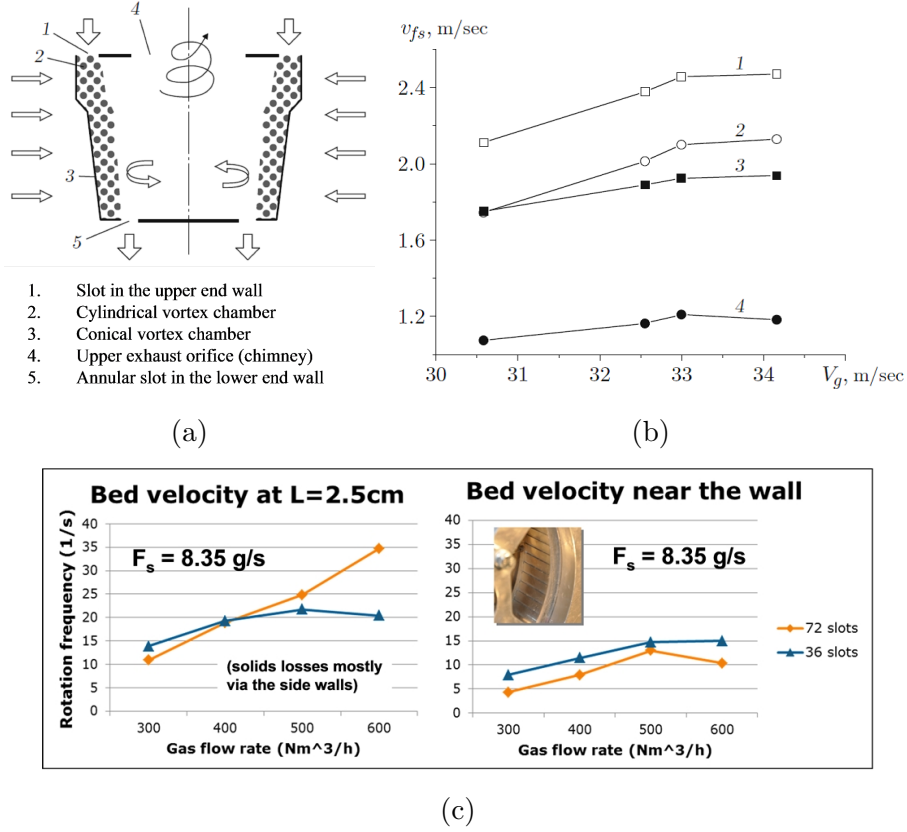


Figure 2.20: (a) Scheme of a conical vortex chamber with upper cylindrical part and lower conical part. (b) Average tangential particle bed velocity versus the mean gas inlet velocity (1) at the axial center and (3) near the end walls of the upper cylindrical part of the vortex chamber. (2) and (4): the same in the lower conical part of the vortex chamber. Vortex chamber: $0.200 \text{ m} \leq D \leq 0.260 \text{ m}$ (conical), $L = 0.127 \text{ m}$ [80]. (c) Particle bed rotational frequency versus gas flow rate using two different gas inlet designs (36 and 72 x 0.5 mm gas inlet slots) while fluidizing 70-85 μm FCC catalyst particles in a vortex chamber: comparison between the particle bed rotational frequency at the axial center (left) and near the end walls (right) of the vortex chamber. Vortex chamber: $D = 0.24 \text{ m}$, $D_c = 0.10$, $L = 0.05 \text{ m}$. Solids feeding rate of 8.35 g/s [76].

tioned, this lack of centrifugal force is likely to happen near the end walls, the particles rotating slower because of friction. As a result, a significant boundary layer secondary flow can be generated [66]. Loftus et al. [63] took advantage of this phenomenon in gas-liquid vortex chambers to remove the heavy phase (liquid) using a secondary concentric chimney (Figure 2.21). Different chimney designs have been proposed to minimize the solids losses via the chimney and to improve the bed stability and density at given operating conditions. Two chimney types can be distinguished. Figures 2.4 and 2.21 show a cyclone flow exit tube which consists of a cylinder with a guide slot that is completely inserted in the vortex chamber. Oppositely, an open ended tube, or "vortex finder", is used in Figures 2.14, 2.16, 2.17 and 2.19. The vortex finder can be inserted in the vortex chamber over a short length or used in combination with concentric rings placed on the end walls to reduce particles losses through the chimney [57]. Better results in terms of particle bed stability and solids losses through the chimney are obtained with the vortex finder design [63, 76]. Indeed, the vortex finder allows to approach a free vortex type flow pattern in the freeboard region (see Figure 2.6a) and, as a result, intensifies the Coriolis effect which returns the particles entrained towards the chimney back to the particle bed. An intensified Coriolis effect consequently reduces particles losses through the chimney. Decreasing the chimney diameter, and as such increasing the size of the freeboard region (see Figure 2.22 [76]), also reduces the solids losses through the chimney because the time during which the particles are subjected to the Coriolis effect is increased. The positive effect of a reduced chimney diameter on the solids retention and particle bed stability was experimentally confirmed by Loftus et al. [63]. The bed stability is improved, however, at the cost of an increased pressure drop over the freeboard and the chimney. A symmetrical double chimney (Figure 2.21), i.e. each on one end wall, allows reducing the pressure drop over the chimney. Carrying experiments with relatively large 1G-Geldart B and D type particles in a vortex chamber with a single chimney, Kochetov et al. come to the conclusion that the optimum ratio between the chimney and vortex chamber diameters was ranging between 0.3 and 0.5. Smaller chimneys are, however, expected to be better suited for smaller/lighter particles.

Working with small particles, Anderson et al. [59] proposed a secondary gas injection to suppress the boundary layer flow inducing the

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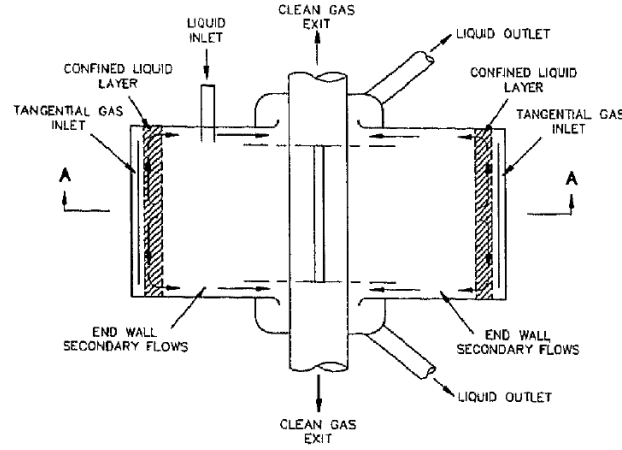


Figure 2.21: Concentric dual-chimney design for the removal of the heavy phase (liquid) with a cyclone flow exit tube with a guide slot for the gas phase [63].

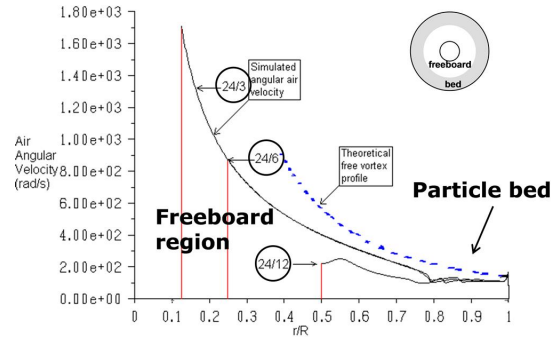


Figure 2.22: CFD simulations of the fluidization of 60 μm FCC catalyst particles in a 0.24 m vortex chamber. Radial profiles of the air angular velocity for three different chimney diameters of 0.03, 0.06 and 0.12 m at given gas flow rate and solids loading [76].

major part of solids losses. As shown in Figure 2.23a, it consists of additional nozzles located centrally at the end of a tapered annex and concentrically with the gas outlet. The impact of the secondary injection gas flow rate $\dot{m}_{EW}$ near the end walls on the solids losses is shown in Figure 2.23b. The rate of solids losses through the chimney is quantified by means of a rate coefficient, k , based on the experimental observation that the rate of solids losses is proportional to the solids loading M in the vortex chamber. When solids are not fed to the vortex chamber, that leads to

$$-\frac{dM}{dt} = k \cdot M \quad (2.35)$$

, that is, an exponential decrease of the solids loading $M = M_0 \exp(-kt)$, M_0 being the solids loading at time 0. The decrease of the coefficient k with the secondary injection gas flow rate $\dot{m}_{EW}$ near the end walls clearly demonstrates the positive effect of the latter on reducing particles losses through the chimney.

Another adaptation of the chimney design consists in profiling the end wall(s) of the vortex chamber connected to the chimney(s) [64, 68]. When the particle bed is sufficiently dense, the rotational motion of the particles approaches the solid body type flow pattern ($u_{ig,t}$ is proportional to r). As a result, with flat end walls, the ratio between the gas-solid drag and centrifugal forces grows when going from the cylindrical outer wall to the chimney. Profiling the end wall connected to the chimney in such a way to compensate for the latter allows balancing the two forces for each radial position. Kuzmin et al. [64] theoretically calculated the end wall profile required to balance the gas-solid drag and centrifugal forces in a vortex chamber with a single chimney. They also added a concentric ring to further reduce solids losses via the chimney (Figure 2.24). Profiling the end wall, however, induces the destruction of the free vortex in the particle bed freeboard region and the related positive effect on maximizing the Coriolis effect that allows reducing significantly the solids losses through the chimney [76].

Finally, a rotating chimney can be used in vortex chambers to reduce the particles losses through the chimney [60, 64]. It consists in a centrally positioned chimney rotating in the same sense as the particle bed, but at higher rotational speeds. As a result, the centrifugal force in the vicinity of the chimney is increased, leading to a reduction of

2.5. Vortex chamber design

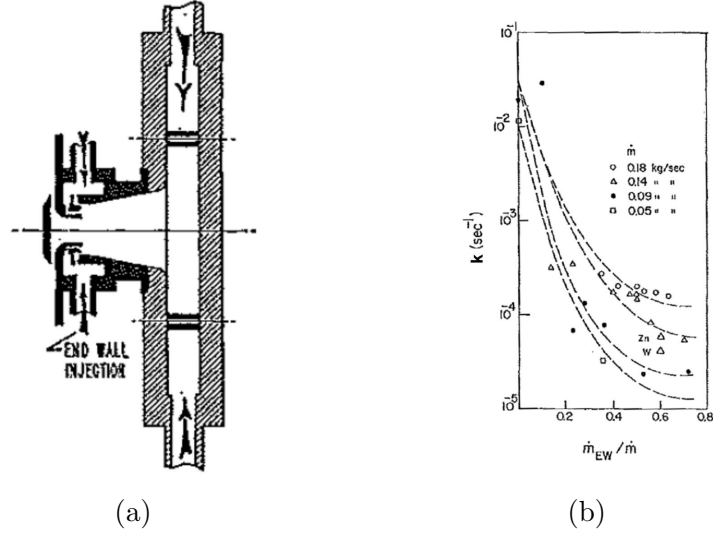


Figure 2.23: Vortex chamber with a vortex finder type chimney and secondary gas injection nozzles located centrally at the end of a tapered annex and concentrically with the gas outlet. Vortex chamber: $D = 0.305$ m, $D_c = 0.15 \rightarrow 0.08$ m (convergent), $L = 0.063$ m, $n = 12$, $s = 0.3$ mm, $\lambda = 0.00376$. (a) Schematic representation. (b) Rate coefficient k for particles losses through the chimney versus the normalized secondary gas injection flow rate for different total gas flow rates. $\dot{m}_{EW}$: secondary injection gas flow rate near the end walls, $\dot{m}$: total gas flow rate [59].

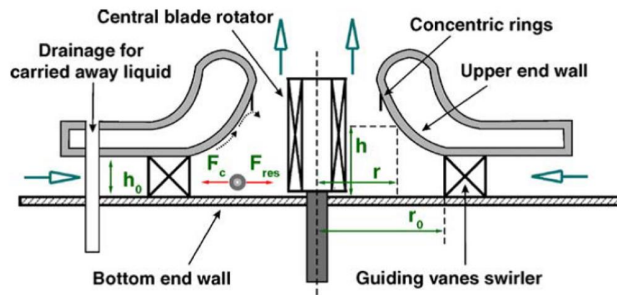


Figure 2.24: Profiled end wall chimney with concentric ring [64].

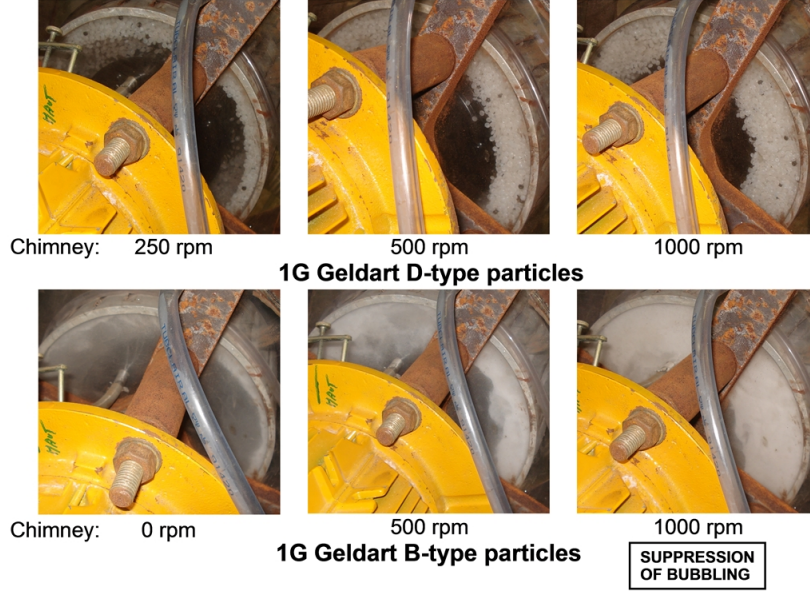


Figure 2.25: Influence of the rotational speed of a rotating chimney on the solids loading and particle bed density and uniformity in a vortex chamber with continuous solids feeding and operated at a gas flow rate of $850 \text{ Nm}^3/\text{h}$. Top: HDPE polymer particles ($\rho_s = 950 \text{ kg/m}^3$, $d_p = 3.5 \text{ mm}$); Bottom: salt particles ($\rho_s = 2200 \text{ kg/m}^3$, $d_p = 350 \text{ }\mu\text{m}$). Vortex chamber: $D = 0.24 \text{ m}$, $D_c = 0.16 \text{ m}$, $L = 0.115 \text{ m}$, $n = 24$, $s = 2.3 \text{ mm}$, $\lambda = 0.073$ [67].

the solids losses through the chimney for a given solids loading and gas flow rate. Experimental studies confirmed this theoretical behaviour and demonstrated that a rotating chimney can be used to control the solids residence time in the vortex chamber [61,67]. Furthermore, the rotating chimney was seen to suppress bubbling and to increase the solids loading while fluidizing 1G-Geldart B particles for the given vortex chamber and operating conditions (Figure 2.25). This effect of particle bed compression allowed extending the range of particle bed rotational speeds for which a uniform and dense particle bed is obtained [67].

2.5.3 Solids inlet design

Solids can be injected via the end walls (side solids inlet) or through the outer cylindrical wall of the vortex chamber. The side solids inlet

2.5. Vortex chamber design

consists typically of a tube pierced through one of the end walls, but a peripheral slot [80] can be used instead. The side solids inlet is the more simple way to feed solids in the vortex chamber. Injection of solids through the outer cylindrical wall can be performed using the gas inlet slots or separately from the gas via additional slots. In both cases, the size of the slots used to inject the particles must be sufficiently large to avoid clogging. However, the gas inlet slots cannot be too large to ensure a stable and dense particle bed (see section 2.5.1) so that feeding solids through the gas inlet slots is not evident. It should also be noticed that for a normal usage of vortex chamber and to facilitate uniform operation, the particle bed rotational motion should be driven mainly by the gas phase and not result from the solids injection. Rosales and De Wilde highlighted by CFD simulations the disturbing effect of using separate solids inlet slots in the outer wall of the fluidization chamber on the particle bed hydrodynamics. Experimental and numerical studies showed more positive results while feeding solids through the gas inlet slots.

2.5.4 Solids outlet design

The solids outlet design is crucial when operating with continuous solids feeding and removal. It can consist of one tube pierced through one of the end walls of the vortex. De Wilde and de Broqueville [60] demonstrated the proper functioning of such solids outlet while continuously feeding and removing 1G-Geldart D type 3 mm diameter polymer particles to and from a rotating fluidized bed in a vortex chamber. With the vortex chamber design and particles characteristics used in those experiments, the particle bed freeboard contained no particles and no solids losses were observed via the chimney. Furthermore, the particle bed height and related solids loading could be controlled by the radial position of the solids outlet. Solids removal via one of the end walls was also performed for relatively large particles ($d_p = 3.5$ mm, $\rho_s = 1420$ kg/m³) by Dvornikov and Belousov [80] using an annular slot (Figure 2.20a). When working with smaller and/or lighter particles, alternative ways of removing the particles may have to be considered, either via the chimney with the gas or either through the outer cylindrical wall of the vortex chamber. As seen in Figure 2.26a, Volchkov et al. [68] used a cyclone type system located on the end wall facing the chimney. The authors reported a proper functioning of this solids outlet design while working with relatively large particles ($d_p = 2$ -8 mm) but its usage with smaller

and/or lighter particles has still to be studied. Goldshtik et al. [66] proposed different configurations for feeding and removing the solids via the slots in a liquid-solid vortex chamber designed for the development of a vortex liquid piston engine (Figure 2.20b). The evacuation of the solids by entrainment with the gas through the chimney results in a reduction of degree of freedom because the solids loading and the related solids residence time cannot be controlled independently from the gas flow rate. Furthermore, removing the solids through the chimney forces the particles to enter the particle bed freeboard region. If the solids concentration in the freeboard region becomes too high, this can lead to the destruction of the free vortex flow pattern in this region, which is, as previously mentioned, detrimental for the particle bed stability and density. The latter also results in a reduced solids loading in the vortex chamber for a given solids feeding rate. In gas-liquid vortex chambers, Loftus et al. [80] took advantage of the boundary flows near the end walls to remove the heavy phase (liquid) through a specially designed secondary chimney to be used in combination with a vortex finder (Figure 2.21).

2.5.5 Other design aspects

Cylindrical, polygonal, or complexly shaped vortex chambers have been designed. De Wilde [69] experimentally and theoretically showed that polygonal vortex chambers offer a more efficient transfer of tangential momentum from the injected gas to the particle bed. Indeed, a cylindrical design leads to friction with the wall in the immediate vicinity of the gas inlet slots which hinders the tangential acceleration of the particles. A polygonal design also facilitates the construction of the vortex chamber and few data tend to demonstrate a reduced attrition in the particle bed. Hybrid vortex chambers with an upper cylindrical part and a lower conical part have also been used for grain drying (Figure 2.20a) [68, 80].

The effect of the length of the vortex chamber on the axial non-uniformities was already discussed in section 2.4.4. The latter can be reduced by increasing the pressure drop over the gas inlet slots. However, Kochetov et al. [57] stated that the length of the vortex chamber should be limited to the half of the vortex chamber diameter in order to ensure the axial uniformity of the bed. This rule was confirmed by more recent experimental data [59, 60, 62, 68, 69, 76].

2.5. Vortex chamber design

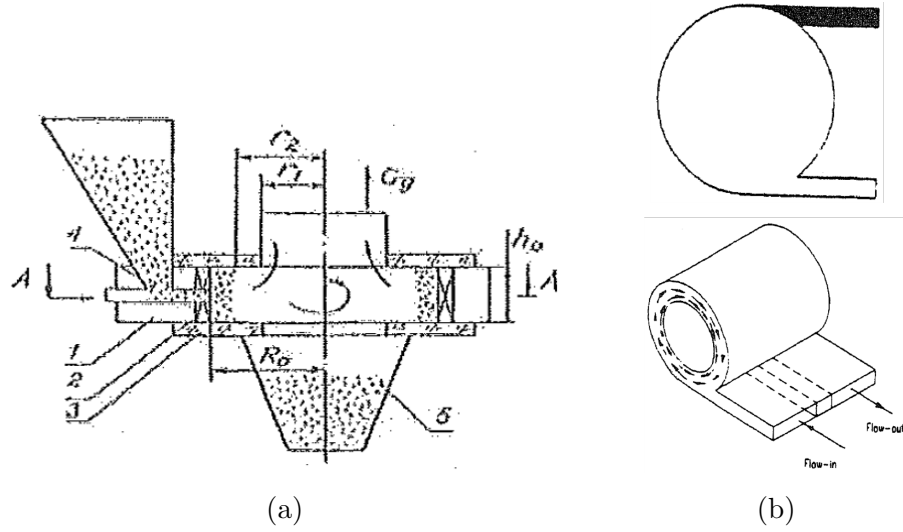


Figure 2.26: (a) Vortex chamber with a cyclone type system for solids removal. (1) Gas distribution chamber (helix); (2) vortex chamber; (3) end walls; (4) solids feeder; (5) collecting hopper [68]. (b) Different configurations for feeding and removing the solids via the slots in a liquid-solid vortex chamber designed for the development of a vortex liquid piston engine [66].

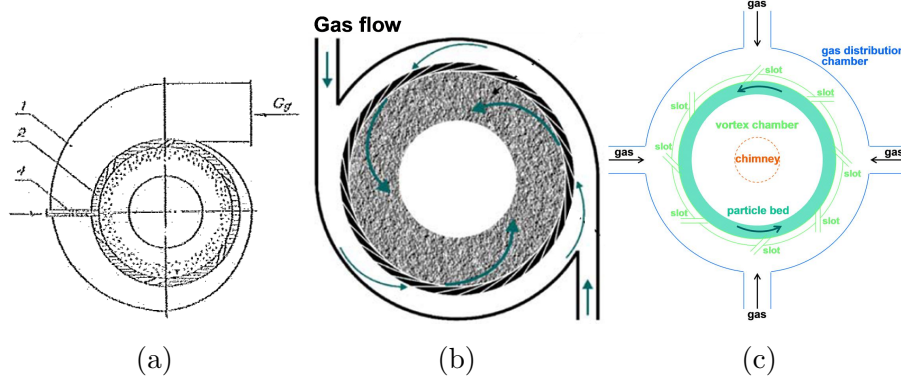


Figure 2.27: Different designs for the gas distribution chamber. (a) Helix type [68], (b) tangentially fed [64] and (c) radially fed distributors [69].

Finally, when operating above the atmospheric pressure, the process gas is generally fed in a gas distributor prior to its injection in the vortex chamber. Different designs have been proposed for the gas distribution chamber. Among them, Figure 2.27 shows the helix type, tangentially fed and radially fed distributors. Rosales and De Wilde [76] demonstrated the minor impact of the gas distributor design on the flow pattern in the vortex chamber if the pressure drop over the gas inlet slots is sufficiently high, i.e. if λ is sufficiently small.

2.6 Heat transfer in vortex chambers

As previously explained, high-G fluidization allows increasing gas-solid slip velocities and the related heat and mass transfer efficiency. De Wilde and de Broqueville [27] theoretically and numerically studied the potential of rotating fluidized beds in vortex chambers for the intensification of interfacial heat transfer. These authors carried out 2D CFD simulations of the response of the particle bed temperature to a step change in the fluidization gas temperature, both for a conventional fluidized bed and for a RFB-SG in a vortex chamber keeping constant the mass of particles in the fluidization chamber. The results, shown in Figure 2.28 demonstrate that higher specific gas flow rates, i.e. per m^3 particle bed and higher slip velocities led to a much faster heating of the particle bed in the RFB-SG than in the conventional fluidized bed. Moreover, the uniformity of the temperature in the particle bed was seen to be

2.7. Multizone operation

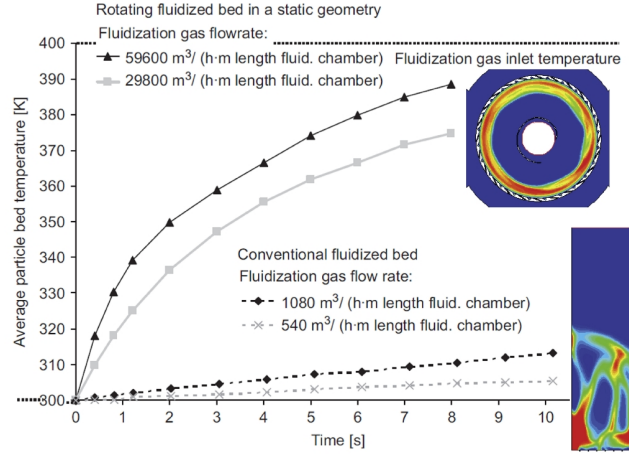


Figure 2.28: 2D CFD simulations of the response of the average particle bed temperature to a step change in the fluidization gas temperature from 300 to 400 K at $t = 0$ s. Comparison of conventional fluidized bed and RFB-SG technologies for the same mass of particles in the fluidization chamber of 33.75 kg/m_{length fluid. chamber}. Profiled end wall chimney with concentric ring. Embedded figures: snapshots of the solids volume fraction profiles [27].

improved in the RFB-SG because of the efficient mixing resulting from the particle bed rotational motion (Figure 2.29).

2.7 Multizone operation

As previously explained in section 2.4.2, the gas injected in a vortex chamber through one gas inlet slot is barely mixed with the gas injected through the next one, allowing tangential multizone operation. Vortex chambers can also be designed for an axial multi-zone operation. In both cases, different gases can be injected through a certain number of gas inlet slots or through a certain part of the slots in the axial direction, allowing integrating different reaction zones in one single vortex chamber. Multizone operation can also be done in terms of heat supply, that is, feeding very hot gas in a limited zone of the vortex chamber. The latter allows to limit the time of exposure of the particle bed to very high temperatures while supplying an important amount of heat into the vortex chamber. CFD simulations demonstrated that significant differ-

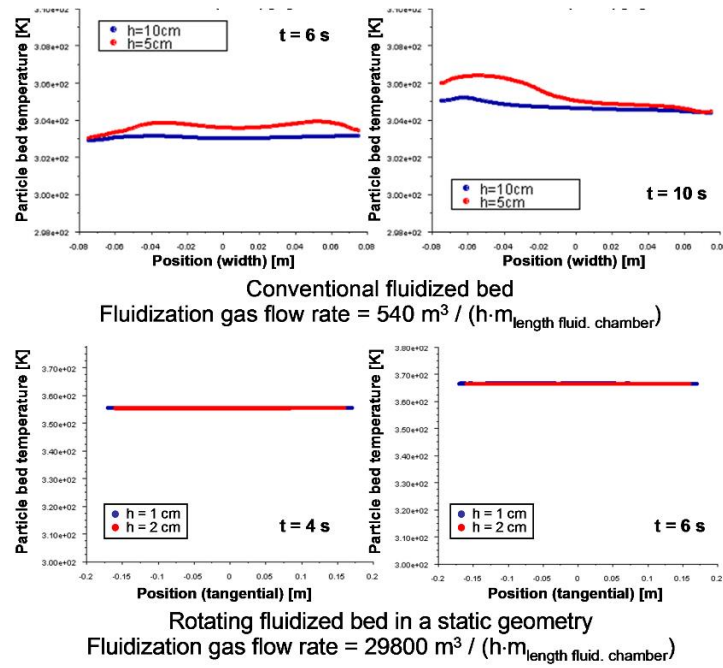


Figure 2.29: 2D CFD simulations of the response of the particle bed temperature to a step change in the fluidization gas temperature from 300 to 400 K at $t = 0$ s. Comparison of the bed uniformity in a conventional fluidized bed and in a RFB-SG [27].

2.7. Multizone operation

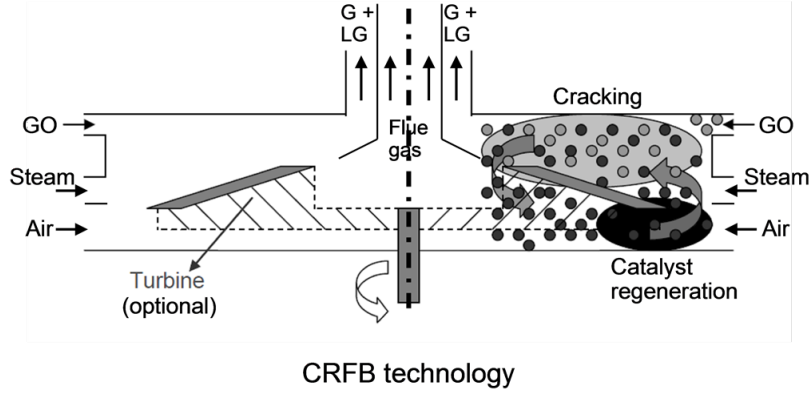


Figure 2.30: Schematic representation of a CRFB designed for catalytic cracking [69].

ences in the temperatures of the gases fed to the different zones resulted in very small temperature differences within the particle bed because of the efficient mixing in vortex chambers [78].

Multizone operation can be extended to the concept of Circulating Rotating Fluidized Beds (CRFBs), in which an axial-radial circulation of the particles is superimposed on the particle bed rotational motion. The axial-radial circulation, which is already present because of the axial gradient in the particle bed rotational speed, can be intensified and controlled by introducing the gas in an axially non-uniform way, as demonstrated by CFD simulations [78]. As seen from Figure 2.30, showing a schematic representation of a CRFB designed for catalytic cracking [69], the CRFBs can be used in combination with a secondary gas outlet. CRFBs were also considered for low-temperature pyrolysis and gasification of biomass [78]. In this case, the vortex chamber is separated between a zone where the biomass is pyrolysed and a zone where the char is combusted. The heat required for the pyrolysis is generated in the combustion zone and transferred by means of the axial circulation of the particles.

Nomenclature

Ar	-	Archimedes number
a_c	m/s^2	centrifugal acceleration
C_D	-	drag coefficient
D	m_r	vortex chamber diameter
D_c	m_r	chimney diameter
d_p	m	particles diameter
F_c	N	centrifugal force over the particle bed
F_d	N	gas-solid drag force over the particle bed
f_c	N/m_r^3	centrifugal force per cubic meter fluidization chamber
f_d	N/m_r^3	gas-solid drag force per cubic meter fluidization chamber
f_{tm}	-	fraction of the gas tangential momentum transferred to the particle bed
g	m/s^2	earth's standard acceleration due to gravity
h	m_r	bed thickness (radial direction)
L	m_r	vortex chamber length
l	m_r	axial coordinate
M	kg	solids loading
$\dot{m}_g$	kg/s	gas phase mass flow rate
n	-	number of gas inlet slots
n_{rot}	-	number of rotation of the gas phase in the particle bed
$p_{inj,t}$	$\text{kg.m}_r/\text{s}$	tangential momentum injected with the gas
Δp_{slot}	Pa	pressure drop over the gas inlet slots
Δp_{bed}	Pa	pressure drop over the particle bed
Q_g	m_g^3/s	gas phase volumetric flow rate
R_f	m_r	radial position of the freeboard, i.e. the inner delimitation of the particle bed.
Re_p	-	particle-based Reynold number
r	m_r	radial coordinate
S_r	m^2	surface area available for the radial motion of the gas
S_t	m^2	surface area available for the tangential motion of the gas
s	m_r	size of the gas inlet slots
s_{gs}	-	tangential gas-solid slip factor

2.7. Multizone operation

t	s	time coordinate
u_{ik}	m_r/s	interstitial velocity of phase k
u_{sk}	$m_k^3/(m_r^2.s)$	superficial velocity of phase k
u_{slip}	m_r/s	gas solid slip velocity
u_{mf}	$m_g^3/(m_r^2.s)$	gas superficial velocity at the minimum fluidization point
u_t	m_r/s	particles terminal or free-fall velocity
V_{bed}	m_r^3	volume taken by the particle bed

Greek letters

β	$kg/(m_r^3.s)$	drag coefficient
Γ	m^2/s	circulation
ε_k	m_k^3/m_r^3	volume fraction of phase k
λ	-	fraction of the chamber circumference taken by gas inlet slots
μ_k	Pa.s	viscosity of phase k
ρ_k	kg/m_k^3	density of phase k
ϕ	rad	tangential coordinate
φ	rad	angle between 2 adjacent slots
φ'	rad	angle such as most of the tangential momentum transfer occurs within $0 \leq \phi \leq \varphi'$
φ''	rad	angle such as most of the radial momentum transfer occurs within $\varphi'' \leq \phi \leq \varphi$
ω	rad/s	rotational speed

Subscripts

b	particle bed
g	gas phase
inj	injection
mf	minimum fluidization
r	reactor (units)
r	radial (variables)
s	solid phase
t	tangential

Biomass drying: state of the art

3.1 Introduction

Biomass can be converted by thermochemical processes such as combustion, pyrolysis and gasification (Figure 3.1) [81]. Woody biomass is generally dried prior to conversion in order to improve their efficiency. Indeed, with wet woody fuel, part of the heat supplied to the boiler or gasifier is used to evaporate the moisture contained in the biomass. A reduced moisture level allows reaching higher flame temperatures and related increased steam production rates. Moreover, a higher flame temperature also results in a more complete combustion with reduced CO and ash levels and a more efficient heat recovery. Evidently, the use of a fuel drier prior to combustion also consumes energy, but the efficiency can be increased by the drier design. Furthermore, the consumed energy can be recovered from the boiler or gasifier flue gas. Dry biomass is also required for pelletization which often precedes combustion or gasification processes. Finally, the drying of biomass leads to a reduced biological activity and density, facilitating transport and storage [82–85].

For combustion the optimum moisture content of woody biomass is between 10 and 25 wt-% on a humid basis. For gasification it varies from 10 to 65 wt-% depending on the type of gasifier used while pyrolysis requires a biomass moisture content ranging between 5 and 10 wt-%. For pelletization, finally, a moisture content below 15 wt-% is needed to obtain stable pellets [81–84].

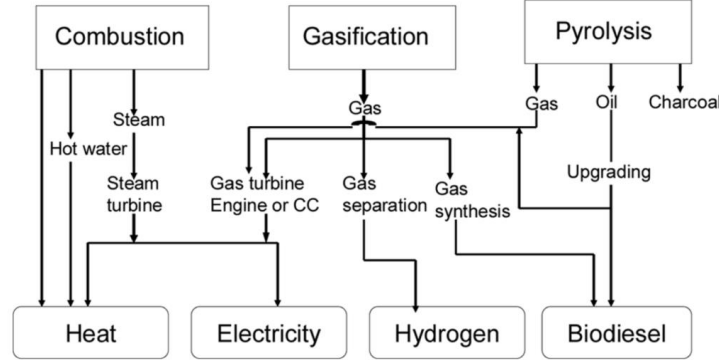


Figure 3.1: Thermochemical processes to convert biomass into energy products [81].

3.2 Granular biomass drying technologies

Conventional biomass dryers are categorized as direct or indirect. In direct dryers, the heat source is in contact with the biomass being dried. In indirect dryers, the heat is supplied through heat exchanging walls. The latter allow recovering the latent heat of vaporization because the water vapor is not diluted in the drying air [84]. On the other hand, indirect dryers are less efficient than direct dryers because of the heat losses resulting from the heat exchange [83].

Biomass dryers can operate at atmospheric pressure or under vacuum to reduce the water boiling point. The drying medium commonly used is air but the heat can also be supplied by a flue gas or superheated steam. In this latter case, the steam can be used in a closed-loop and a large part of the energy can be recovered from the excess steam [82–84]. Moreover, the use of superheated steam rather than air prevents the risk of fire.

Finally, it should be remarked that a cyclone is generally required downstream to the dryer to separate entrained particles from the drying medium.

3.2.1 Rotary drum dryers

The rotary drum dryer is widely used for biomass drying. It consists of a slowly rotating cylinder in which the biomass is fed continuously. As seen in Figure 3.2a, the drum has an inclination of 3 to 5 ° to ensure the axial motion of the biomass. The rotation of the drum combined with the presence of flights lift the particles in the dryer (Figure 3.3). Consequently, the particles cascade down through the drying air, leading to efficient gas-solids contacts and related mass and heat transfer. The flights also ensure a uniform end product as all the particles have the same probability of coming into contact with the drying air [81]. Rotary drums are typically operated at air inlet temperatures between 250 and 550 °C and rotation speeds between 2 and 8 rpm, leading to a capacity between 0.5 and 150 tons of biomass per hour depending on the drum diameter. Direct rotary drums dryers can be operated co- or counter-currently, the latter being more efficient. However, it should be avoided in case of fire risk or with small particles that could be entrained by the air flow [84,86]. Typically, the co-current operation is chosen for biomass drying [87]. Rotary drum drying is simple, robust, well-known, flexible and is suited for all particle sizes. A rotary drum is, however, a heavy equipment because of its moving parts. Moreover, the poor heat transfer efficiency requires a high air throughput and a high air inlet temperature, resulting in a pronounced emission of volatile organic compounds (VOCs) [87].

An indirect rotary drum dryers is shown in Figure 3.2b. Indirect heating is preferred when the granular material being dried cannot be exposed to air or combustion products. Usually, a heat exchanger with steam tubes is located inside the drum [83,87,89].

3.2.2 Perforated floor dryers

A perforated floor dryer (Figure 3.4) is essentially a fixed bed of biomass through which the air is flowing. It is commonly operated batch-wise and is suited for small plants [87]. The main drawback of fixed bed dryers is the product quality, with large vertical gradient in the moisture content. Indeed, the particles located at the bottom of the bed are in contact with fresh air during the whole batch time and, as a result, are more dried than those located on the top of the bed.

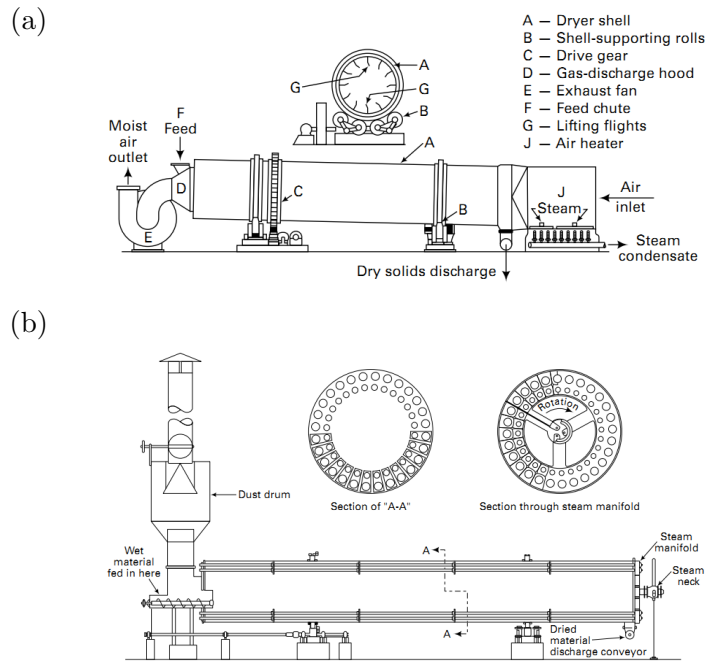


Figure 3.2: (a) Direct and (b) indirect rotary drum drying [88].

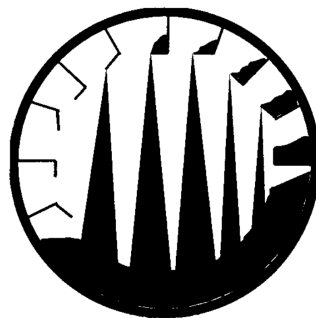


Figure 3.3: Flight arrangement in direct rotary drum dryers [89].

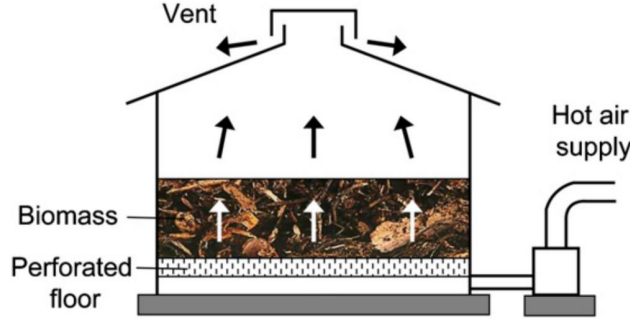


Figure 3.4: Perforated floor dryer [87].

3.2.3 Packed moving bed dryers

The packed moving bed dryer is typically compartmented (Figure 3.5). The particles are fed continuously on a conveyor belt, forming a fixed bed moving from one compartment of the dryer to another. To feed the air, the conveyor belt is perforated [81, 90]. Single-pass and multi-pass (Figure 3.6) conveyors exist. In the latter, the biomass is discharged at the end of a conveyor to the one below. Hot air is flowing upwards in the first half of the dryer and is then recycled by flowing downwards in the second half. The vertical gradient of moisture content created in the first half of the dryer is thereby reduced in the second half, leading to a relatively homogeneous end product. As in rotary drums, a counter-current configuration can be used to enhance the drying efficiency if there is no risk of fire. In this case, fresh air is fed upwards in the second half of the dryer and recycled to the first half. A typical air inlet temperature ranges between 80 and 120 ° C. This rather low drying temperature limits the risk of fire and the emissions of VOCs [81, 83, 90]. However, the air inlet temperature must be kept high enough to ensure that the exhaust air temperature is appropriate to achieve the required moisture level. The drying time can be easily controlled in packed moving bed dryers by setting the conveyor belt speed.

3.2.4 Fluidized bed dryers

Fluidized beds are also used to dry the biomass [87, 89, 91, 92]. Depending on the fluidization regime, fluidized bed dryers are categorized as fluid bed, spouted bed, pneumatic or flash dryers. Fluid bed dryers (Figure 3.7a), or spouted bed dryers/cascade dryers when fluidizing D-

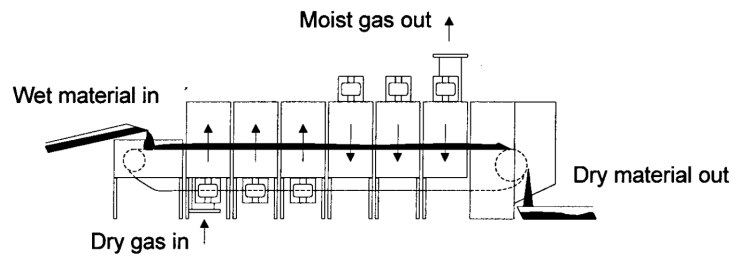


Figure 3.5: Packed moving bed dryer [89].

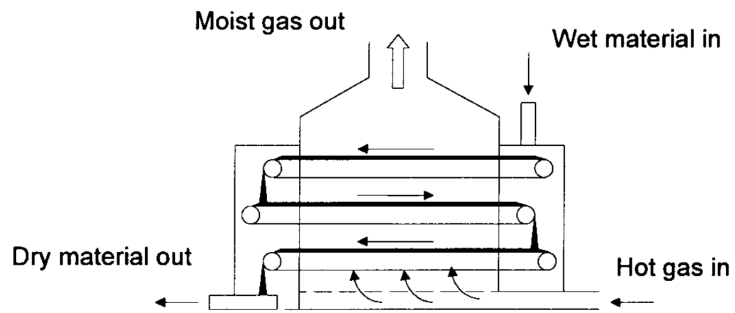


Figure 3.6: Multipass packed moving bed dryer [89].

3.2. Granular biomass drying technologies

type Geldart particles, make use of a fluidization regime without particle entrainment ($u_{sg} \leq u_t$). The particles are suspended and dried by upward flowing hot air. The air inlet temperature in fluidized bed dryers ranges between 100 and 200 °C, but operation slightly above ambient temperatures is also possible. Fluid bed dryers can be operated both batch-wise or continuously. In the latter case, however, the residence time distribution may become challenging [89]. Fluidized bed drying is fast, efficient and requires small reactor volumes because of the efficient mixing and high heat and mass transfer rates [87,89]. On the other hand, it is only applicable to fluidizable particles, i.e. with diameter ranging from 100 μm to 10 mm and with a reasonably narrow particle size distribution. Moreover, conventional fluidized bed dryers offer a limited flexibility with respect to the working operational conditions [87,89]. Pulsed fluidized beds, in which the gas flow rate oscillates periodically around a mean value to improve the particle bed uniformity, have also been used for drying [93,94].

Pneumatic or flash dryers (Figure 3.7b) make use of the riser regime (see section 1.4.2), i.e. the fluidization regime in which the material to be dried, continuously fed, is fully entrained. Such dryers are mostly used to rapidly remove small amounts of moisture from small particles. Pneumatic dryers also include ring dryers (Figure 3.7c), suited for bigger particles, in which the particles are fluidized in an endless ring duct. The main drawback of pneumatic dryers is their large size and high drying air throughput resulting from the low solids volume fraction in the reactor [87,89].

Rotating fluidized beds have also been considered for drying. The use of centrifugal fluidized beds with a rotating chamber for the continuous drying of diced food materials was studied in the '70s [51,52], but industrial applications were limited because of challenges with respect to sealing, vibrations and continuous operation [53–55]. The use of vortex chambers, in which a so-called rotating fluidized bed in a static geometry (RFB-SG) is generated, for the removal of superficial moisture from several materials such as methylsiloxane-nylon-6 (MSN) copolymer, sulfanilamide, melalite, emulsion polystyrene or sand, was also studied by Kochetov et al [57,58] and by Volchkov et al. [68]. Few results on the drying of wheat grain were also reported [68].

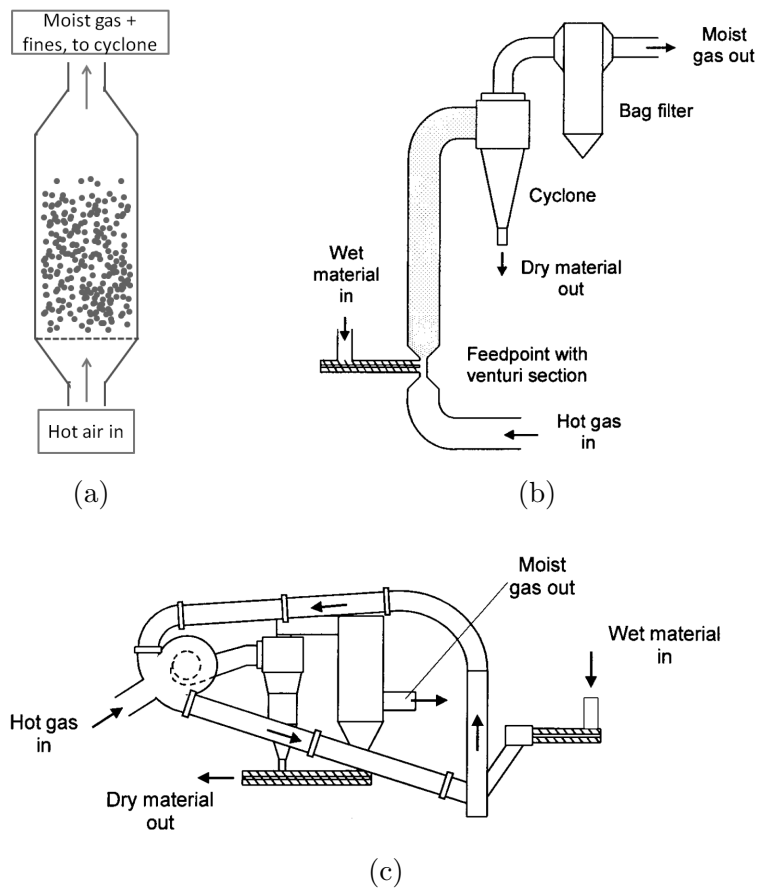


Figure 3.7: (a) Fluidized bed dryer, (b) flash dryer [89] and (c) ring dryer [89].

3.2.5 Use of alternative forms of energy

Microwave dryers and solar energy are also used to dry biomass. Open-air drying using solar energy is cheap but very slow and requires large surface areas. Oppositely, microwave drying allows fast drying with an accurate control of the operating conditions and a good energy efficiency. On the other hand, microwave drying is costly [83].

3.3 Particle drying stages

In conventional biomass dryers, the drying typically occurs in two stages. In the first stage - the "constant rate period" - the vapour pressure at the particle surface equals the saturation pressure at the surface temperature. Interfacial heat and mass transfers determine the drying rate [82–84]. The second stage of drying starts once the vapour pressure at the particle surface decreases below the saturated vapour pressure. During this so-called "falling rate period", the drying rate is also determined by intra-particle diffusion limitations. The transition between the two stages and their relative importance depend on the structure of the biomass, e.g. its porosity and pore size distribution, and the operating conditions [82–84, 95–97].

Figure 3.8 shows typical curves of the solid moisture content, air outlet humidity and air outlet temperature versus time for batch drying of a porous material. These curves have been obtained for the drying of 175 g of 1 mm diameter yeast particles in a 0.18 m diameter, 0.34 m tall fluidized bed. The air flow rate of 135 Nm³/h was injected at a constant temperature of 27 °C [95].

The first zone corresponds to the transient behavior related to the feeding of the material to be dried. The absolute air outlet humidity increases because water vapor is transferred from the solids to the air while the air outlet temperature decreases because of the latent heat of vaporization. The second zone corresponds to the constant rate period where the solids moisture content is decreasing linearly with time. The quantity of evaporated water transferred from the solid to the gas phase is thus constant, leading to constant air outlet humidity and temperature during this period. The third zone begins when the intra-particle diffusion limitations become a limiting factor. In this falling drying rate period, the drying rate decreases with time and, as a result, the outlet

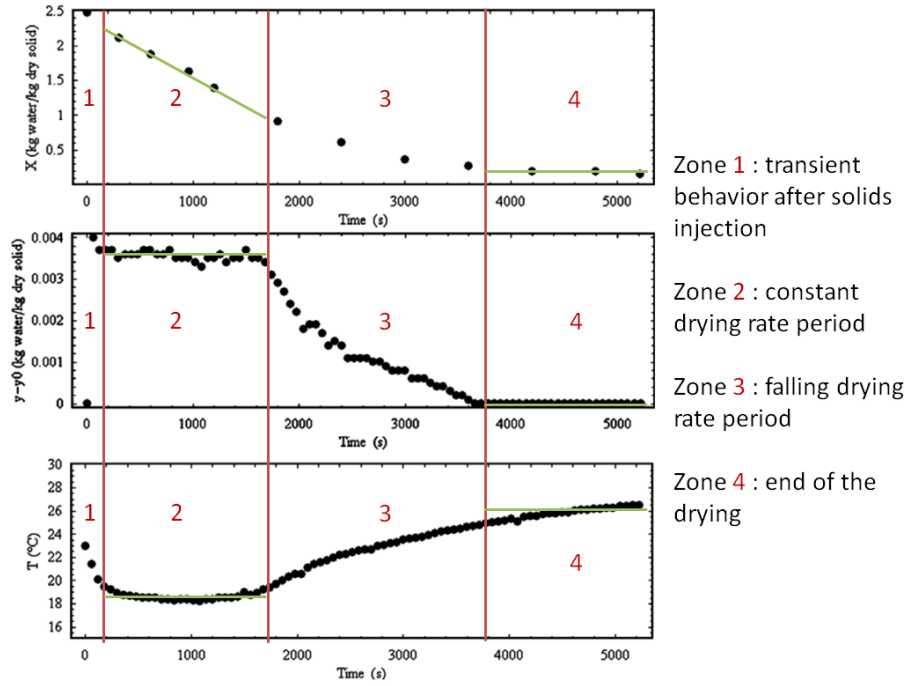


Figure 3.8: Typical curves of solid moisture content, air outlet humidity and air outlet temperature versus time for batch drying of a porous material [95].

3.3. Particle drying stages

air humidity decreases towards the inlet air humidity while the outlet air temperature increases towards the inlet air temperature. The fourth zone corresponds to the end of the drying. The solid moisture content has reached a residual value while the outlet air humidity and temperature are equal to the inlet values by lack of water transfer.

Drying of biomass particles: Experimental study and comparison of the performance of a conventional fluidized bed and a rotating fluidized bed in a static geometry

The content of this chapter has been published in Drying Technology, Vol. 31 (2), pp. 236-245, 2013.

4.1 Introduction

Because of the different factors limiting the drying rate, state-of-the-art biomass dryers are relatively large devices. To improve the economy of woody biomass based processes and facilitate local treatment prior to transport, smaller and more efficient dryers are needed.

Rotating fluidized beds allow intensifying fluidized bed processes by operating in a high-gravity (high-G) field [53–55]. The latter in particular allows maintaining a dense and uniform particle bed at high gas-solids slip velocities, as such intensifying interfacial heat and mass transfer and opening perspectives for developing dryers of reduced size. The intensity of interfacial heat and mass transfer is reflected in the specific biomass drying rate, expressed per cubic meter reactor. High-gravity operation also allows fluidizing smaller, even cohesive particles. Two types of rotating fluidized beds exist: those with either (i) a rotating chamber or

(ii) a (static) vortex chamber. The use of centrifugal fluidized beds with a rotating chamber for the continuous drying of diced food materials has been considered in the 70s [51, 52], but industrial applications were limited because of challenges with respect to sealing, vibrations and continuous operation [53–55]. Therefore, vortex chambers, in which a so-called rotating fluidized bed in a static geometry (RFB-SG) is generated, are focused on. The use of a vortex chamber for the removal of superficial moisture from several materials such as methylsiloxane-nylon-6 (MSN) copolymer, sulfanilamide, melalite, emulsion polystyrene or sand, was previously studied by Kochetov et al [57, 58] and by Volchkov et al. [68]. Few results on the drying of wheat grain were also reported [68]. No direct comparison with the performance of a conventional fluidized bed was, however, made. Furthermore, a detailed investigation of the drying of porous materials is lacking and the chamber design requires further optimization.

This chapter presents experimental results of porous biomass drying in both a conventional fluidized bed and a vortex chamber or RFB-SG. Comparison of their performance allows quantifying the process intensification that can be achieved by making use of a vortex chamber. The product quality and the efficiency of the air utilisation are also analysed.

4.2 Rotating fluidized bed in a static geometry

A RFB-SG makes use of a vortex chamber [57, 58, 68]. As shown in Figure 4.1, the particle bed rotational motion results from introducing the process gas tangentially in the vortex chamber, through multiple gas inlet slots in its outer cylindrical wall [60]. Upon contact with particles, the gas phase motion quickly shifts from tangential to mainly radial and the gas leaves the vortex chamber via a centrally positioned chimney. Due to their inertia, the particles follow a different flow path and form a rotating particle bed. The latter can be radially fluidized, but it does not have to be. Whether fluidization in the radial direction occurs depends on the magnitudes of the radial gas-solid drag force and the counteracting centrifugal force. Because these forces both depend on the gas flow rate, balancing them requires a careful vortex chamber design [60–62]. With the vortex chamber design and operating conditions used in the present work, measurements of the particle bed rotational speed showed

4.2. Rotating fluidized bed in a static geometry

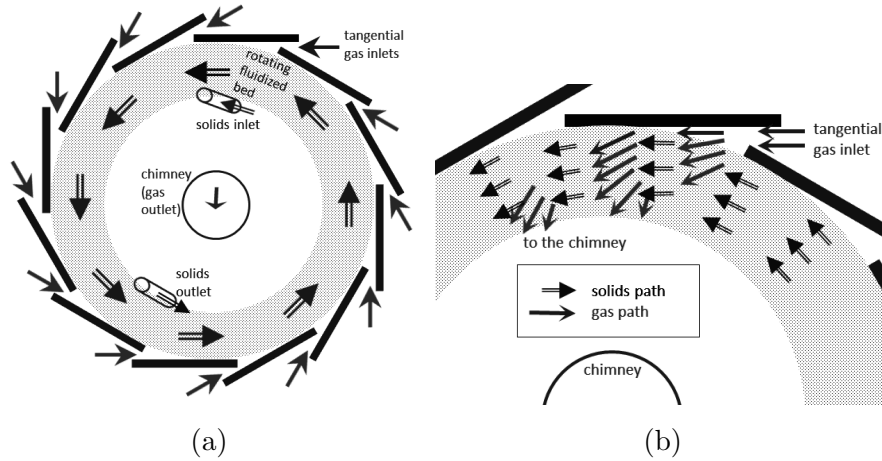


Figure 4.1: The use of a vortex chamber to generate a rotating fluidized bed in a static geometry (RFB-SG).

that the centrifugal force experienced by a particle is significantly higher than the radial gas-solid drag force, explaining why no radial fluidization of the particle bed was observed.

The gas flow rate having a similar effect on the radial gas-solid drag force and the counteracting centrifugal force results in a unique flexibility with respect to the gas flow rate [60–62]. The tangential fluidization of the particle bed, on the other hand, improves radial mixing in the particle bed [60–62].

Experimental and numerical studies of the fluidization behavior of different types of particles in vortex chambers have been carried out [61, 62]. Based on the obtained data, guidelines for vortex chamber design, including optimal dimensional ratios, have been developed [57, 76, 98]. Rosales Trujillo and De Wilde [76] showed how to adapt the gas inlet design - i.e. the number and size of the gas inlet slots - to the type of particles being fluidized in order to balance the centrifugal force and the radial gas-solid drag force. In general, the fluidization of smaller particles requires smaller gas inlet slots. In parallel, the potential of vortex chambers for a number of applications has been evaluated, e.g. for carrying out heterogeneous catalytic processes [77, 99]. The poten-

tial of improved interfacial heat transfer was numerically demonstrated by de Broqueville and De Wilde [27]. The present chapter reports on an experimental evaluation of the potential of vortex chambers for the drying of biomass.

Limited information on the scale-up of vortex chambers is available in the literature, especially for fluidizing small particles. For larger particles, like those used in the present work, Kochetov et al. [57] found experimentally that when scaling up a vortex chamber, ranges for different dimensional ratios have to be respected:

- The ratio between the chimney diameter and the chamber diameter should range between 0.3 and 0.5.
- The ratio between the chamber length and the chamber diameter should range between 0.2 and 0.5.
- The ratio between the cumulative inlet slot width and the chamber diameter should range between 0.05 and 0.2.

Rosales Trujillo and De Wilde [76] found that for fine powders, other ranges apply. A more detailed investigation is required. Scale-up of vortex chambers has, however, been proven and industrial vortex chamber devices for biomass drying and cleaning have been operated [70, 80].

4.3 Experimental set-up

The biomass used in this study was 4 mm long, 4 mm diameter cylindrical beech wood particles (Figure 4.2) with a density of about 600 kg/m^3 on a dry basis. The wood particles were made by cutting long stems of 4 mm diameter beech wood. According to their size and density, the particles can be classified in the 1G-Geldart D group [17]. Their superficial minimum fluidization velocity and terminal velocity are respectively 0.8 and 7.8 m/s in the earth gravitational field. The biomass particles were mechanically sufficiently strong and no attrition was observed during the experiments. Prior to an experiment, the particles were wetted in water for 24h to reach an absolute moisture content of $850 \text{ g}_{\text{water}}/\text{kg}_{\text{dry biomass}}$ after removing the superficial moisture using a towel. The biomass was then dried either batch-wise or continuously, both in a conventional fluidized bed and in a rotating fluidized bed in a

4.3. Experimental set-up



Figure 4.2: Beech wood particles.

static geometry. Continuous biomass drying will be focused on.

The conventional fluidized bed (FB) was a 0.1 m diameter, 1.5 m tall cylindrical column with a volume of $11.8 \cdot 10^{-3} \text{ m}^3$. The air was distributed via a cone and a perforated plate and the solids were fed near the bottom of the column via an inlet in the side wall. The vortex chamber, referred to as RFB-SG in what follows, had a 0.43 m diameter and a 0.1 m diameter chimney. The length of the RFB-SG was 0.05 m and its volume $6.9 \cdot 10^{-3} \text{ m}^3$. In the RFB-SG, the air was fed via 72 gas inlet slots. The slots were 2 mm wide and 30° inclined with respect to the tangent. The solids enter the RFB-SG via a hole in one of the end plates. As a comparison, the vortex chambers used by Kochekov et al. [58] and Volchkov et al. [68] had diameters between 0.12 and 0.36 m and lengths between 0.024 and maximum half the chamber diameter, the latter to avoid an axial non-uniform distribution of the particles in the chamber. The vortex chambers used by Kochetov et al. [58] have a limited number of gas inlet slots, one single, 5 non-evenly distributed or 6 evenly distributed. Volchkov et al. [68] indicated that a sufficiently high number of gas inlet slots (> 20) allowed to optimize the circulation rate of the particle bed for a given gas flow rate per unit chamber length. An important design parameter is the fraction of the chamber circumference occupied by the gas inlet slots, λ . For the RFB-SG used in the present work this fraction is 0.107. For the chambers used by Kochetov et al. [58], it was between 0.006 and 0.048; for those used by Volchkov et al. [68] between 0.03254 and 0.112. A theoretical relation

between the required λ and the fluid and particle characteristics can be derived [76], showing that for the relatively large particles used in this work, $\lambda = 0.107$ is sufficiently small.

In all experiments, the air flow rate was controlled by means of a mass flow controller (Bronkhorst model F-106EI, Gefran, Olen Belgium). The solids feeding rate was controlled by means of a screw (Acrison model AI-105-00, TBMA, Wondelgem Belgium) and a calibrated sealing rotary valve (TBMA MAX-150, TBMA, Wondelgem Belgium). In both the FB and the RFB-SG, no separate solids outlet was foreseen. This implies that in continuous operation, the biomass is forced to leave the chambers via the gas outlet, after entrainment into the particle bed freeboard region. To calculate the residence time of the biomass in the particle bed, the solids loading in the chambers was also measured. The biomass leaving the vortex chamber was quickly separated from the gas by means of two specially designed cyclones [50] and its quantity measured by means of on-line balances (Sartorius F61S-D2 and F300S-D2, Sartorius, Vilvoorde Belgium).

In the experiments with the FB, the air flow rate was $110 \text{ Nm}^3/\text{h}$ with a corresponding fluidization velocity of $4.5 \text{ m}_g^3/(\text{m}_r^2.\text{s})$. At this gas flow rate, a spouted fluidized bed behaviour and bubbling were observed with the biomass particles, as expected with Geldart D-group particles. Studies of the gas distributor geometry have been carried out to obtain more uniform beds with reduced bubbling [9], but the latter can never be completely eliminated in conventional fluidized beds. The gauge pressure upstream of the air distributor was 110 mbar. The biomass feed rate was varied between 1 and 9 g/s on a humid basis. The FB could not be operated at higher gas flow rates without exceeding the terminal velocity of the particles resulting in pronounced particles entrainment. The RFB-SG can be operated at much higher gas velocities. Experiments were carried out with an air flow rate of $700 \text{ Nm}^3/\text{h}$ which resulted in a dense, uniform, pseudo-fluidized bed, as can be seen in Figure 4.3c. As expected from the radial force balance on the particles, no radial fluidization or bubbling was observed. The gauge pressure upstream of the gas distribution chamber was 400 mbar. The biomass feed rate was varied between 3 and 18 g/s on a humid basis. With the given air flow rate, the gas-solid slip velocities in the RFB-SG were only slightly higher than in the FB. Operation at higher gas flow rates in the

4.3. Experimental set-up

	Conventional fluidized bed	Rotating fluidized bed in a static geometry
Dimensions	$D = 0.10$ m $H = 1.50$ m $V = 11.8 \cdot 10^{-3}$ m ³	$D = 0.43$ m $D_{\text{chimney}} = 0.10$ m $L = 0.05$ m $V = 6.9 \cdot 10^{-3}$ m ³
Gas distribution	Cone and perforated plate	72 gas inlet slots: width = 2 mm; inclination with tangent = 30 °
Solids feeding	Via side wall at $h = 0.05$ m	Via end plate
Particle characteristics	Beech wood, cylindrically shaped, $d_p = 4$ mm, $h_p = 4$ mm	
Operating conditions		
T_{bed} [K]	318.15	
p_{out} [Pa]	101325	
Gas mass flow rate [Nm ³ /h]	110	700
Pressure upstream of the gas distributor [bar]	1.1	1.4
Compressor estimated power consumption [kW]	0.31	7.8
Solids mass feed rate [g _{humid biomass} /s]	1, 2, 3, 6, 9	3, 6, 9, 12, 15, 18
Solids inlet moisture content [kg _{water} /kg _{dry biomass}]	0.850	

Table 4.1: Drying chamber characteristics and operating conditions.

RFB-SG is, however, possible but requires a more powerful compressor. The dimensions of the drying chambers and the operating conditions are summarized in Table 4.1. Pictures of the experimental set-up and drying chambers are shown in Figure 4.3. It should be remarked that with smaller particles, radial fluidization and bubble formation occur more easily. To avoid these phenomena, the chamber has to be modified as discussed by Rosales Trujillo and De Wilde [76].

Since the rates of heat and mass transfer strongly depend on the particle bed temperature, comparison of the conventional and rotating fluidized bed performances is facilitated by operating at equal particle bed temperature. The biomass was dried at the relatively low temperature of 318.15 K (45 ° C). The particle bed temperature was controlled by means of a PID controller connected to a fluidization gas pre-heater (12 kW resistor: Kamenev, Liège Belgium). The air flow was delivered by a

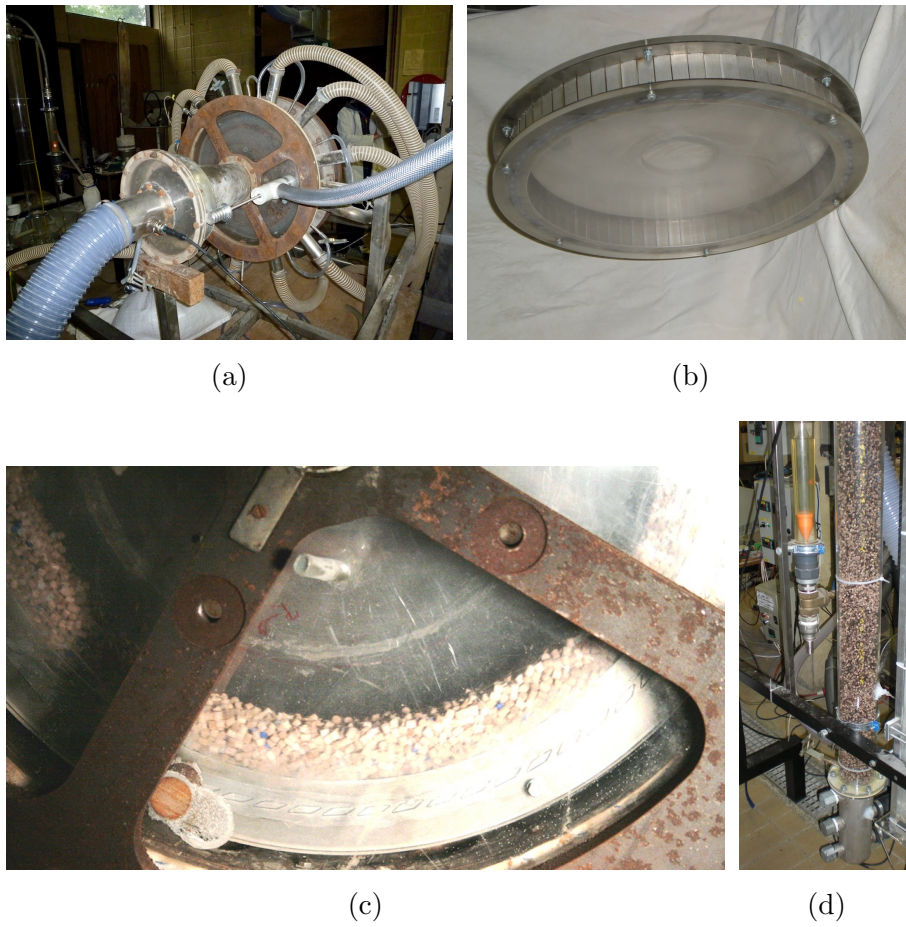


Figure 4.3: The use of a vortex chamber to generate a rotating fluidized bed in a static geometry (RFB-SG).

Roots Blowers RVT70 (Aerzen, Kortenberg Belgium) compressor. The air humidity, temperature and pressure at the inlet and outlet of the drying chambers were continuously measured (with respectively: Rotronic HC2, Krautli, Dilbeek Belgium; PT100, Fabritius, Brussels Belgium; and Gefran TSA probes, Gefran, Olen Belgium). After reaching continuous, that is, steady state operation, samples of the biomass were taken, as well as from the feed, out of the drying chamber and from the outlet, and their moisture content determined. Because the product uniformity - related to the particle bed hydrodynamics - is an important issue, the distribution of the moisture content of the biomass leaving the drying chamber was also studied.

4.4 Results and discussion

Three aspects are focused on. First, the drying chamber volume required to dry a given biomass feed rate to a certain moisture content - in other words, the intensity of the process using FB and RFB-SG type drying chambers. Second, the product uniformity and related chamber design aspects. Third, the required air flow rates and the air usage efficiency.

4.4.1 Process intensification and product quality

The experimentally measured (a) air inlet and outlet humidity, (b) moisture content of the biomass fed, biomass in the drying chamber and biomass leaving the drying chamber, (c) solids loading in the drying chamber, and (d) residence time of the biomass in the particle bed are shown in Tables 4.2 and 4.3 for respectively the FB and the RFB-SG and as a function of the biomass feed rate. For each set of operating conditions, the reproducibility of the experimental results was verified and the experimental errors measured. The mean standard deviations are also reported in Tables 4.2 and 4.3.

A comparison between the drying performance of the FB and the RFB-SG is shown in Figure 4.4. The mean biomass moisture content at the outlet of the drying chamber is shown as a function of the specific biomass feed rate, i.e. the biomass feed rate per m^3 drying chamber. The experimental errors obtained from repetitive experiments are also shown in Figure 4 and illustrate that the differences in behaviour be-

Chapter 4: Drying of biomass particles: Experimental study and comparison of the performance of a FB and a RFB-SG

$\dot{m}_{\text{humid biomass}}$ [g _{humid biomass} /s]	$H_{\text{air-in}}$ [g _{water} /kg _{dry air}]	$H_{\text{air-out}}$ [g _{water} /kg _{dry air}]	m_{bed} [g _{humid biomass}]	τ_{biomass} [s]
1	5.10 ± 0.50	8.75 ± 0.25	160.40 ± 1.90	236.40 ± 6.84
2	4.80 ± 0.40	13.25 ± 0.25	223.05 ± 15.95	158.03 ± 9.01
3	6.10 ± 0.30	18.00 ± 2.00	170.10 ± 29.30	79.12 ± 14.13
6	6.70 ± 0.20	21.50 ± 3.50	341.85 ± 21.95	72.76 ± 6.41
9	6.90	30.00	469.00	63.13

$\dot{m}_{\text{humid biomass}}$ [g _{humid biomass} /s]	$H_{\text{biomass-in}}$ [kg _{water} /kg _{dry biomass}]	$H_{\text{biomass-out}}$ [kg _{water} /kg _{dry biomass}]	$H_{\text{biomass-bed}}$ [kg _{water} /kg _{dry biomass}]
1	0.872 ± 0.015	0.203 ± 0.002	0.271 ± 0.041
2	0.884 ± 0.000	0.287 ± 0.005	0.328 ± 0.019
3	0.865 ± 0.000	0.394 ± 0.002	0.338 ± 0.009
6	0.870 ± 0.001	0.513 ± 0.019	0.467 ± 0.034
9	0.875	0.572	0.547

Table 4.2: Biomass drying in a FB. Inlet and outlet air humidity; moisture content of the biomass fed, the biomass in the chamber and the biomass leaving the chamber (chimney); mass of the biomass particle bed; and biomass residence time for different biomass feed rates. Experimental set-up and conditions: see Table 4.1.

$\dot{m}_{\text{humid biomass}}$ [g _{humid biomass} /s]	$H_{\text{air-in}}$ [g _{water} /kg _{dry air}]	$H_{\text{air-out}}$ [g _{water} /kg _{dry air}]	m_{bed} [g _{humid biomass}]	τ_{biomass} [s]
3	4.80 ± 0.10	9.90 ± 0.10	788.05 ± 1.95	413.81 ± 0.30
6	5.65 ± 0.75	13.00 ± 0.90	841.15 ± 12.25	204.50 ± 3.09
9	6.35 ± 0.15	15.95 ± 0.25	862.70 ± 5.70	132.91 ± 0.50
12	5.35 ± 0.25	16.55 ± 0.45	888.40 ± 11.30	101.84 ± 0.41
15	5.25 ± 0.55	18.30 ± 0.80	887.90 ± 15.00	82.10 ± 0.71
18	5.30 ± 0.10	18.40 ± 0.10	875.65 ± 7.25	65.28 ± 0.63

$\dot{m}_{\text{humid biomass}}$ [g _{humid biomass} /s]	$H_{\text{biomass-in}}$ [kg _{water} /kg _{dry biomass}]	$H_{\text{biomass-out}}$ [kg _{water} /kg _{dry biomass}]	$H_{\text{biomass-bed}}$ [kg _{water} /kg _{dry biomass}]
3	0.843 ± 0.027	0.138 ± 0.021	0.170 ± 0.019
6	0.857 ± 0.012	0.241 ± 0.006	0.273 ± 0.007
9	0.831 ± 0.011	0.358 ± 0.015	0.321 ± 0.004
12	0.819 ± 0.027	0.377 ± 0.002	0.322 ± 0.008
15	0.857 ± 0.011	0.417 ± 0.004	0.338 ± 0.003
18	0.840 ± 0.018	0.442 ± 0.001	0.371 ± 0.012

Table 4.3: Biomass drying in a RFB-SG. Inlet and outlet air humidity; moisture content of the biomass fed, the biomass in the chamber and the biomass leaving the chamber (chimney); mass of the biomass particle bed; and biomass residence time for different biomass feed rates. Experimental set-up and conditions: see Table 4.1.

4.4. Results and discussion

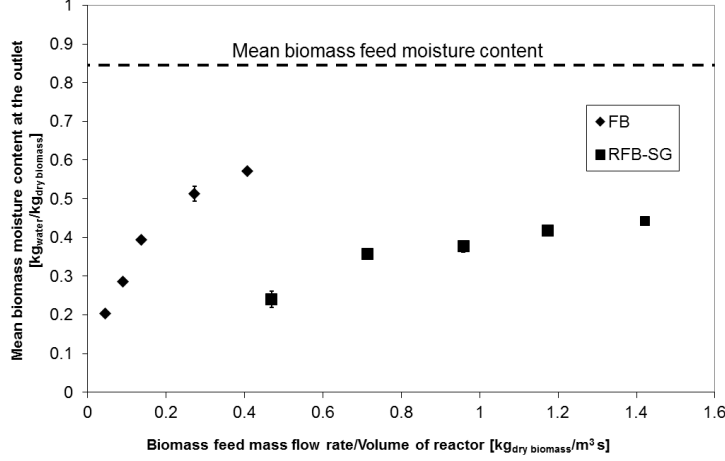


Figure 4.4: FB and RFB-SG biomass drying performance. Mean biomass moisture content at the outlet of the drying chamber as a function of the specific biomass feed rate. Experimental set-up and conditions: see Table 4.1.

tween the FB and the RFB-SG are statistically significant.

As shown in Tables 4.2 and 4.3, the solids hold-up or particle bed mass increases with increasing biomass feed rate in the FB (except for $\dot{m}_{\text{humid biomass}} = 3 \text{ g/s}$), but hardly in the RFB-SG. The latter allows a easier control of the biomass residence time as well as operation in a wider biomass residence time range in the RFB-SG. With increasing biomass feed rate and decreasing related biomass residence time, the biomass outlet moisture content increases both in the FB and the RFB-SG - as expected. As shown in Tables 4.2 and 4.3 and from Figure 4.4, the biomass outlet moisture content increases less rapidly with decreasing biomass residence time in the RFB-SG than in the FB, offering increased flexibility with respect to the biomass feed rate when using an RFB-SG, especially for initial drying of biomass.

Figure 4.4 demonstrates that for a given mean biomass moisture content at the outlet of the drying chamber, the RFB-SG allows much higher specific biomass feed rates than the FB, as already suggested by Kochetov et al. [58] and Volchkov et al. [68]. In other words, the same biomass flow rate can be treated in a significantly smaller drying

chamber when using a RFB-SG. The improvement can be quantified by means of a process intensification (PI) factor defined as the ratio of the specific drying rates in the RFB-SG and in the FB:

$$\text{PI factor} = \frac{(\text{sp. drying rate})_{\text{RFB-SG}}}{(\text{sp. drying rate})_{\text{FB}}} \quad (4.1)$$

The specific drying rate is the quantity of water transferred per unit time and per m^3 drying chamber. It can be calculated based on either the air humidity or the biomass moisture content:

$$\begin{aligned} \text{sp. drying rate} &= \frac{\dot{m}_{\text{dry biomass}}}{V} (H_{\text{biomass-in}} - H_{\text{biomass-out}}) \\ &= \frac{\dot{m}_{\text{dry air}}}{V} (H_{\text{air-out}} - H_{\text{air-in}}) \end{aligned} \quad (4.2)$$

Consequently, the PI factor accounts for the reduction of the biomass moisture, the biomass feed rate, and the required drying chamber volume.

PI factors between 10 and 13 were obtained using the RFB-SG, depending on the mean biomass moisture content at the outlet of the drying chamber. The PI factor decreases as the desired biomass outlet moisture content decreases due to diffusion limitations within the pores of the biomass particles. The latter could be avoided by using smaller biomass particles. Three factors contribute to the process intensification that can be achieved using an RFB-SG:

- Higher gas-solid slip velocities than in the FB and as a result higher coefficients of interfacial mass and heat transfer.
- A higher particle bed density than in the FB.
- A more uniform particle bed than in the FB.

With the drying chamber geometries used in the present study and the available compressor capacity, the gas-solid slip velocities in the RFB-SG could only be slightly higher than those in the FB (9 m/s in the RFB-SG compared to 5 m/s in the FB). The PI factor exclusively due to the improvement of interfacial mass and heat transfer is therefore only about 1.3. Hence, most of the process intensification is due to the increased particle bed density and the improved particle bed uniformity in the RFB-SG. This implies that additional process intensification

4.4. Results and discussion

could be achieved with a more powerful compressor or a smaller RFB-SG drying chamber.

Figure 4.5 compares the specific drying rates expressed per m^3 drying chamber (a) and per g dry solids (b). Whereas (a) reflects the overall PI that can be achieved by using the RFB-SG, (b) reflects the PI exclusively due to the improvement of the interfacial mass and heat transfer and of the particle bed uniformity in the RFB-SG. The comparison of (a) and (b) and the knowledge of the PI factor due to the improvement of interfacial mass and heat transfer allow the separate calculation of the PI factor resulting from the improvement of the particle bed uniformity (about 1.5) and the PI factor resulting from the densification of the particle bed (about 5.5) - clearly the most important contribution to the overall PI for the conditions considered. In the RFB-SG, the particle bed is close to packed and only pseudo-fluidized, as can be seen from Figure 4.3c. Considering a biomass particle, the measured particle bed rotational speed and particle bed density allow to calculate that the centrifugal force ($F_c = \rho_s \pi d_p^3 v_t^2 / (6R) = 3.46 \cdot 10^{-3} \text{ N}$) is significantly higher than the radial gas-solid drag force ($F_{r,d} = \rho_g \pi d_p^2 C_D u_{i,r}^2 / 8 = 2.63 \cdot 10^{-4} \text{ N}$), explaining the absence of radial fluidization in the RFB-SG.

Volchkov et al. [68] indicated that aerodynamic blow-off of superficial moisture could significantly contribute to the removal of water, in particular when operating at mild temperatures. To evaluate if aerodynamic blow-off of moisture plays a significant role in the present experiments, the specific drying rates that can be theoretically achieved by evaporation only have been calculated considering interfacial mass and heat transfer and neglecting intra-particle diffusion limitations. Different correlations for the mass and heat transfer coefficients were used: Gunn [29], Dwivedi and Upadhyay [45], Balakrishnan and Pei [28] and Gupta et al. [49]. It should be remarked that the correlations predict local values of the interfacial mass and heat transfer coefficients, whereas experimentally only average values for the chamber were measured. Non-uniformities in the particle bed will affect these average values. The results are shown in Table 4.4, for the FB with a solids feeding rate of 6 g/s and for the RFB-SG with a solids feeding rate of 12 g/s, and a comparison is made with the experimental data.

Table 4.4 shows that the experimentally observed specific drying

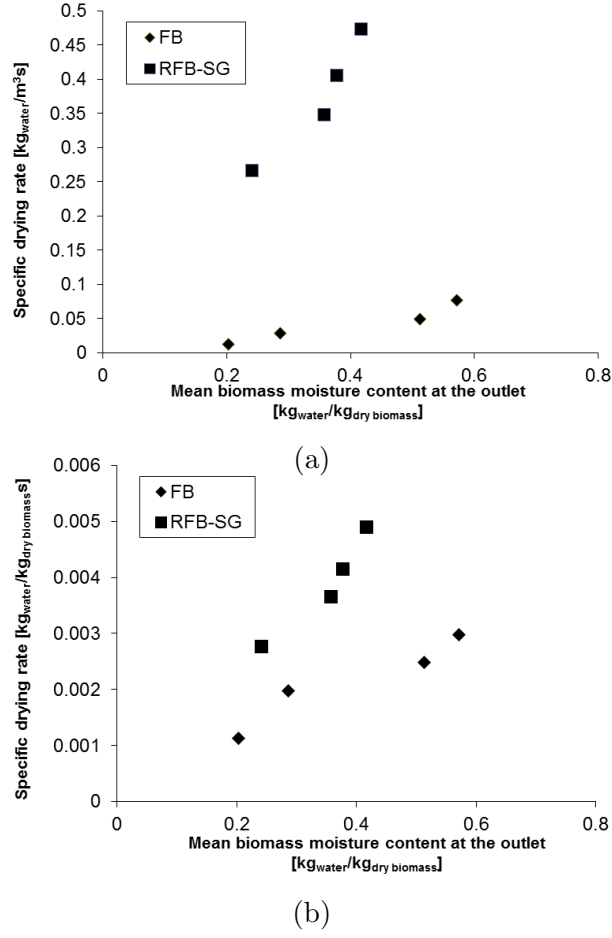


Figure 4.5: Comparison of the specific drying rates expressed (a) per m^3 drying chamber and (b) per kg dry solids. Experimental set-up and conditions: see Table 4.1.

4.4. Results and discussion

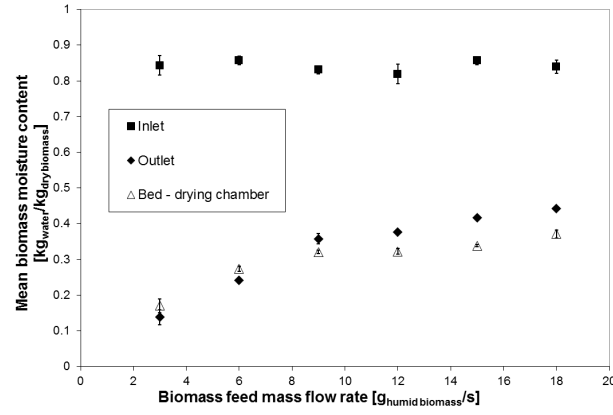
	k_g [m _g ³ /(m _i ² s)]	h_f [J/(K m _i ² s)]	sp. drying rate [kg _{water} /(m _r ³ s)]
Mass: Gunn			
Heat: Gunn			
<i>FB 6 g/s</i>	0.1718	184.5	0.35
<i>RFB-SG 12 g/s</i>	0.4598	494.2	1.23
Mass: Dwivedi & Upadhyay			
Heat: Balakrishnan & Pei			
<i>FB 6 g/s</i>	0.2016	116.9	0.25
<i>RFB-SG 12 g/s</i>	0.3775	438.3	1.07
Mass: Dwivedi & Upadhyay			
Heat: Gupta et al.			
<i>FB 6 g/s</i>	0.2016	230.7	0.43
<i>RFB-SG 12 g/s</i>	0.3775	431.1	1.06
Experiment			
<i>FB 6 g/s</i>			0.05
<i>RFB-SG 12 g/s</i>			0.41

Table 4.4: Theoretical calculations versus experimental data of the specific biomass drying rates. Theoretical values based on different correlations for the coefficients of interfacial heat and mass transfer and neglecting intra-particle diffusion limitations and non-uniformities in the particle bed. FB at 6 g/s and RFB-SG at 12 g/s solids feeding rate. Experimental set-up and other conditions: see Table 4.1.

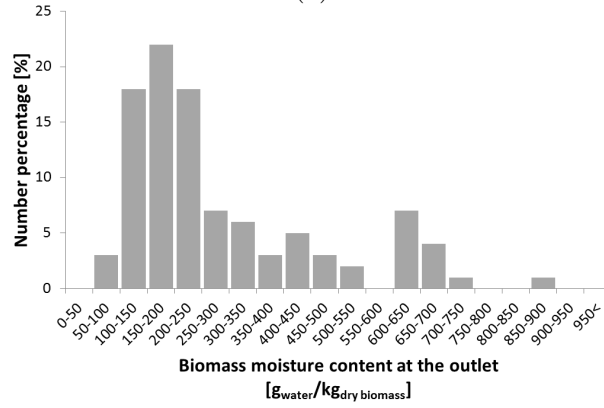
rates are lower than the theoretically predicted values, indicating that aerodynamic blow-off of moisture does not significantly contribute to the biomass drying. The lower than theoretically predicted specific drying rates result from the theoretical calculations not accounting for (i) intra-particle diffusion limitations, (ii) interfacial heat transfer limitations and (iii) non-uniformities in the particle bed, i.e. bubbling, leading to a less efficient gas-solid contact. The second factor is seen to be more pronounced in the FB than in the RFB-SG, as visually observed - see Figure 4.3.

In the FB, bubbling results in pronounced mixing in the particle bed. The particle bed hydrodynamics in the RFB-SG requires some more attention. As seen from Table 4.3, the residence time of the biomass in the particle bed is such that the biomass particles complete multiple rotations in the drying chamber (the particle bed rotates at about 4-5 rps, corresponding to about 18G for the given RFB-SG geometry and operating conditions). Hence, the behavior of the bed of biomass particles drying in the RFB-SG is expected to approach that of a completely mixed particle bed (continuous stirred tank reactor, or CSTR) on which an axial motion is superimposed. This implies that the mean moisture content of the biomass particles should be the same inside the drying chamber and in the outlet, i.e. in the chimney. Table 4.3 and Figure 4.6a show that this is reasonably well the case with biomass feeding rates of 3 and 6 g humid biomass per s. With higher biomass feeding rates, however, the biomass moisture content becomes significantly higher in the outlet than in the drying chamber. An explanation was found in a certain degree of bypassing of the biomass particles, the latter going directly from the side solids inlet into the chimney. This bypassing becomes more important with increasing biomass feeding rates. Bypassing of the particle bed by part of the biomass fed was reflected in the distribution of the moisture content of the biomass leaving the drying chamber, as shown for a solids feeding rate of 3 g humid biomass per s in Figure 4.6b. For the conditions of figure 4.6b, about 15% of the biomass fed bypasses the particle bed. An improved solids feeding system, e.g. closer to- or via the cylindrical wall, and a separate solids removal system can minimize bypassing. Different solids outlet configurations have been proposed in the literature. [63,68].

4.4. Results and discussion



(a)



(b)

Figure 4.6: Biomass drying in an RFB-SG. Experimental set-up and conditions: see Table 4.1. (a) Comparison of the biomass moisture content in the particle bed and in the drying chamber outlet. (b) Distribution of the biomass moisture content in the drying chamber outlet for a solids feeding rate of 3 g humid biomass per s.

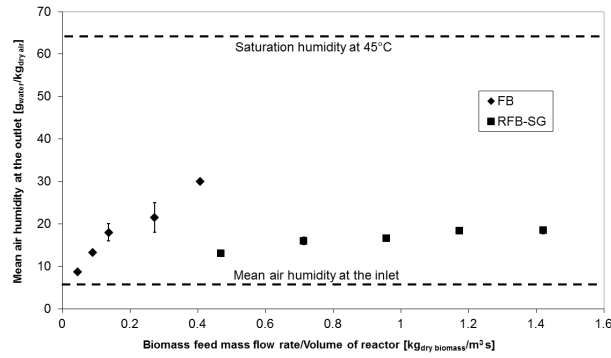
4.4.2 Air flow rates and the air usage efficiency

The air flow rate used with the RFB-SG is 6.4 times higher than that used with the FB, leading to an increased power consumption of the compressor (see Table 4.1). The efficiency of the drying air usage is therefore an important issue. Figure 4.7a compares the air humidity at the outlet of the FB and RFB-SG drying chambers for varying specific biomass feeding rates. Figure 4.7b illustrates the similar air utilization in the FB and RFB-SG at equal biomass moisture content at the outlet of the drying chamber, meaning that the ratio $\frac{\dot{m}_{\text{dry air}}}{\dot{m}_{\text{dry biomass}}}$ is similar in both reactors. For example, comparing at 0.4 kg water per kg dry biomass shows a roughly equal outlet air humidity of about 18 g water per kg dry air for the FB and the RFB-SG. The specific biomass feed rate was 0.14, respectively 1.2 kg dry biomass per m³ drying chamber and per s for the FB and the RFB-SG (Figure 4.3). In this case, the process intensification comes at the cost of a higher gas flow rate only. If the air leaving the drying chamber is far from saturated, recycling can be considered. Also, optimizing the RFB-SG drying chamber design, allowing operation at lower air flow rates, is possible. Rosales Trujillo and De Wilde [76] have shown that reducing the slots size allows maintaining the same bed rotational speed with a lower gas flow rate.

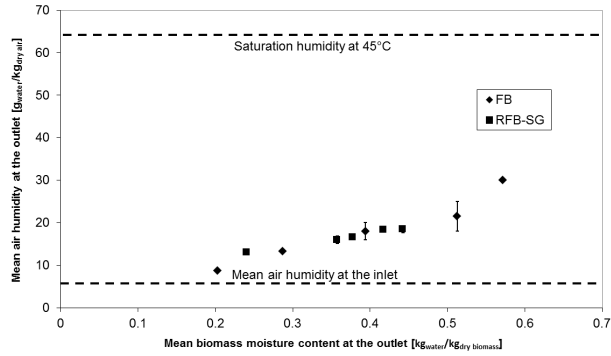
4.5 Conclusions and outlook

The drying of woody biomass using a conventional fluidized bed and using a vortex chamber based rotating fluidized bed in a static geometry has been experimentally studied. High-gravity operation allows denser and more uniform particle beds at higher gas-solid slip velocities and as such intensification of the biomass drying process. The experimental study confirms that significant process intensification - by about one order of magnitude - can be achieved by using a rotating fluidized bed in a static geometry. The process intensification is the most pronounced during the initial stage of drying, intra-particle diffusion limitations gaining importance as the outlet biomass moisture content decreases. Smaller particles could, however, also be easily treated. The rotating fluidized bed in a static geometry is operated at a much higher gas flow rate than the conventional fluidized bed, but the experimental data show that for a given biomass outlet moisture content, the air usage efficiency is maintained. Finally, in the biomass drying study carried out, the particle bed

4.5. Conclusions and outlook



(a)



(b)

Figure 4.7: Comparison of the inlet and outlet air humidity in the FB and RFB-SG as a function of (a) the specific biomass feed rate and (b) the mean biomass moisture content at the outlet of the drying chamber. Experimental set-up and conditions: see Table 4.1.

is close to completely mixed. Some bypassing of the particle bed by the biomass fed was observed, mainly at high biomass feed rates, indicating that optimization of the biomass inlet and outlet design is required. The investigation opens perspectives for the development of more compact and more economic biomass dryers for the local treatment of biomass, e.g. prior to transport.

Nomenclature

D	m_r	drying chamber diameter
H	m_r	conventional drying chamber height
$H_{\text{air-in}}$	$g_{\text{water}}/kg_{\text{dry air}}$	absolute air humidity at the inlet of the drying chamber
$H_{\text{air-out}}$	$g_{\text{water}}/kg_{\text{dry air}}$	absolute air humidity at the outlet of the drying chamber
$H_{\text{biomass-bed}}$	$g_{\text{water}}/kg_{\text{dry biomass}}$	absolute moisture content of the biomass in the drying chamber
$H_{\text{biomass-in}}$	$g_{\text{water}}/kg_{\text{dry biomass}}$	absolute moisture content of the biomass fed
$H_{\text{biomass-out}}$	$g_{\text{water}}/kg_{\text{dry biomass}}$	absolute moisture content of the biomass at the drying chamber outlet
h_f	$W/(m_i^2 K)$	heat transfer coefficient
k_g	$m_g^3/(m_i^2 s)$	mass transfer coefficient
L	m_r	vortex chamber length
m_{bed}	$g_{\text{humid biomass}}$	biomass bed mass on a humid basis
$\dot{m}_{\text{humid biomass}}$	$g_{\text{humid biomass}}/s$	biomass feed mass flow rate on a humid basis
$\dot{m}_{\text{dry biomass}}$	$g_{\text{dry biomass}}/s$	biomass feed mass flow rate on a dry basis
$\dot{m}_{\text{dry air}}$	$g_{\text{dry air}}/s$	air mass flow rate on a dry basis
PI factor	-	process intensification factor
sp. drying rate	$kg_{\text{water}}/(m_r^3 s)$	specific drying rate
V	m_r^3	drying chamber volume

Greek letters

τ_{biomass} s biomass residence time in the drying chamber

Subscripts

g gas
i interface area
r reactor

Modeling and simulation of biomass drying in vortex chambers

The content of this chapter has been submitted for publication in Drying Technology.

5.1 Introduction

The use of pre-dried biomass allows improving the efficiency of biomass combustion, pyrolysis, or gasification processes. Drying the biomass prior to transportation is economically advantageous, but requires local, small and efficient drying units. Conventional biomass dryers are of rotating drum or fluidized bed type [82–85]. Fluidized beds allow more efficient mass and heat transfer, but their performance is limited by (i) the bed density which decreases with increasing gas velocity, (ii) the non-uniformity of the particle bed with pronounced bubbling, and (iii) the gas-solid slip velocity which cannot exceed the terminal velocity of the particles in the earth gravitational field [5].

Rotating fluidized beds (RFBs) allow overcoming these limitations by operating high-G. In such reactors, the particle bed is rotating in a cylindrical chamber. The particles undergo a radially outward centrifugal force and a counteracting drag force by the gas flowing radially inward [53]. Gas-solid slip velocities and related interfacial mass, heat and momentum transfer rates can be higher than in a conventional fluidized bed while maintaining a dense and more uniform particle bed [27]. RFBs are as such promising for the development of compact dryers of

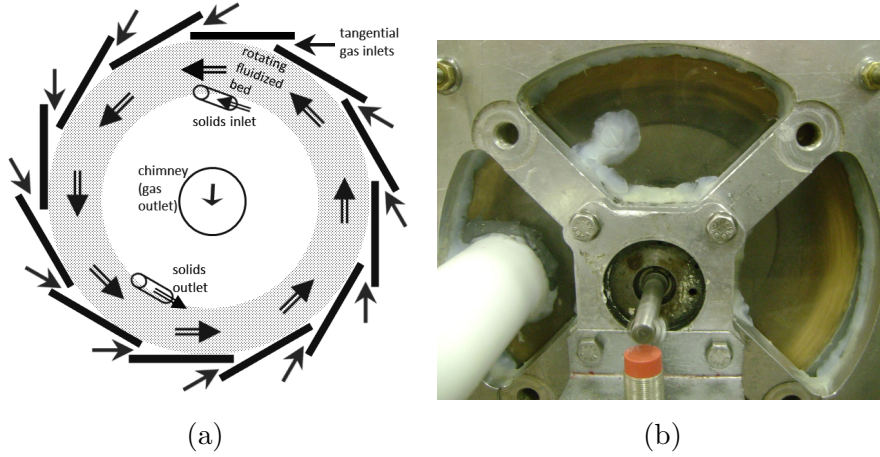


Figure 5.1: (a) The use of a vortex chamber to generate a rotating fluidized bed in a static geometry (RFB-SG). (b) Picture of a rotating bed of biomass particles.

granular materials, in particular biomass.

RFBs make use of either a rotating chamber or a static vortex chamber. RFBs with a rotating chamber have been considered for the continuous drying of diced food materials [51, 52] but industrial application is challenging because of sealing and vibrations issues. RFBs in a static geometry (RFB-SGs) make use of a vortex chamber. The particle bed rotational motion is generated by the tangential injection of the process gas through multiple gas inlet slots in the outer cylindrical wall (Figure 5.1a). The gas is deflected radially inward upon contact with the particles and leaves through a centrally positioned chimney [60]. In vortex chambers, radial fluidization is not required as the bed is tangentially fluidized. In case the bed is radially fluidized, fast entrainment of solids into the chimney has to be avoided. This requires balancing the centrifugal force and the radial gas-solid drag force. Because both forces are impacted in a similar way by the gas flow rate, the vortex chamber has to be designed to ensure this balance for given particle characteristics. With a proper design, RFB-SGs offer then an increased flexibility with respect to the gas flow rate.

Drying of porous materials typically occurs in two distinct regimes. In the first "constant drying rate" regime, the drying rate is determined

5.2. Batch experiments

by interfacial heat and mass transfer. If the drying medium inlet temperature and humidity are constant, the particle moisture content (the liquid water contained in the particles) decreases linearly with time when drying batch-wise. At a given moisture content, intra-particle diffusion, i.e. diffusion of vapor in the particle pores, becomes limiting and the drying rate decreases. This is referred to as the "falling drying rate" regime. The moisture content at which the transition occurs depends on the structure of the particles, i.e. its porosity and pore distribution, and the operating conditions [82, 95, 97]. The increased bed density and improved bed uniformity allow process intensification in both the constant and falling drying rate regimes. Only in the constant drying rate regime, additional advantage is taken from the intensified interfacial mass and heat transfer.

Vortex chambers have been successfully used to remove the superficial moisture from various non-porous materials [57, 58, 68]. More recently, Eliaers and De Wilde [100] reported experimental results on continuous (porous) biomass drying in a vortex chamber. In the present paper, the modelling and simulation of biomass drying in vortex chambers are addressed and optimization and scale-up of the technology studied. Discrimination between different simulation models and estimation of the model parameters is carried out using experimental data of both previously reported continuous [100] and additional batch drying experiments using 2 different vortex chambers. Batch experiments offer many data points per experiment, a well-known advantage of non-stationary experimental techniques. Next, with the retained model, continuous operation at typical commercial scale is focused on and different vortex chamber designs (single or multi-chamber) and air feeding systems (uniform or co- and counter-current) are studied. Maximum product quality and uniformity and minimum air consumption are aimed at. A comparison with biomass drying in a conventional fluidized bed is also made.

5.2 Batch experiments

5.2.1 Experimental set-up

Drying of 4 mm long, 4 mm diameter cylindrical beech wood particles with a density of 600 kg/m^3 on a dry basis is studied. To complement the continuous drying data of Eliaers and De Wilde [100], batch exper-

iments were carried out. Non-stationary data allow modelling with a sharply reduced number of experiments. In addition to the conventional fluidized bed and vortex chamber used by Eliaers and De Wilde [100], a smaller-diameter vortex chamber with a different design was tested. The chamber design parameters and operating conditions are summarized in Table 5.1.

To build up a dense and uniform rotating particle bed in a vortex chamber (Figure 5.1b), the centrifugal force generated by the gas injection has to be maximized. The amount of tangential momentum injected for a given gas flow rate can be increased by reducing the total inlet slot surface area, either by reducing the number of slots or their size. To guarantee a uniform gas distribution, a minimum number of slots is, however, required. Another constraint is that the tangential momentum injected has to be transferred efficiently to the particle bed. With 4 mm sized particles, this necessitates millimeter-sized slots. Tests using vortex chambers with smaller 0.2 and 0.5 mm gas inlet slots demonstrated inefficient momentum transfer to millimeter-sized particles. No stable bed could be obtained. The experimental procedure for the continuous experiments was described in [100]. The procedure for the batch experiments is explained hereafter.

In the conventional fluidized bed, the air flow rate was 40 Nm³/h, relatively low ($u_{sg} \simeq 2 \times u_{mf}^{1G}$) to minimize bubbling. At time t_0 , the pre-defined mass of humid biomass was introduced almost instantaneously in the fluidization column. The particle bed temperature was controlled via the air inlet temperature by means of a PID controller, but cannot be kept constant during a complete experiment. The vortex chamber experiments were carried out at an air flow rate of 730 N³/h or 500 Nm³/h ($u_{sg,r} \simeq 4.5 - 5.5 \times u_{mf}^{1G}$), for respectively vortex chamber 1 and 2. With the obtained bed density, the radial gas-solid slip velocity was respectively 7 and 8.5 m/s. As explained in De Wilde and de Broqueville [60], the minimum fluidization or terminal velocity in RFB-SGs cannot really be defined as both increase with increasing gas flow rate. The experimental procedure was similar to that in the conventional fluidized bed. The particles are, however, pneumatically conveyed to the vortex chamber. As a result, the solids feeding cannot be considered instantaneous. The solids feeding time was ranging between 20 and 35 s depending on the solids feeding rate and the quantity of biomass fed.

5.2. Batch experiments

	Conventional fluidized bed (FB)	Rotating fluidized bed: vortex chamber 1	Rotating fluidized bed: vortex chamber 2
Dimensions	D = 0.10 m H = 1.50 m V = $11.8 \cdot 10^{-3} \text{ m}^3$	D = 0.43 m D _{chimney} = 0.10 m L = 0.05 m V = $6.9 \cdot 10^{-3} \text{ m}^3$	D = 0.24 m D _{chimney} = 0.12 m L = 0.05 m V = $1.7 \cdot 10^{-3} \text{ m}^3$
Gas distribution	Cone and perforated plate	72 gas inlet slots: width: 2 mm; $\lambda = 0.11$; inclination with tangent: 30°	24 gas inlet slots: width: 3 mm; $\lambda = 0.10$; inclination with tangent: 30°
Solids feeding	Via side wall at h = 0.05 m	Via end plate, opposite to the chimney	

(a)

Particle characteristics	Pelletized beech wood, cylindrically shaped, $d_p = 4 \text{ mm}$, $h_p = 4 \text{ mm}$				
P _{out} [Pa]	101325				
Inlet solids moisture content [kg _{water} /kg _{dry b}]	0.8 - 0.95				
Inlet solids temperature [°C]	20				
Inlet air humidity [kg _{water} /kg _{dry a}]	$3.6 \cdot 10^{-3}$				
Experiment n °	Set-up	Mass of solids in the particle bed [g _{dry b}]	Feeding time [s]	Feed gas temperature [°C]	Gas flow rate [Nm ³ /h]
1	FB	80	instantaneous	Variable (PID)	40
2	FB	110	instantaneous	Variable (PID)	40
3	FB	135	instantaneous	Variable (PID)	40
4	FB	160	instantaneous	Variable (PID)	40
5	Vortex chamber 1	640	35	60	730
6a,b,c	Vortex chamber2	205, 195, 190	22	50	500
7a,b,c	Vortex chamber2	280, 210, 195	28	50	500
8a,b,c	Vortex chamber2	290, 290, 250	33	50	500

(b)

Table 5.1: (a) Drying chamber characteristics and (b) operating conditions for batch biomass drying experiments in a conventional fluidized bed or a rotating fluidized bed in a static geometry (vortex chamber).

The in- and outlet temperature and humidity of the air were measured on-line. During drying, biomass samples were taken from the bed to measure the evolution in time of the biomass moisture content. The sampling time was about 2 s. Each sample was weighted prior to and after drying in an oven at 110 °C to determine its moisture content. The measured gas inlet temperature and air inlet humidity profiles were imposed in the simulations of the experiments and the measured and calculated profiles of the gas outlet temperature compared, together with the measured and calculated profiles of the air outlet humidity and the biomass moisture content.

5.2.2 Results and discussion

Figures 5.5, 5.6 and 5.7 show typical profiles in time of the particle moisture content and the air temperature and humidity at the inlet and outlet of the drying chambers. In Figure 5.5 a conventional fluidized bed was used, in Figures 5.6 and 5.7 two different vortex chambers. From the moment the particles are fed, a gradual decrease of the moisture content is observed. The uniformity of the particle bed was verified by measuring the moisture content distribution in a particle sample taken at a given time t . In all experiments, the air outlet temperature is seen to decrease rapidly after feeding the particles. In most experiments (e.g. Figure 5.5), over a relatively narrow moisture content range, the drying rate is seen to be initially nearly constant. This so-called constant drying rate period is a well-known phenomenon in particle drying, indicating the saturation vapor pressure is reached at the outer surface of the particles so that the driving force for interfacial mass transfer remains constant. The saturation vapor pressure at the outer surface of the particles is determined by the bed temperature, but an equilibrium or quasi-steady state between the heat supplied and the heat consumed in the bed is normally rapidly found. The constant drying rate period is reflected in a quasi-constant air outlet temperature (until about 125 s in Figure 5.5). In some experiments the air feed temperature was adapted with a PID controller to maintain a constant air outlet temperature (e.g. Figure 5.5), but the rapid initial drop of the temperature after feeding the particles cannot be avoided. From a certain particle moisture content on, the drying is seen to slow down as a result of intra-particle diffusion limitations. This is logically reflected in an increasing air temperature at the outlet of the drying chamber.

5.3. Simulation model

In the experiments with the rotating fluidized beds in a vortex chamber (e.g. Figures 5.6 and 5.7), the constant drying rate period was seen to be very short, in certain cases overlapping with the solids feeding time. This reflects the intensification of interfacial mass and heat transfer that can be achieved by high-G operation in vortex chambers. As soon as intra-particle diffusion limitations are encountered, the intensification of interfacial mass and heat transfer becomes less relevant. Significant process intensification can, however, still be achieved by using a vortex chamber because of the increased bed density and improved particle bed uniformity both related to the intensification of interfacial momentum transfer. When comparing a rotating fluidized bed in a vortex chamber and a conventional fluidized bed, the batch experiments confirm the about one order of magnitude process intensification earlier reported for continuous biomass drying [100]. In conventional fluidized bed Experiment 2 (Figure 5.5), for example, the moisture content of 110 g of biomass (dry basis) is reduced from 0.85 to 0.4 kg_{water}/kg_{dry biomass} in 320 s and requires a 11.8 l chamber. In vortex chamber Experiments 5 (Figure 5.6), respectively 7a (Figure 5.7), a same moisture content reduction for 640/280 g of biomass (dry basis) is achieved in 115/105 s and requires a 6.9/1.7 l chamber.

The responses of the air humidity and temperature at the outlet of the drying chambers are expected to follow the evaporation rate. A correlation between the two responses is observed and explained by the way the absolute humidity is obtained. The air humidity probe measures the relative air humidity and the air temperature and from the combination of the two calculates the absolute air humidity. The reliability or accuracy of the air humidity measurement was also less than that of the biomass moisture content and air temperature. Therefore, in the modeling and parameter estimation, significantly more weight was given to the biomass moisture content and air temperature data than to the air humidity data.

5.3 Simulation model

Based on experimental observations and CFD results [67,69], the simulation model assumes plug flow for the air and a CSTR type behavior for the particle bed. The particle mixing time scale is indeed several orders

of magnitude lower than the drying time scale, both in the conventional fluidized bed and in the RFB-SGs in which the particle bed is rotating at 5 to 15 rps. The amount of sampled solids is assumed negligible compared to the total mass of the particle bed. Furthermore, the particle bed was uniformly fluidized in both the conventional and rotating fluidized bed. According to these assumptions, the species continuity equations for water in the gas and solid phase are given by

$$\varepsilon \tilde{\rho}_a \frac{dy}{dt} = -G_a \frac{dy}{dz} + \tilde{\rho}_{B,dry} \cdot a_m \cdot j \quad (5.1)$$

$$m_b \frac{dx}{dt} = \dot{m}_b (x_{in} - x) - m_b \cdot a_m \left[\frac{1}{V_B} \int_{V_B} j \, dV \right] \quad (5.2)$$

The initial and inlet conditions are:

- $x(t = 0) = x_0$
- $y(t = 0, z) = y_{in}(t = 0)$
- $y(t, z = 0) = y_{in}(t)$

In equations (5.1) and (5.2), x is the biomass moisture content and y is the absolute humidity of the air, both on a dry basis. As the absolute humidity of the air is relatively small, y is always much smaller than 1. For a RFB-SG, equation (5.1) is strictly only valid if the thickness of the bed is sufficiently small compared to the diameter of the fluidization chamber.

Assuming quasi-stationary interfacial mass transfer and that no liquid film is present on the solid surface, the local vaporization mass flux can be expressed as

$$j = k_g (p_{H_2O,s}^s - p_{H_2O}) M_{H_2O} \quad (5.3)$$

For an adiabatic reactor and isothermal particles, the gas and solid phase heat balances can be written as

$$\varepsilon \rho_g c_{p,g} \frac{dT_g}{dt} = -G_g c_{p,g} \frac{dT_g}{dz} - \tilde{\rho}_{B,dry} \cdot a_m \cdot h_f (T_g - T_s) \quad (5.4)$$

5.3. Simulation model

$$\begin{aligned}
m_b (c_{p,b} + c_{p,\text{H}_2\text{O}} \cdot x) \frac{dT_s}{dt} &= \dot{m}_b (c_{p,b} + c_{p,\text{H}_2\text{O}} \cdot x_{in}) (T_{s,in} - T_s) \\
&+ m_b \cdot a_m \left[\frac{1}{V_B} \int_{V_B} j \, dV \right] (-\Delta H_{vap}) \\
&+ m_b \cdot a_m \left[\frac{1}{V_B} \int_{V_B} h_f (T_g - T_s) \, dV \right]
\end{aligned} \quad (5.5)$$

The initial and inlet conditions are:

- $T_s(t = 0) = T_{s,0}$
- $T_g(t = 0, z) = T_{g,in}(t = 0)$
- $T_g(t, z = 0) = T_{g,in}(t)$

The isothermicity of the particles is justified as the Biot number $Bi = \frac{h_f d_p}{\lambda_{hb}}$ is always smaller than 0.1, both in the conventional and rotating fluidized beds.

The vapour partial pressures $p_{\text{H}_2\text{O}}$ and $p_{\text{H}_2\text{O},s}^s$ are linked respectively to the absolute humidities y and y_s^s according to the ideal gas law:

$$p_{\text{H}_2\text{O}} = p_{tot} y_{\text{H}_2\text{O}} = p_{tot} \frac{y}{\frac{M_{\text{H}_2\text{O}}}{M_a} + y} \quad (5.6)$$

The gas density is estimated from the ideal gas law while the air viscosity and the vapour diffusivity in air are calculated as a function of the temperature using respectively Sutherlands [101] and Bolz and Tuve's [102] equations:

$$\mu_a = 1.512 \cdot 10^{-6} \frac{T_g^{3/2}}{T_g + 120} \quad (5.7)$$

$$D_{\text{H}_2\text{O},a} = -2.775 \cdot 10^{-6} + 4.479 \cdot 10^{-8} T_g + 1.6561 \cdot 10^{-10} T_g^2 \quad (5.8)$$

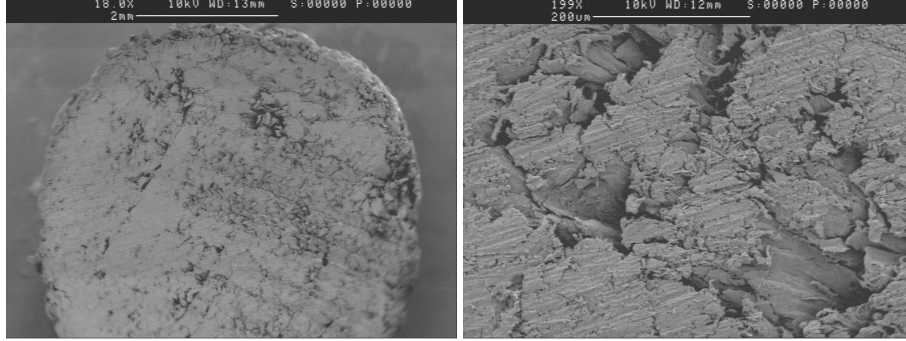


Figure 5.2: SEM pictures of the beech wood particles at different magnifications.

In the heat balances, the heat capacity of air, water and dry biomass are assumed constant ($c_{p,a} = 1005 \text{ J/(kg.K)}$, ($c_{p,\text{H}_2\text{O}} = 4186 \text{ J/(kg.K)}$), ($c_{p,b} = 2800 \text{ J/(kg.K)}$) while the latent heat of vaporization varies linearly with the solids temperature according to

$$\Delta H_{vap} = 3.175 \cdot 10^6 - 2.456 \cdot 10^3 T_s \quad (5.9)$$

The calculation of the vapor pressure at the external surface of the solid $p_{\text{H}_2\text{O},s}^s$ and the resulting vapor mass flux j can be dealt with through the different models described below. In these models, particle shrinkage is always neglected and the beech particles are considered 4 mm equivalent diameter particles. Capillarity effects are not considered important in the relatively large pores in beech wood particles [103]. SEM images of the pores are shown in Figure 5.2.

For batch drying experiments, equation (5.2) and (5.5) are simplified as no solids is fed to the reactor ($\dot{m}_b = 0$) except during the start-up phase in the RFB-SGs. For continuous drying experiments, the steady-state is focused on and, as a result, the temporal derivatives are set to 0 in equations (5.1)-(5.2) and (5.4)-(5.5).

5.3.1 Drying models 1 & 2

Drying model 1 and 2 assume a uniform repartition of the moisture in the particles and fast intra-particle diffusion leading to a uniform vapor pressure within the particle pores ($p_{\text{H}_2\text{O},s} = cste = p_{\text{H}_2\text{O},s}^s$). The

5.3. Simulation model

equilibrium between the vapor pressure within the pores and the moisture content can be accounted for by an equilibrium sorption isotherm. Among the different sorption isotherms reported in the literature, the GAB equation was selected in this work as it allows covering a wide range of vapor activities in the solid phase ($0.05 < a_{\text{H}_2\text{O}} < 0.85$) [104].

$$x = \frac{x_m c_G k_G a_{\text{H}_2\text{O}}}{(1 - k_G a_{\text{H}_2\text{O}})(1 - k_G a_{\text{H}_2\text{O}} + c_G k_G a_{\text{H}_2\text{O}})} \quad (5.10)$$

$$\text{where } a_{\text{H}_2\text{O}} = \frac{p_{\text{H}_2\text{O},s}}{p_{\text{H}_2\text{O},sat}}.$$

The three parameters c_G , k and x_m for beech wood at 20 °C are given by Bratasz et al. [105] both for water adsorption (7.11, 0.70 and 0.0698 kg_{H₂O}/kg_{dry b}) and desorption (4.88, 0.53 and 0.1192 kg_{H₂O}/kg_{dry b}). Figure 5.3 shows the sorption isotherms of beech wood, both for adsorption and desorption. A slight hysteresis behaviour is observed. The desorption isotherm is representative for the drying process while the adsorption isotherm is used for wetting. However, particles subject to multiple drying/wetting cycles, as those used in this work, do not show a hysteresis behaviour anymore. A generalized sorption isotherm, i.e. an average between the desorption and adsorption isotherms, is then used to relate x with $a_{\text{H}_2\text{O}}$ [82]. Figure 5.3 also indicates that the vapor pressure inside the particle pores is nearly equal to the saturation vapor pressure if the local moisture content is above 0.22 kg_{H₂O}/kg_{dry b}. Below a moisture content of 0.22 kg_{H₂O}/kg_{dry b}, the water is retained by the solid leading to a vapor pressure lower than the saturation vapor pressure.

The vapour saturation pressure at the solid phase temperature is calculated using Dupré's equation

$$p_{sat} = \exp \left[46.784 - \frac{6435}{T_s} - 3.868 \ln(T_s) \right] \cdot 133.322 \quad [\text{Pa}] \quad (5.11)$$

Model 2 resolves equations (5.4) and (5.5) for the heat balances while model 1 assumes infinitely fast heat transfer ($h_f \rightarrow \infty$). As a result, the gas temperature is assumed equal to the solid phase temperature and uniform in the particle bed. A global heat balance can then be written as

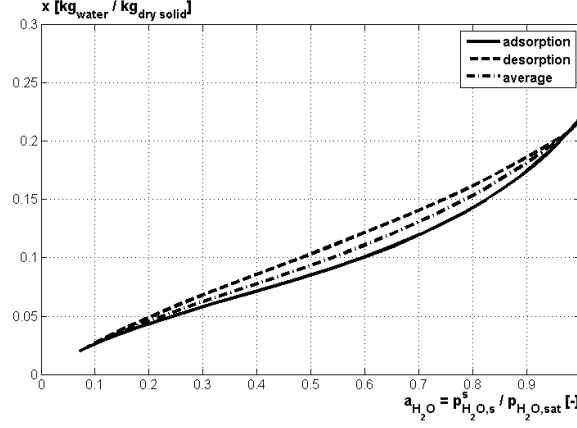


Figure 5.3: GAB equation type sorption isotherm for beech wood.

$$\begin{aligned}
 [m_b (c_{p,b} + c_{p,H_2O} \cdot x) + V_B \varepsilon \rho_g c_{p,g}] \frac{dT}{dt} &= \dot{m}_b (c_{p,b} + c_{p,H_2O} \cdot x_{in}) (T_{s,in} - T) \\
 &+ m_b \cdot a_m \left[\frac{1}{V_B} \int_{V_B} j \, dV \right] (-\Delta H_{vap}) \\
 &+ G_g \cdot S \cdot c_{p,g} (T_{g,in} - T)
 \end{aligned} \tag{5.12}$$

with initial condition: $T(t = 0) = T_{s,0}$.

5.3.2 Drying model 3

In a third drying model, a shrinking core type behaviour is assumed for the moisture content distribution inside the particles below a critical moisture content x_{crit} [16,95,103]. The moisture contents in the core and in the shell are uniform with respectively $x_{core} = x_{crit}$ and $x_{shell} = x_{res}$, a residual moisture content. Furthermore, intra-particle diffusion limitations in the dry shell are accounted for. With the core at critical moisture content shrinking, the water evaporates at the core-shell interface R_{core} . The water vapor is transported to the external surface of the particles by diffusion in the pores of the particles. Assuming quasi-stationarity of the intra-particle diffusion, neglecting intra-particle Stefan flow and accounting for the uniform solid phase temperature, the equation for the vapor pressure profile in the shell can be written

5.3. Simulation model

$$\left[\frac{D_{eff}}{RT_s} \right] \frac{d}{dr} \left(r^2 \frac{dp_{H_2O,s}}{dr} \right) = 0 \quad (5.13)$$

so that:

$$p_{H_2O,s}(r) = \frac{A}{r} + B \quad (5.14)$$

The integration constants A and B are determined by the boundary conditions $p_{H_2O,s}(R_p) = p_{H_2O,s}^s$ and $p_{H_2O,s}(R_{core}) = p_{H_2O,s}^{R_{core}}$. The vapor pressure at the core-shell interface $p_{H_2O,s}^{R_{core}}$ is linked to the moisture content x_{crit} of the humid core via the sorption isotherm (eq. (5.10)), so that $p_{H_2O,s}^{R_{core}} = a_{H_2O}(x_{crit}) \cdot p_{H_2O,sat}$. Then A and B are:

$$\begin{aligned} A &= \frac{a_{H_2O}(x_{crit}) \cdot p_{H_2O,sat} - p_{H_2O,s}^s}{\frac{1}{R_{core}} - \frac{1}{R_p}}; \\ B &= \frac{R_p p_{H_2O,s}^s - R_{core} a_{H_2O}(x_{crit}) \cdot p_{H_2O,sat}}{R_p - R_{core}}. \end{aligned} \quad (5.15)$$

With the known $p_{H_2O,s}(r)$ -profile, the vapour flux at the solid surface can be calculated from

$$\begin{aligned} j &= -M_{H_2O} \left[\frac{D_{eff}}{RT_s} \right] \frac{dp_{H_2O,s}}{dr} \Big|_{r=R_p} = M_{H_2O} \left[\frac{D_{eff}}{RT_s} \right] \frac{A}{R_p^2} \\ &= M_{H_2O} \left[\frac{D_{eff}}{RT_s} \right] \frac{a_{H_2O}(x_{crit}) \cdot p_{H_2O,sat} - p_{H_2O,s}^s}{R_p \left(\frac{R_p}{R_{core}} - 1 \right)} \\ &\triangleq k_{shell} (a_{H_2O}(x_{crit}) \cdot p_{H_2O,sat} - p_{H_2O,s}^s) M_{H_2O} \end{aligned} \quad (5.16)$$

As no water vapor accumulation can occur at $r = R_p$, $p_{H_2O,s}^s$ can be eliminated from

$$\begin{aligned} j &= k_{shell} (a_{H_2O}(x_{crit}) \cdot p_{H_2O,sat} - p_{H_2O,s}^s) M_{H_2O} \\ &= k_g (p_{H_2O,s}^s - p_{H_2O}) M_{H_2O} \end{aligned} \quad (5.17)$$

leading to

$$j = \left(\frac{1}{k_g} + \frac{1}{k_{shell}} \right)^{-1} (a_{\text{H}_2\text{O}}(x_{crit}) \cdot p_{\text{H}_2\text{O},sat} - p_{\text{H}_2\text{O}}) M_{\text{H}_2\text{O}} \quad (5.18)$$

The radial position of the core-shell interface R_{core} is found from the mean particle moisture content x :

$$\frac{4}{3}\pi R_p^3 x = \frac{4}{3}\pi R_{core}^3 x_{crit} + \frac{4}{3}\pi (R_p^3 - R_{core}^3) x_{res} \Leftrightarrow \frac{R_p}{R_{core}} \sqrt[3]{\frac{x_{crit} - x_{res}}{x - x_{res}}} \quad (5.19)$$

Drying model 3 introduces two model parameters to be determined from the experimental data, the critical moisture content, x_{crit} , and the effective diffusivity, $D_{eff} = \frac{\varepsilon_s}{\tau} D_{\text{H}_2\text{O},a}$. $D_{\text{H}_2\text{O},a}$ is calculated with the solid phase temperature according to equation (5.8).

Figure 5.4 shows a schematic representation of the different models assumptions on the moisture content, water vapor partial pressure and temperature profiles.

The mass and heat transfer coefficient are commonly modelled introducing the Colburn j-factors for mass (j_D) and heat (j_H) transfer [5]:

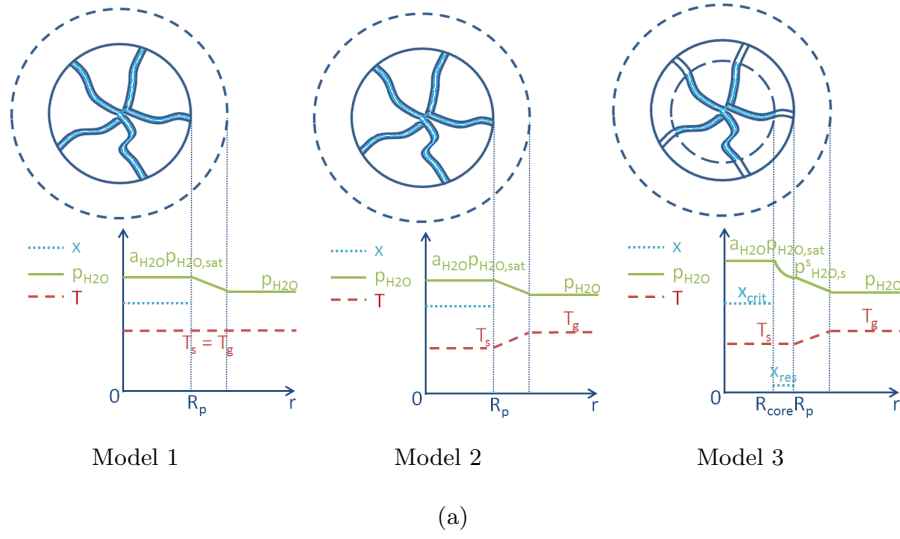
$$j_D = \frac{Sh}{ReSc^{1/3}} \Leftrightarrow k_g = j_D \frac{G_g}{y_f M_g p_{tot}} Sc^{-2/3} \quad (5.20)$$

$$j_H = \frac{Nu}{RePr^{1/3}} \Leftrightarrow h_f = j_H G_g c_{p,g} Pr^{-2/3} \quad (5.21)$$

Transfer of water vapor through a quasi-stagnant air film is assumed. The film factor y_f [5, 50] is then $(y_{a,film})_{logmean}$ with:

$$(y_{a,film})_{logmean} = \frac{(1 - y_{\text{H}_2\text{O}}) - (1 - y_{\text{H}_2\text{O},s}^s)}{\ln(1 - y_{\text{H}_2\text{O}}) - \ln(1 - y_{\text{H}_2\text{O},s}^s)} \quad (5.22)$$

5.3. Simulation model



Model	Accounts for
1	Interfacial mass transfer limitations
2	Interfacial mass and heat transfer limitations
3	Interfacial mass and heat transfer limitations Shrinking core & intra-particle diffusion limitations in the dry shell

(b)

Figure 5.4: (a) Schematic representation of the moisture content, water vapour partial pressure and temperature profiles and (b) assumptions for drying model 1, 2 and 3.

The Colburn j-factors can be calculated as a function of the particle Reynolds number $Re_p = \frac{d_p G_g}{\mu_g}$ or a modified Reynolds number $Re_{pm} = \frac{Re_p}{(1-\varepsilon)}$. For the simulations of the conventional fluidized bed, j_D and j_H are calculated from the fluidized bed correlations proposed by respectively Dwivedi et al. [45] and Gupta et al. [49]:

$$\varepsilon j_D = 0.765 Re_p^{-0.82} + 0.365 Re_p^{-0.386} \quad (5.23)$$

$$\varepsilon j_H = 2.876 Re_p^{-1} + 0.3023 Re_p^{-0.35} \quad (5.24)$$

Because in a RFB-SG the bed is dense, nearly packed, the use of the correlations of Yoshida et al. [39] and Handley and Heggs [48] for a packed bed has also been considered.

$$j_D = 0.84 \left(\frac{Re_{pm}}{6} \right)^{-0.51} \quad \text{if } \left(\frac{Re_{pm}}{6} \right) \leq 50$$

$$0.57 \left(\frac{Re_{pm}}{6} \right)^{-0.41} \quad \text{if } \left(\frac{Re_{pm}}{6} \right) > 50 \quad (5.25)$$

$$\varepsilon j_H = 0.255 Re_p^{-0.335} \quad (5.26)$$

5.4 Model discrimination and parameter estimation

Model discrimination and estimation of the model parameters x_{crit} and $\frac{\varepsilon_s}{\tau}$ are based on the comparison of experimentally measured and simulated data. Only the new batch experiment data were used for the estimation of the model parameters. The same parameter values were then used to simulate the previously published continuous drying experiments [100].

Figures 5.5, 5.6 and 5.7 compare for respectively batch Experiments 2, 5 and 7a the experimentally measured and simulated moisture content, air outlet humidity and air outlet temperature profiles. Simulations with the three different models are shown. For the heat and mass transfer coefficients, the correlations of Dwivedi [45] and Gupta [49] were used for the conventional fluidized bed. For the dense and uniform rotating fluidized bed in a vortex chamber, the Yoshida [39] and Handley [48] correlations for packed beds gave slightly better results. For the model

5.5. Simulation of an industrial scale vortex chamber for continuous biomass drying

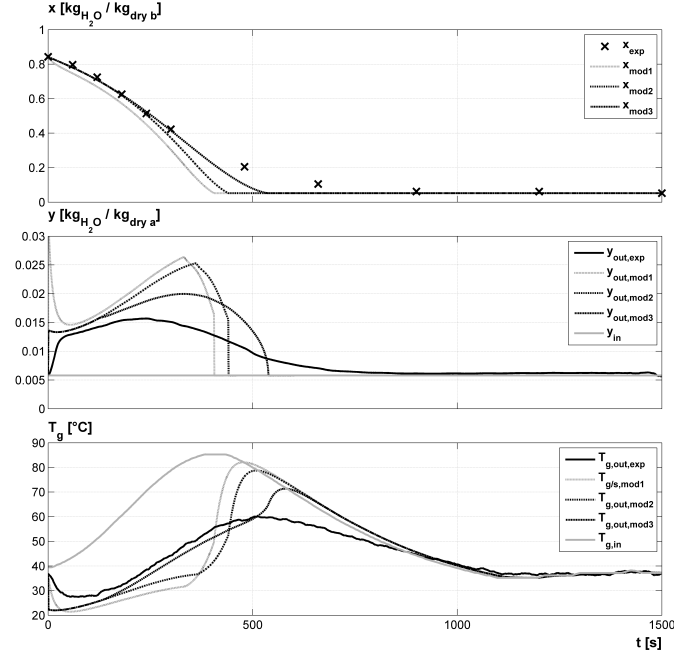
parameters of drying model 3, optimal values for $\frac{\varepsilon_s}{\tau}$ and x_{crit} of respectively $0.55 \text{ m}_{\text{void}}^3/\text{m}_{\text{b}}^3$ and $0.70 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ were found. Drying model 1 does not account for interfacial heat transfer limitations and as a result overestimates the drying rate already during the constant drying rate period. Drying model 2 which accounts for interfacial mass and heat transfer limitations improves the prediction of the drying rate down to the critical moisture content. An accurate prediction of the drying rate below that moisture content clearly requires accounting for intra-particle diffusion limitations. Using the above described shrinking core approach when the moisture content is below x_{crit} (drying model 3) significantly improves the prediction of the drying rate in the falling drying rate period.

Parity plots obtained with the three models are shown in Figures 5.9, 5.10 and 5.11, comparing as well the biomass moisture content (Figure 5.9), the absolute humidity (Figure 5.10) and temperature (Figure 5.11) of the air leaving the drying chamber. In these plots, simulations of the continuous drying experiments of Eliaers and De Wilde [100] are included. The latter are also shown separately in Figure 5.8. An overall good agreement is found when using drying model 3 with the model parameters derived from the batch experiments and given above.

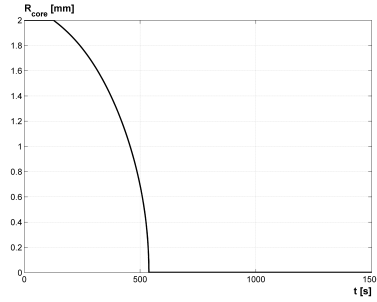
5.5 Simulation of an industrial scale vortex chamber for continuous biomass drying

5.5.1 Vortex chamber designs

Continuous operation is considered. To reach a typical commercial capacity of 5 tons/h humid biomass while keeping similar residence times as those in Eliaers and De Wilde [100], the volume of the vortex chamber has to be multiplied by 250. As for the gas flow rate, a similar air to biomass ratio as in [100] would require $175000 \text{ Nm}^3/\text{h}$. However, since the air utilization was not optimal, that is, saturation was not reached, operation at lower air flow rate can be considered as long as high-G operation is ensured. With a standard vortex chamber (Design A, Figure 5.12), a minimum flow rate of $12000 \text{ Nm}^3/\text{h}$, leading to operation at about 7G, is required. To further improve the performance, a higher gas inlet temperature can also be considered, but the biomass temperature has to remain sufficiently low to avoid its degradation. Because of the



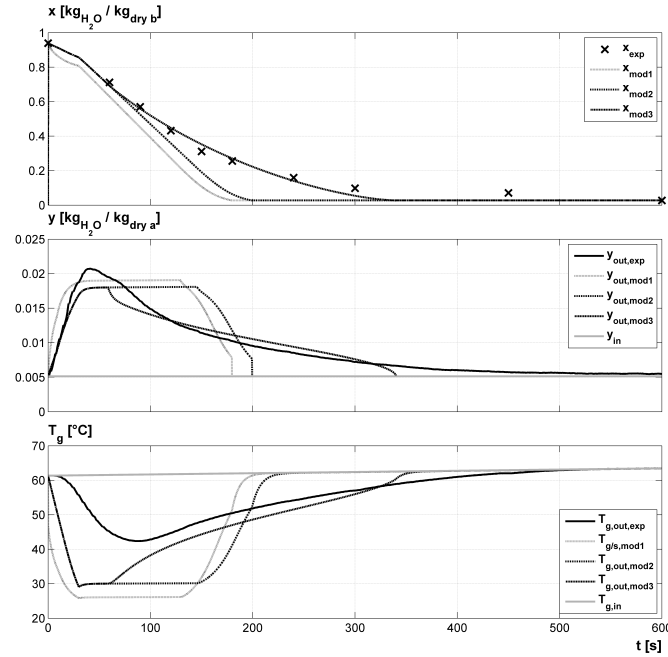
(a)



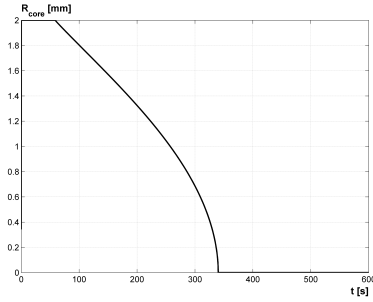
(b)

Figure 5.5: Batch biomass drying in a conventional fluidized bed. (a) Experimentally measured versus simulated data using model 1, 2 and 3 (Figure 5.4). (b) Evolution of the shrinking core radius with time (model 3). Chamber design and operating conditions: see Table 5.1, Exp. 2. Mass and heat transfer coefficients calculated with the Dwivedi/Gupta correlations. $\frac{\varepsilon_s}{\tau} = 0.55$ and $x_{\text{crit}} = 0.70 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ in model 3.

5.5. Simulation of an industrial scale vortex chamber for continuous biomass drying

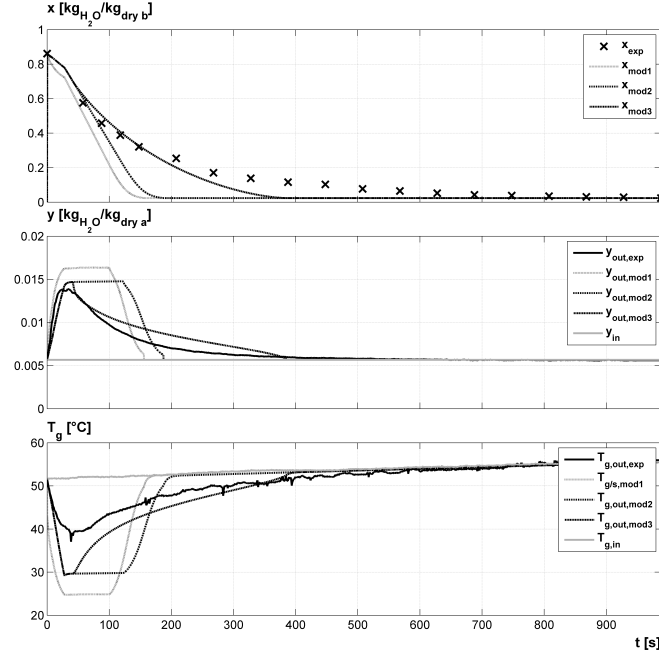


(a)

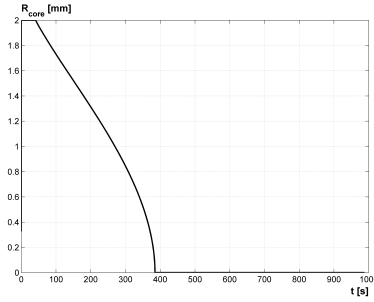


(b)

Figure 5.6: Batch biomass drying in a rotating fluidized bed in a vortex chamber. (a) Experimentally measured versus simulated data using model 1, 2 and 3 (Figure 5.4). (b) Evolution of the shrinking core radius with time (model 3). Chamber design and operating conditions: see Table 5.1, Exp. 5. Mass and heat transfer coefficients calculated with the Yoshida/Handley correlations. $\frac{\varepsilon_s}{\tau} = 0.55$ and $x_{\text{crit}} = 0.70 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ in model 3.



(a)



(b)

Figure 5.7: Batch biomass drying in a rotating fluidized bed in a vortex chamber. (a) Experimentally measured versus simulated data using model 1, 2 and 3 (Figure 5.4). (b) Evolution of the shrinking core radius with time (model 3). Chamber design and operating conditions: see Table 5.1, Exp. 7a. Mass and heat transfer coefficients calculated with the Yoshida/Handley correlations. $\frac{\varepsilon_s}{\tau} = 0.55$ and $x_{\text{crit}} = 0.70 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ in model 3.

5.5. Simulation of an industrial scale vortex chamber for continuous biomass drying

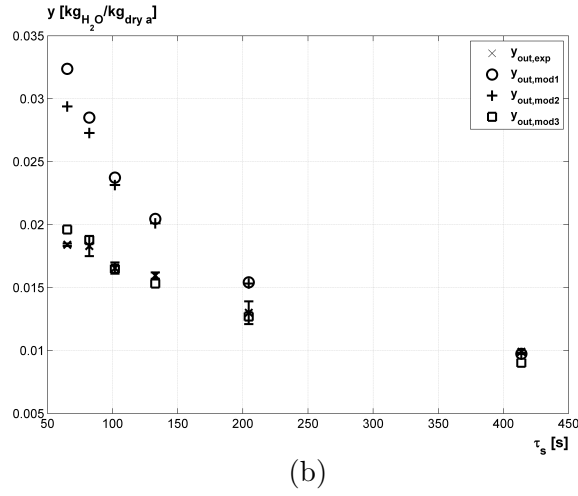
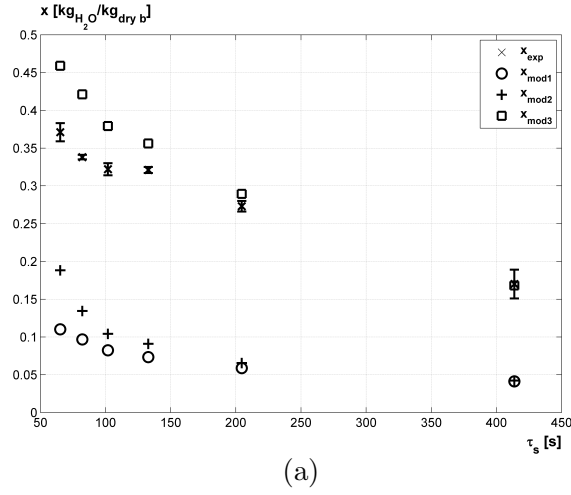
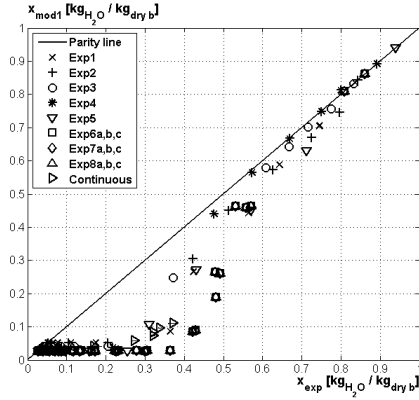
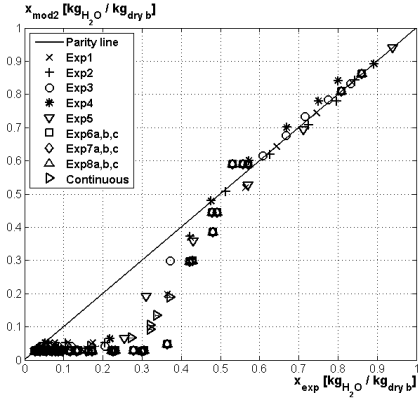


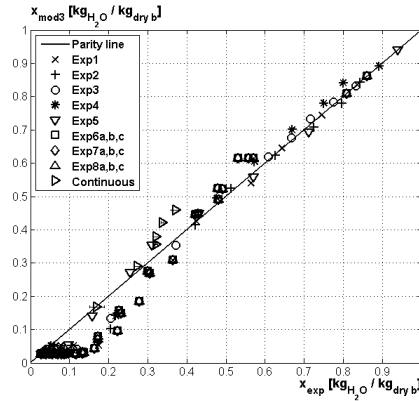
Figure 5.8: Continuous biomass drying in a rotating fluidized bed in a vortex chamber. Experimentally measured [100] versus simulated data using model 1, 2 and 3 (Figure 5.4). Chamber design and operating conditions: see [100]. Mass and heat transfer coefficients calculated with the Yoshida/Handley correlations. $\frac{\varepsilon_s}{\tau} = 0.55$ and $x_{crit} = 0.70$ $\text{kgH}_2\text{O/kg}_{dry\ b}$ in model 3.



Model 1



Model 2



Model 3

Figure 5.9: Simulation of biomass drying in a conventional fluidized bed and in rotating fluidized beds in vortex chambers. Batch drying (Table 5.1) and continuous drying experiments of Eliaers and De Wilde [100] simulated. Parity plots of the biomass moisture content for models 1, 2 and 3 (Figure 5.4). $\frac{\varepsilon_s}{\tau} = 0.55$ and $x_{crit} = 0.70$ kg_{H₂O}/kg_{dry b} in model 3.

5.5. Simulation of an industrial scale vortex chamber for continuous biomass drying

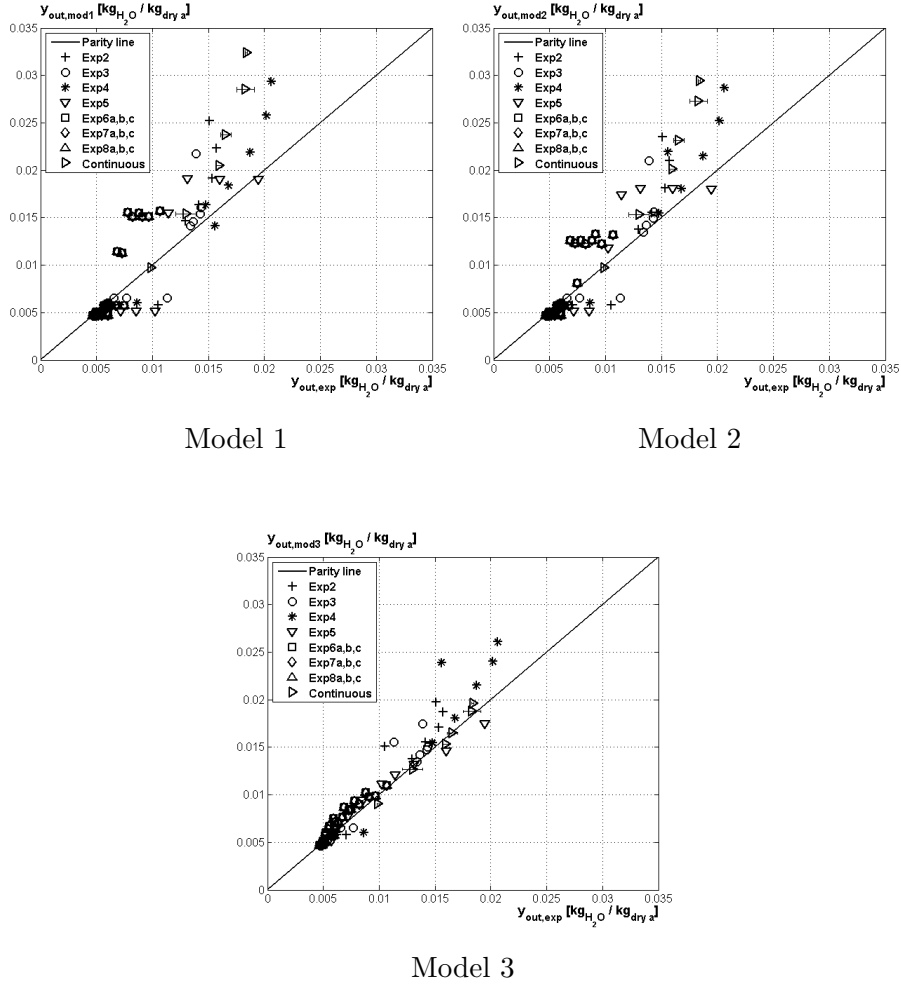


Figure 5.10: Simulation of biomass drying in a conventional fluidized bed and in rotating fluidized beds in vortex chambers. Batch drying (Table 5.1) and continuous drying experiments of Eliaers and De Wilde [100] simulated. Parity plots of the absolute air humidity at the outlet for models 1, 2 and 3 (Figure 5.4). $\frac{\varepsilon_s}{\tau} = 0.55$ and $x_{crit} = 0.70 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ in model 3.

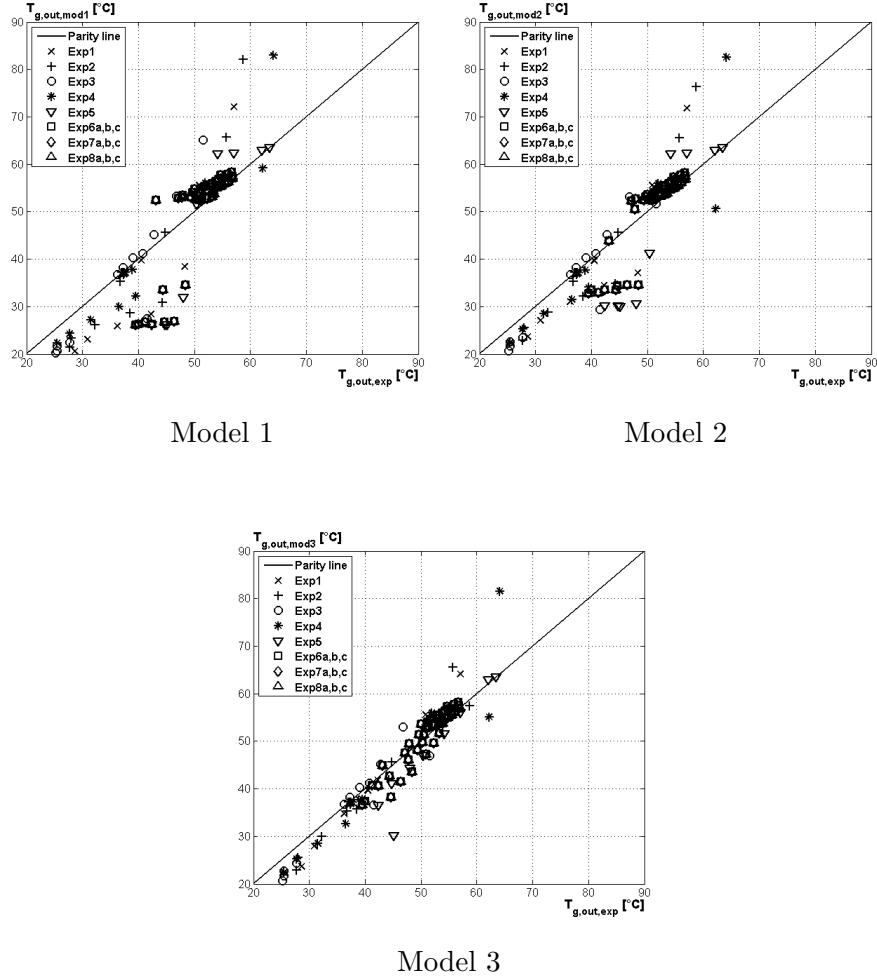


Figure 5.11: Simulation of biomass drying in a conventional fluidized bed and in rotating fluidized beds in vortex chambers. Batch drying (Table 5.1) and continuous drying experiments of Eliaers and De Wilde [100] simulated. Parity plots of the air temperature at the outlet for models 1, 2 and 3 (Figure 5.4). $\frac{\varepsilon_s}{\tau} = 0.55$ and $x_{crit} = 0.70$ kg_{H₂O}/kg_{dry b} in model 3.

5.5. Simulation of an industrial scale vortex chamber for continuous biomass drying

good particle bed mixing, the product uniformity is also an issue [100]. To address the air consumption/utilization and the product uniformity, different designs were considered, as shown in Figure 5.12. The vortex chamber dimensions and operating conditions are summarized in Table 5.2. Design A is a simple vortex chamber. In Design B, the vortex chamber was compartmented, injecting fresh air in all compartments. Practically, rings can be used to compartment the vortex chamber. The biomass moves from one compartment to another, introducing a well-defined axial motion and leading to a CSTR-in series type behavior for the biomass. This allows significantly improving the product uniformity, i.e. the uniformity of the moisture content of the biomass particles leaving the last compartment. Design C/C_{bis} considers different (smaller) vortex chambers in parallel and re-utilizing the air leaving a given vortex chamber in the next one. The biomass is distributed over the different vortex chambers. Whereas the product uniformity will not be improved, the air consumption can be significantly reduced and the air utilization optimized. The D- and E-Designs combine the advantages of vortex chamber compartmenting and air re-utilization. Co-current operation (Design D/D_{bis}) and counter-current operation are then possible. The latter requires two compressors (Design E1/E1_{bis}) or a suction compressor (Design E2/E2_{bis}), drastically increasing the cost of the unit. Therefore, these designs were not further considered in the context of biomass drying. The bis-designs include air reheating in between vortex chambers (Design C_{bis}) or vortex chamber compartments (Design D_{bis} or E_{bis}) to supply the required energy. Note that only vortex chambers or vortex chamber compartments of equal size have been considered in this work.

5.5.2 Results and discussion

Figure 5.13 shows the biomass moisture content and temperature and the air relative and absolute humidity at the outlet as a function of the air inlet temperature for the different air flow rates and biomass feeding rates considered using a Design A (Figure 5.12 and Table 5.2) vortex chamber. For biomass feeding rates up to 5 tons/h, a low biomass moisture content can be achieved. A high air flow rate and air inlet temperature are, however, required. The temperature of the biomass has to be maintained below 80 °C to avoid degradation. This limits the air inlet temperature, as seen from Figure 5.13b. Only at the lower biomass feeding rate (2.5 tons/h), a risk of overheating the biomass is observed

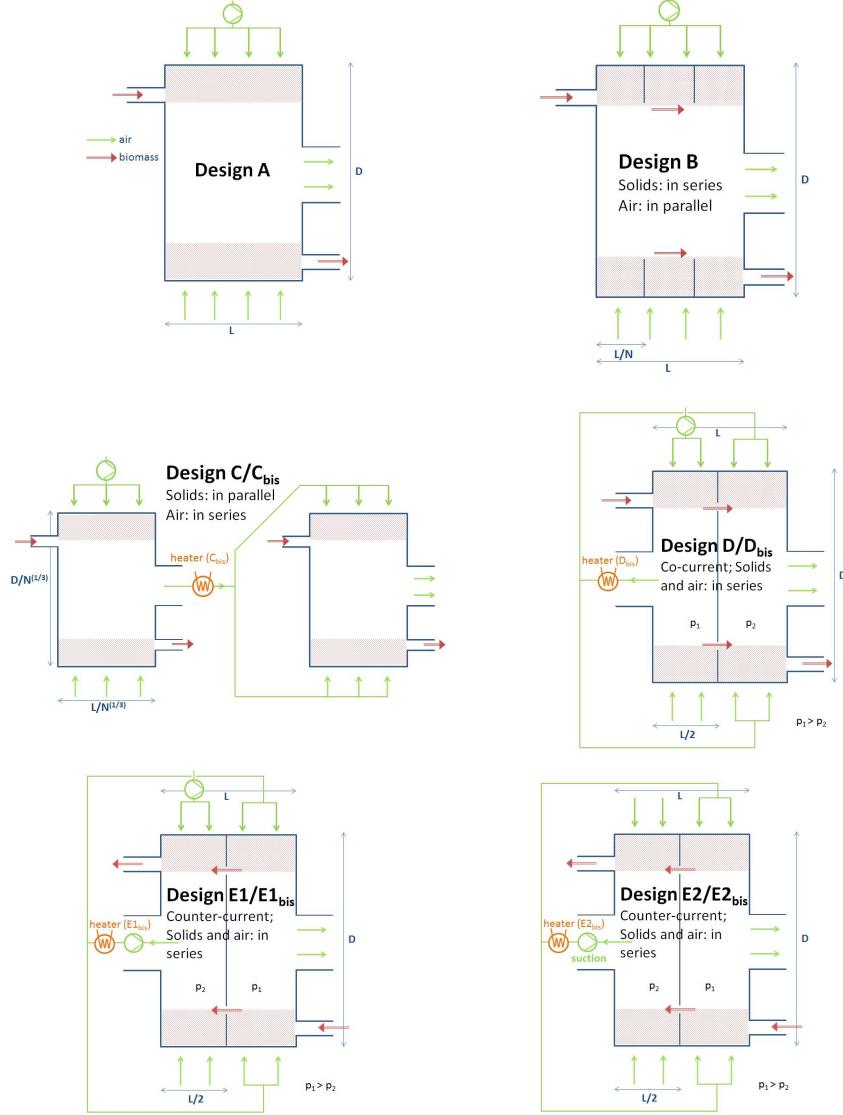


Figure 5.12: Different designs of a vortex chamber biomass dryer considered for the scale-up study. Design A: Single chamber; Design B: Compartmented chamber (operation in series for the biomass and in parallel for the air); Design C/C_{bis}: Multiple chambers in parallel (operation in parallel for the biomass and in series for the air); Design D/D_{bis}/E1/E1_{bis}/E2/E2_{bis}: Chamber with two compartments and dual-chimney design (operation in series for the solids and the air: $p_1 > p_2$), Design D/D_{bis}: Co-current, Design E1/E1_{bis} and E2/E2_{bis}: Counter-current with two compressors or a suction compressor. The bis-designs are equipped with air reheating in between chambers/compartments.

5.5. Simulation of an industrial scale vortex chamber for continuous biomass drying

All designs:	
Vortex chamber(s) total volume [m ³]	1.73
Biomass particle characteristics	Pelletized beech wood, cylindrically shaped, $d_p = 4$ mm, $h_p = 4$ mm
Biomass feed temperature [°C]	20
Biomass feed moisture content [kgH ₂ O/kg _{dry} b]	0.85
Air inlet temperature [°C]	50, 100, 150, 200, 250, 300 (bis designs: air re-heated between compartments/chambers)
Air inlet humidity (absolute) [kgH ₂ O/kg _{dry} a]	$5 \cdot 10^{-3}$
Design A: Single-chimney vortex chamber:	
Vortex chamber dimensions	D = 2.7 m, D _{chimney} = 0.63 m, L = 0.32 m, V = 1.73 m ³
Biomass feed rate (humid) [tons/h]	2.5, 5, 10
Air flow rate [Nm ³ /h]	12000, 20000
p _{out} [Pa]	101325
Design B: Compartmented single-chimney vortex chamber with N compartments: operation in series for the solids / in parallel for the gas:	
Vortex chamber dimensions	D = 2.7 m, D _{chimney} = 0.63 m, L = 0.32 m, V = 1.73 m ³
Number of compartments N	1 (Design A), 2, 3, 5
Compartment dimensions	D = 2.7 m, L = (0.32/N) m, V = (1.73/N) m ³
Biomass feed rate (humid) [tons/h]	5
Total air flow rate [Nm ³ /h]	12000, 20000
Air flow rate per compartment [Nm ³ /h]	12000/N, 20000/N
p _{out} [Pa]	101325
Design C/C_{bis}: N single-chimney vortex chambers: operation in parallel for the solids / in series for the gas:	
Vortex chambers dimensions	D = (2.7/N ^{1/3}) m, D _{chimney} = (0.63/N ^{1/3}) m, L = (0.32/N ^{1/3}) m, V = (1.73/N) m ³
Number of vortex chambers N	1 (Design A), 2, 3, 5
Biomass feed rate per vortex chamber (humid) [tons/h]	5/N
Total air flow rate = air flow rate per chamber [Nm ³ /h]	12000/N, 20000/N
p _{out} [Pa]	101325
Design D/D_{bis}: Dual-chimney vortex chamber with 2 compartments: operation in series for gas and solids and co-current:	
Vortex chamber dimensions	D = 2.7 m, D _{chimney} , L = 0.32 m, V = 1.73 m ³
Number of compartments N	2
Compartment dimensions	D = 2.7 m, L = (0.32/2) m, V = (1.73/2) m ³
Biomass feed rate (humid) [tons/h]	5
Total air flow rate [Nm ³ /h]	12000, 20000
Air flow rate per compartment [Nm ³ /h]	12000/2, 20000/2
p _{out} compartment 1 [Pa]	1.1×101325
p _{out} compartment 2 [Pa]	101325

Table 5.2: Vortex chamber designs and operating conditions considered for the scale-up study.

in the range of operating conditions studied. Figure 5.13 shows that 2.5 tons/h of biomass can be safely dried to less than $0.1 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ feeding $12000 \text{ Nm}^3/\text{h}$ air at 250°C , 5 tons/h feeding $20000 \text{ Nm}^3/\text{h}$ at 300°C . Under such conditions, the relative humidity of the air leaving the vortex chamber is relatively low (Figure 5.13c), about 28% and 16% in the two above-mentioned cases, although its absolute humidity is relatively high (Figure 5.13d). Optimal air utilization, while guaranteeing a biomass moisture content below $0.5 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ is not possible using Design A. As mentioned before, the well-mixed particle bed behavior results in significant product non-uniformity.

As theoretically illustrated in Figure 5.14, compartmenting the vortex chamber and introducing a CSTR-in-series behavior for the biomass allows improving the product uniformity. For the conditions given in Table 5.2, the performance of a Design B (Figure 5.12 and Table 5.2) compartmented vortex chamber with fresh air distributed in parallel over the different compartments is shown in Figure 5.15. A 5 tons/h biomass dryer was focused on. Profiles of the biomass (a) moisture content and (b) temperature and of the (c) relative humidity of the air exiting are shown as a function of the air inlet temperature for a vortex chamber with 1 ($\equiv$ Design A), 2, 3 and 5 compartments and for two different total air flow rates. Profiles for the first (-1) and last (-N) compartment are shown. Fresh air being fed to each compartment, the differences between the profiles in the first and last compartment become the more pronounced with increasing number of compartments and with increasing air flow rate. As the number of compartments, N , increases, the moisture content of the biomass leaving the first compartment logically increases (Figure 5.15a-1). The total vortex chamber volume is kept constant and, hence, the size of each compartment and the related biomass residence time per compartment decreases with increasing N . The evaporation rate in the first compartment is to a certain extent negatively affected by the low temperature at which the biomass is fed (20°C). The moisture content of the biomass leaving the last compartment decreases slightly with increasing number of compartments (Figure 5.15a-N). This is due to intra-particle diffusion limitations only being faced in the last compartment(s). Note that only the average moisture content is shown in Figure 5.15a, not the improved product uniformity which follows from the CSTR-in-series behavior. For the design(s) and conditions considered, reducing the biomass content below $0.3 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ requires an

5.5. Simulation of an industrial scale vortex chamber for continuous biomass drying

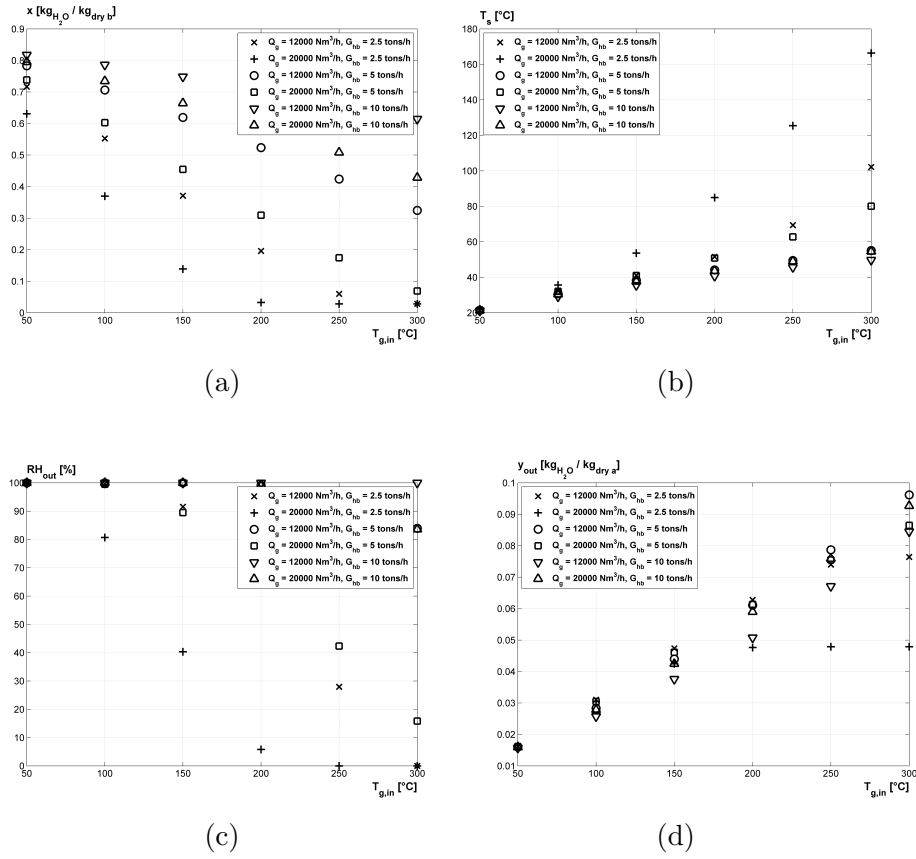


Figure 5.13: Simulations of commercial-scale biomass drying in a Design A vortex chamber for different air flow rates and biomass feeding rates (design and other operating conditions: Table 5.2 and Figure 5.12). Biomass: (a) moisture content (dry basis) and (b) temperature; Air: (c) relative and (d) absolute humidity at the outlet as a function of the gas inlet temperature.

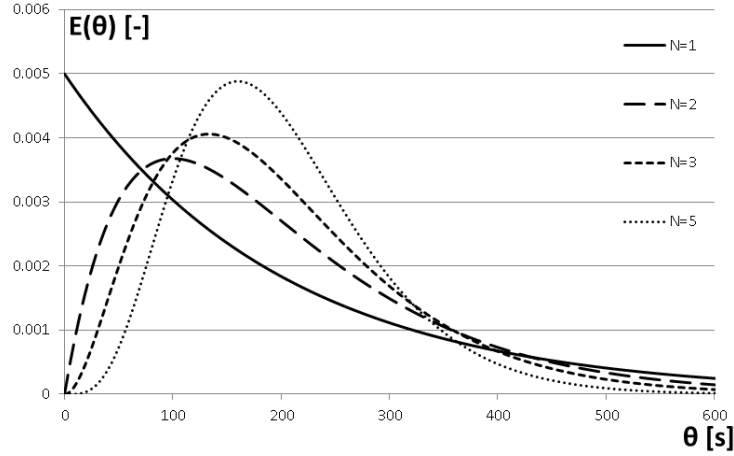


Figure 5.14: Theoretical impact of compartmenting the vortex chamber on the biomass residence time distribution and related moisture content distribution. N: number of compartments (equal size).

air flow rate higher than $12000 \text{ Nm}^3/\text{h}$. When compartmenting, the biomass temperature can be kept easily below 80°C in the first compartment(s) where the biomass is the wettest (Figure 5.15b-1). In the last compartment(s), however, the risk of overheating the biomass increases (Figure 5.15b-N), the biomass in those compartment(s) already being relatively dry. A similar behavior is observed in the air utilization (Figure 5.15c). With increasing number of compartments, the air is better utilized in the first compartment(s), but less good in the last compartment(s). The biomass overheating and air utilization problems in the last compartment(s) are not observed at the lower air flow rate, the biomass then being still relatively wet in the last compartment. When operating at the higher air flow rate, working with a lower air inlet temperature in the last compartment(s), where the biomass is already relatively dry and intra-particle diffusion limitations are the strongest, could resolve the problem of biomass overheating and improve the air utilization.

Figure 5.16 shows the performance of the Design C_{bis} vortex chambers (Figure 5.12 and Table 5.2) in which air reutilization, that is, operation in series for the air, with air reheating in different vortex chambers was considered. Simulations without air reheating (Design C - see Figure

5.5. Simulation of an industrial scale vortex chamber for continuous biomass drying

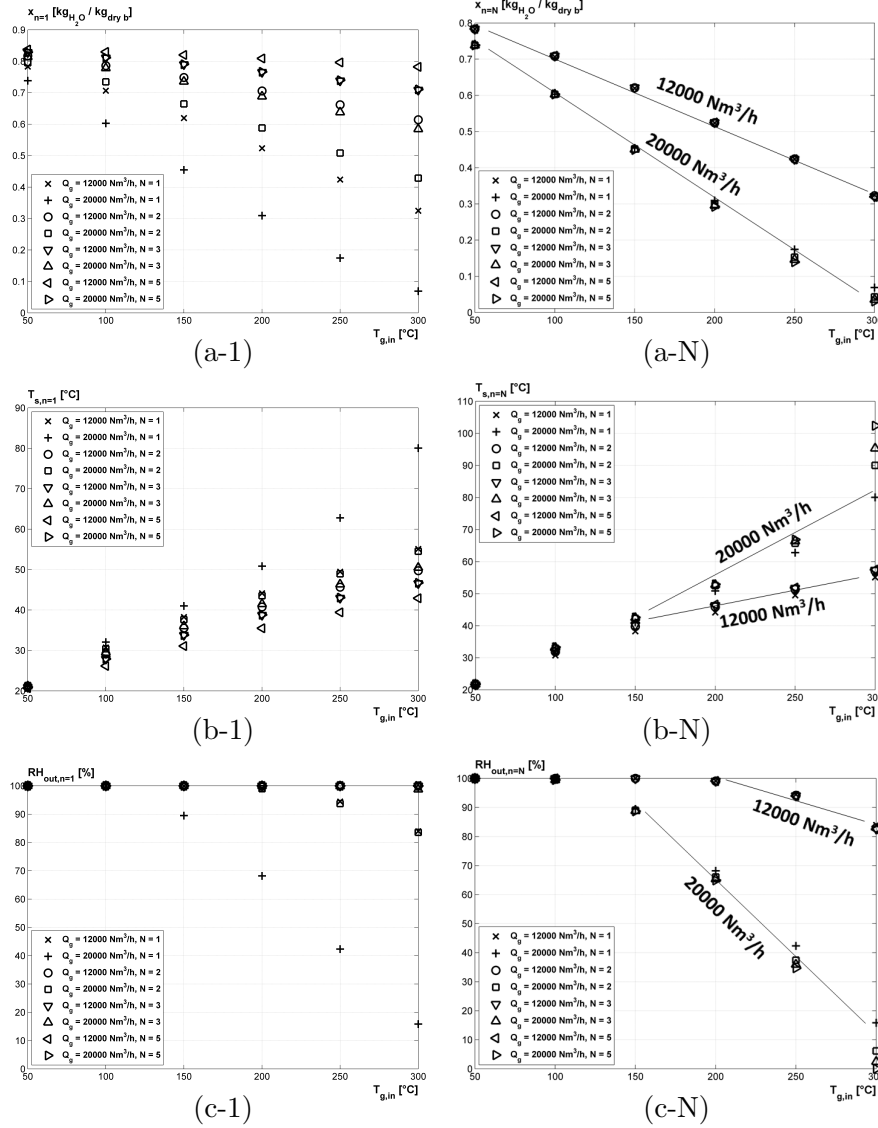


Figure 5.15: Simulations of commercial-scale biomass drying in a Design B vortex chamber with N equally sized compartments for different total air flow rates (design and other operating conditions: Table 5.2 and Figure 5.12). Biomass: (a) moisture content (dry basis) and (b) temperature; Air: (c) relative humidity at the outlet as a function of the gas inlet temperature. (...-1) indicates for the first compartment, (...-N) for the last. Simulations for 1, 3 and 5 compartments.

B.1, annex B) showed very inefficient drying from the second chamber on as a result of insufficient energy supply and air saturation. For 1 ($\equiv$ Design A), 2, 3 and 5 vortex chambers operated in parallel for the biomass, profiles of the biomass (a) moisture content and (b) temperature and of the (c) relative humidity of the air exiting are shown as a function of the air inlet temperature. Profiles in the first (...-1) and last (...-N) vortex chamber are shown. A total biomass feeding rate of 5 tons/h was focused on. Air flow rates of $(12000/N)$ and $(20000/N)$ Nm^3/h were simulated, with N the number of vortex chambers. The volume of each equally sized vortex chamber being inversely proportional with N, high-G operation can still be guaranteed at the lower gas flow rates. The ratio air-to-biomass fed to each vortex chamber is independent of N, explaining the identical profiles for different N in the first chamber (Figure 5.16a-1, b-1 and c-1) in which fresh air is fed. The humidity of the reutilized air increasing gradually from the first to the last vortex chamber, the biomass moisture content achieved in the last vortex chamber increases with increasing number of vortex chambers (Figure 5.16a-N), despite the air reheating. Furthermore, drying in the last chamber becoming slower, an increase of the biomass temperature is observed (Figure 5.16b-N). With 5 vortex chambers, a biomass moisture below $0.1 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ in the first vortex chamber and slightly above $0.1 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$ in the last vortex chamber can be achieved when feeding $4000 \text{ Nm}^3/\text{h}$ air at 300°C , but the biomass temperature in the last vortex chamber reaches values around 85°C . Feeding air at lower temperature resolves the biomass overheating problem, but results in somewhat higher biomass moisture contents. Optimizing the air inlet temperature for each vortex chamber can be considered. Figure 5.16c confirms that air reutilization in different vortex chambers allows significantly reducing the air consumption and improving the air utilization. Feeding air at 300°C and aiming at a low biomass moisture content of around $0.1 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{dry b}}$, for example, using Design C_{bis} with 5 vortex chambers allows reducing the air flow rate from 20000 to $4000 \text{ Nm}^3/\text{h}$ and increasing the relative humidity of the air leaving the system from 16% to 46% (Figure 5.16c-N). The product uniformity cannot be improved using Design C/ C_{bis} , fresh biomass being fed to each vortex chamber and air being reutilized. Figure 5.16a shows that the difference in the moisture content of the biomass leaving the first and the last vortex chamber logically increases with the number of vortex chambers and with decreasing gas flow rate, but remains limited. In the range studied, the influence of the gas inlet temperature

5.5. Simulation of an industrial scale vortex chamber for continuous biomass drying

on the product uniformity is less pronounced.

The performance with Design D_{bis} (Figure 5.12 and Table 5.2) with 2 equally sized compartments and co-current operation in series for the biomass and the reheated air is shown in Figure 5.17, focusing on a capacity of 5 tons/h of biomass. Efficient drying in the second compartment could not be achieved without air reheating (Design D - see Figure B.2, annex B). With two compartments, a dual-chimney design can be adopted, facilitating the practical application. Profiles in the first (...-1) and second/last (...-2) compartment of the biomass (a) moisture content and (b) temperature and of the (c) relative humidity of the air exiting are shown as a function of the air temperature for two different air flow rates. For comparison, the profiles with Design A (1 compartment) are included. Using Design D_{bis}, a combination of improved product uniformity and reduced air consumption/improved air utilization is aimed at. The net effect of compartmenting and air re-utilization on the moisture content of the biomass leaving the system is limited. The influence on the biomass temperature and on the air utilization, i.e. relative humidity, is much more pronounced. Because of the co-current operation, already partially wetted biomass and humidified air enter the second chamber, resulting in lower drying rates and an increased risk of biomass overheating. Intra-particle diffusion limitations contribute to the lower drying rates in the second chamber. The lower drying rates in the second chamber also explain the relatively modest improvement in the air utilization that can be achieved by using Design D_{bis} (Figure 5.17c-2). The reduction of the air flow rate by a factor 2 and the improvement of the product uniformity should, however, be kept in mind. Note that the biomass temperature and drying rate in the first compartment are somewhat lowered by the low biomass feeding temperature (20 °C). Figure 5.17 shows that with an air flow rate of 10000 Nm³/h and air inlet temperature of 250 °C, the moisture content of the biomass leaving the system can be kept below 0.2 kg_{H₂O}/kg_{dry b}, while maintaining the biomass temperature below 80 °C. The relative humidity of the air exiting the system is 52%, about 10% higher than when using a simple chamber (Design A). With co-current operation, going to a lower biomass moisture content is challenging. Biomass overheating and poor air utilization in the last compartment have to be simultaneously addressed, eventually by applying different air inlet temperatures in the different compartments. Counter-current operation for

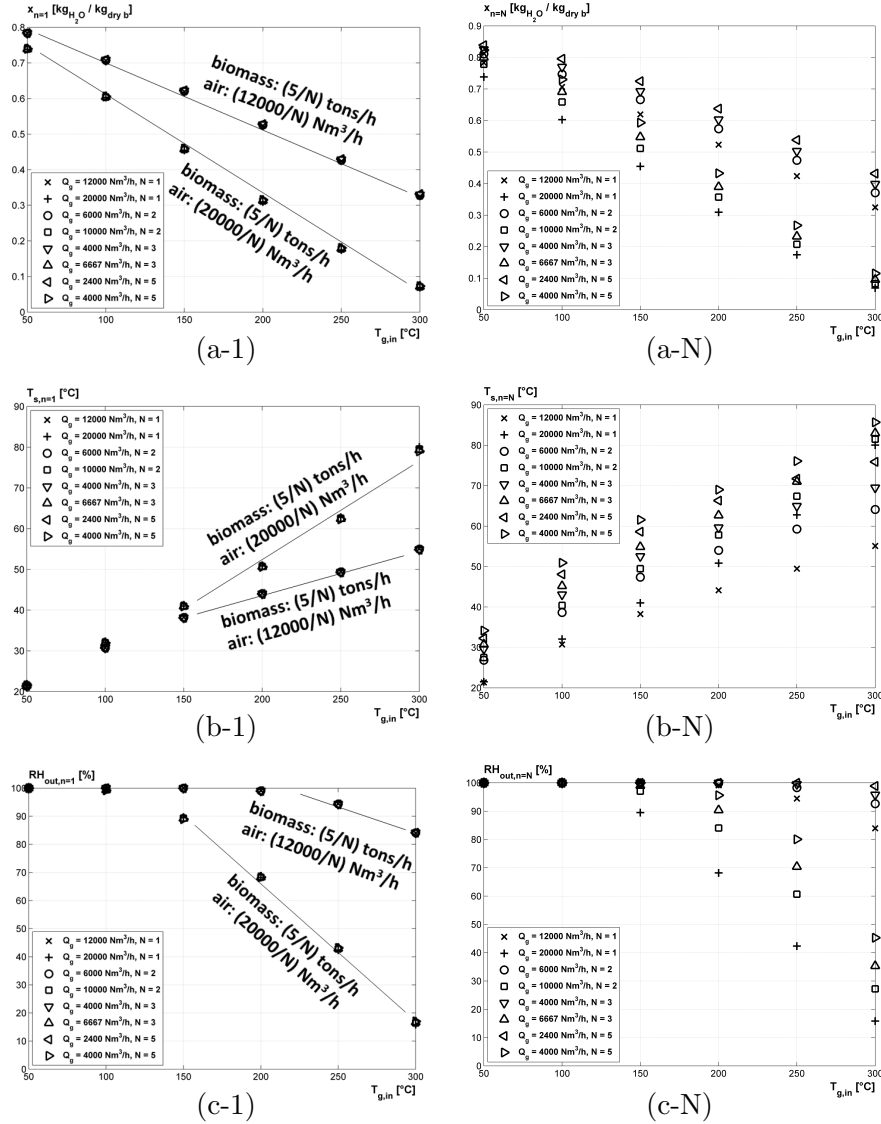


Figure 5.16: Simulations of commercial-scale biomass drying for Design C_{bis} with N equally sized parallel vortex chambers for different total air flow rates (design and other operating conditions: Table 5.2 and Figure 5.12). Biomass: (a) moisture content (dry basis) and (b) temperature; Air: (c) relative humidity at the outlet as a function of the gas inlet temperature. (...-1) indicates for the first chamber, (...-N) for the last. Simulations for 1, 3 and 5 vortex chambers.

the biomass and the air (Design E/E_{bis}, Figure 5.12) is possible, but was considered too complex for the biomass application studied in this work, as already mentioned.

The performance of the different designs in terms of air consumption and biomass moisture content and uniformity is directly compared in Figure 5.18. A significant reduction of the air consumption while maintaining a given biomass moisture content and improving the product uniformity is possible, although care has to be taken not to overheat the biomass.

5.6 Conclusions

Batch drying experiments of granular woody biomass in a vortex chamber confirm the process intensification that can be achieved as earlier observed in continuous drying experiments. Simulating both the batch and continuous experiments shows that the drying model has to account for interfacial mass and heat transfer limitations and for a non-uniform distribution of the moisture in the particles when the moisture content is below a critical value. A shrinking core model describes relatively well the observed behavior, provided that intra-particle diffusion limitations in the dry part of the particles are included and packed bed type correlations for the coefficients of interfacial mass and heat transfer are used. The latter is explained by the high bed density and improved bed uniformity obtained in rotating fluidized beds in vortex chambers. The two model parameters introduced in respectively the interfacial mass and heat transfer correlations for the RFB-SGs and the shrinking core drying model were determined from the batch experiment data and as such used for the simulation of previously published continuous drying experiments. Deviations between the experimentally measured and simulated values of the biomass moisture content, air humidity and air temperature are attributed to a slight non-ideality of the flow pattern of the gas and/or solids. In the simulation model, plug flow was assumed for the air and complete mixing for the particle bed, based on the high particle bed rotational motion and previously reported smoke tracer experiments.

Scale-up of the technology is challenging. Focusing on continuous operation, two important aspects were studied with the retained dry-

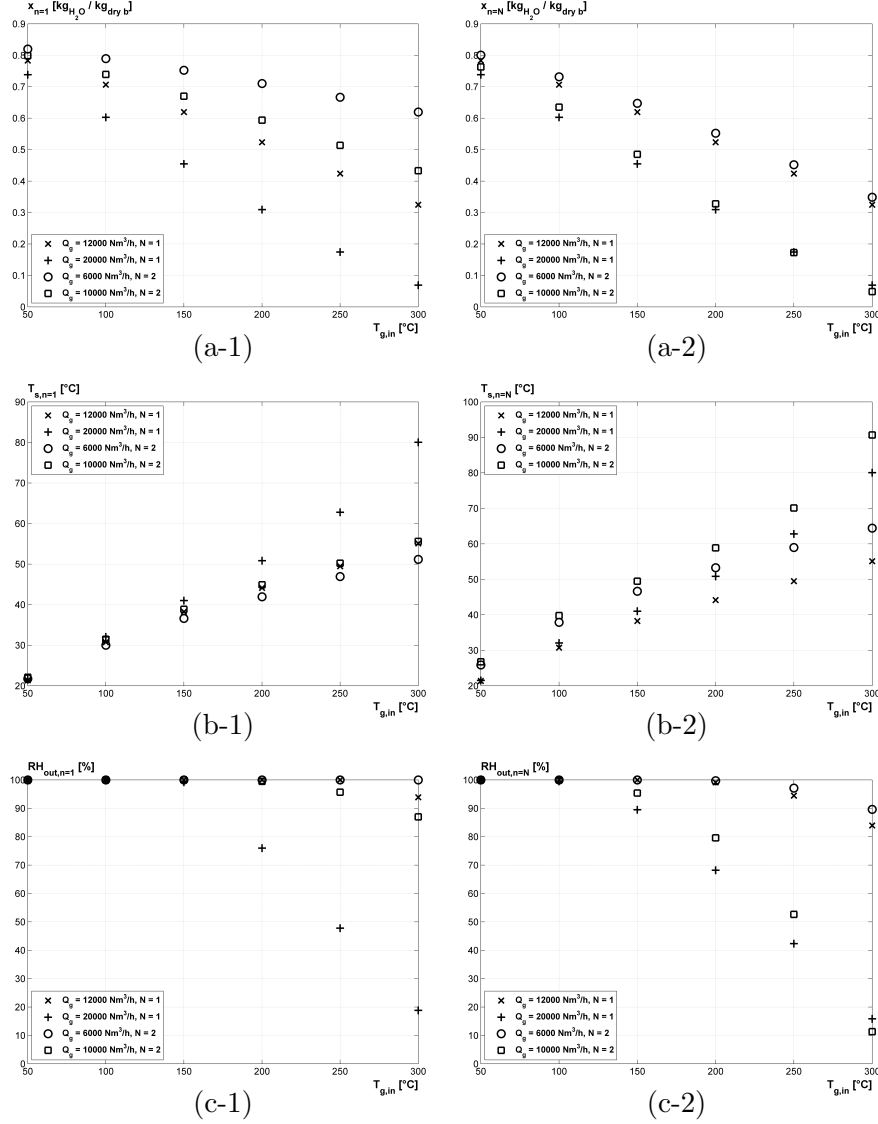


Figure 5.17: Simulations of commercial-scale biomass drying in a Design D_{bis} vortex chamber with 2 equally sized compartments for different total air flow rates (design and other operating conditions: Table 5.2 and Figure 5.12). Biomass: (a) moisture content (dry basis) and (b) temperature; Air: (c) relative humidity at the outlet as a function of the gas inlet temperature. (...-1) indicates for the first compartment, (...-2) for the second. Comparison with one-compartment vortex chamber shown.

5.6. Conclusions

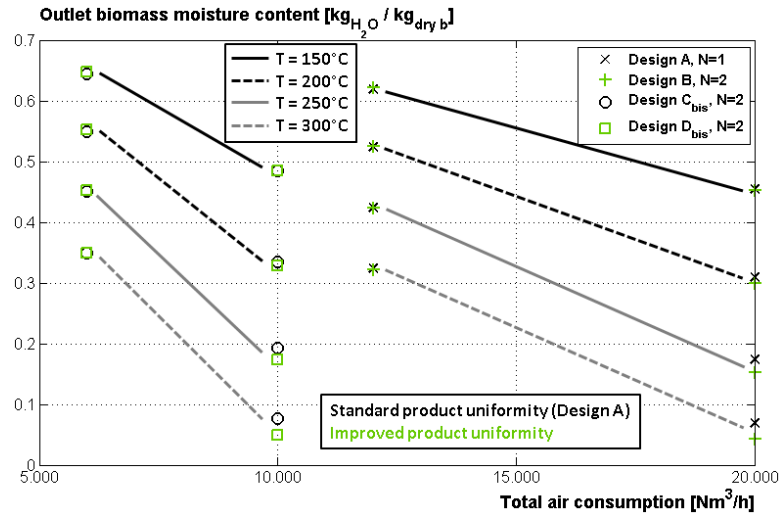


Figure 5.18: Outlet biomass moisture content and uniformity (green symbol indicates improved uniformity) vs. air consumption for a humid biomass feed rate of 5 tons/h and different air inlet temperatures (lines are indicative) using different vortex chamber dryer designs as shown in Figure 5.12. Vortex chamber design and other operating conditions: see Table 5.2 and Figure 5.12.

ing model: (i) the product uniformity and (ii) the air consumption and related air utilization. When using a single chamber, the well-mixed (CSTR-type) behavior in the fast rotating particle bed results in a non-uniform product. Compartmenting the vortex chambers, e.g. by means of rings, allows introducing a CSTR-in-series type behavior for the particles, significantly improving the product uniformity. The air consumption can be significantly reduced and the air utilization improved by re-using the air leaving one vortex chamber or vortex chamber compartment in another one, i.e. operation in series for the air. A complete removal of particles eventually entrained by the air into a chimney and reheating the air are, however, required. Co- and counter-current operation in series for both biomass and air is possible and allows combining advantages. Counter-current operation was considered technologically too costly for biomass drying. With co-current operation, lower drying rates towards the end of the system imply a risk of overheating the biomass and poor air utilization. Optimization of the air inlet temperature for each vortex chamber or vortex chamber compartment can be considered.

Nomenclature

a_m	$\text{m}_{\text{int}}^2/\text{kg}_{\text{dry b}}$	interfacial area per mass unit of dry biomass
$a_{\text{H}_2\text{O}}$	-	water vapour activity in the solid phase
Bi	-	Biot number
$c_{p,i}$	$\text{J}/(\text{kg}.\text{K})$	specific heat of i
D	m_r	vortex chamber diameter
D_{chimney}	m_r	chimney diameter
$D_{\text{H}_2\text{O},a}$	m^2/s	molecular diffusivity of water vapour in air
D_{eff}	$\text{m}_g^3/(\text{m}_p.\text{s})$	effective diffusivity
d_p	m_p	particle mean diameter
G	$\text{kg}/(\text{m}_r^2.\text{s})$	superficial mass flow velocity
ΔH_{vap}	$\text{J}/\text{kg}_{\text{H}_2\text{O}}$	latent heat of vaporization
h_f	$\text{W}/(\text{m}_{\text{int}}^2.\text{K})$	heat transfer coefficient
j	$\text{kg}/(\text{m}_{\text{int}}^2.\text{s})$	water vapour mass flux at the gas-solid interface
j_D	-	Colburn j-factor for mass transfer
j_H	-	Colburn j-factor for heat transfer
k_g	$\text{mol}/(\text{bar}.\text{m}_{\text{int}}^2.\text{s})$	mass transfer coefficient
k_{shell}	$\text{mol}/(\text{bar}.\text{m}_{\text{int}}^2.\text{s})$	pseudo mass transfer coefficient in the dry shell
L	m_r	vortex chamber length
M_i	kg/mol	molecular weight of i
m_b	$\text{kg}_{\text{dry b}}$	mass of dry biomass in the particle bed
$\dot{m}_b$	$\text{kg}_{\text{dry b}}/\text{s}$	dry biomass feeding rate
N	-	number of vortex chambers or compartments
Nu	-	Nusselt number
p_i	bar	partial pressure of i
$p_{i,\text{sat}}$	bar	saturation partial pressure of i
p_{tot}	bar	total pressure
Pr	-	Prandtl number
r	m_p	spatial coordinate in the biomass particles
R	$\text{J}/(\text{mol}.\text{K})$	ideal gas law constant
R_{core}	m_p	particle core radius
R_p	m_p	particle radius

Re_p	-	particle-based Reynolds number
Re_{pm}	-	modified particle-based Reynolds number
S	m_r^2	surface normal to the gas flux
Sc	-	Schmidt number
Sh	-	Sherwood number
t	s	time
t_{feed}	s	solids feeding time for batch experiments in vortex chambers
T_j	K	temperature of phase j
V	m_r^3	volume
x	$kg_{H_2O}/kg_{dry\ b}$	biomass moisture content
y	$kg_{H_2O}/kg_{dry\ a}$	absolute air humidity
y_i	mol_i/mol_g	molar fraction of i in the gas phase
z	m_r	spatial coordinate along the fluidization chamber

Greek letters

α	-	mass and heat transfer coefficient parameter in the RFB-SG.
ε	m_g^3/m_r^3	particle bed void fraction
ε_s	m_g^3/m_p^3	biomass particles void fraction
λ	-	fraction of the chamber circumference taken by gas inlet slots
λ_i	W/(m.K)	thermal conductivity of i
μ_i	Pa.s	viscosity of i
ρ_i	kg/m^3	density of i
$\tilde{\rho}_a$	$kg_{dry\ a}/m_g^3$	air density on a dry basis
$\tilde{\rho}_{B,dry}$	$kg_{dry\ b}/m_r^3$	bed density on a dry basis
τ	-	biomass particles tortuosity

Subscripts

a	air
b	biomass
B	particle bed
crit	critical
dry	on a dry basis
g	gas phase
H ₂ O	water
hb	humid biomass
int	interface
p	particle
r	reactor
res	residual
s	solid phase

Superscripts

s	solid surface
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Particle coating: state of the art

6.1 Introduction

Particle coating is used for different purposes in the feed and food, pharmaceutical, agricultural, chemical and cosmetic industries [106–110]. A coating can protect the core material from the environment (e.g. humidity, oxygen, light) during storage or allows a controlled release of an active component from the core. A coating may also be used to mask odor or bad taste. The coating material may be chosen to break up under certain conditions of pH, pressure, temperature, solvent, etc. Finally, surface properties can be modified, e.g. to reduce hygroscopicity or to improve the flowability of the granular material.

6.2 Particle coating techniques

6.2.1 Wet Coating processes

Wet coating processes include techniques in which the coating material is fed in the coater as a liquid solution or suspension. The solvent is commonly water but alcohols, cetones, ethers or esters can be used in some cases to increase the coating material content in the solution or to reduce the heat required for its evaporation [111]. These processes generally make use of a spray nozzle that injects the liquid in the form of small droplets. These droplets spread onto the core material surface and liquid bridges are built with the core material particles. The solvent is dried by a flow of hot process gas leaving one or multiple layers of coating material on the core particles (Figure 6.1) [106, 107, 109, 112, 113]. If liquid bonds are formed between the core material particles, agglomeration occurs and granules are formed [106, 107]. Agglomeration is mostly

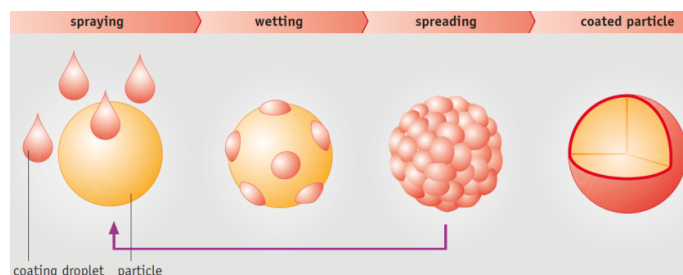


Figure 6.1: Wet coating mechanism [114].

observed when the droplet size is relatively large compared to the size of the particles to be coated [108,110]. Both efficient mixing and circulation of the core material and good contact between the core material particles and the process gas are required. Therefore, wet coating processes are commonly carried out in fluidized beds or rotating drums.

Fluidized bed coaters

Fluidized bed coaters make use of the fluidized bed technology described in chapter 1. In conventional, i.e. gravitational, fluidized bed coaters, core material particles are suspended by upward flowing gas. In this case, the process gas, commonly air, is used both to fluidize the core material and to evaporate the solvent of the coating solution/suspension [112]. The relatively high slip velocities between the solids and the gas and the related high interfacial heat and mass transfer rates allow a relatively low coating process time in fluidized bed compared to other technologies. Moreover, the efficient mixing makes the temperature uniform in the particle bed and lead to efficient heat exchanges with the surrounding.

Depending on the spray nozzle position and orientation in the fluidized bed column, the coater can be categorized as top, bottom or tangentially-sprayed (Figure 6.2). Top-sprayed fluidized bed coaters (Figure 6.2a) offer capacity and an easy access to the nozzle for replacement/cleaning during a run [108,112,113]. However, top-sprayed coaters do not allow controlling the distance the droplets travel before colliding with a core material particle. As a result, a high quality and uniform coating with controlled agglomeration is difficult to achieve. Furthermore, a poor coating efficiency may be obtained because a significant

6.2. Particle coating techniques

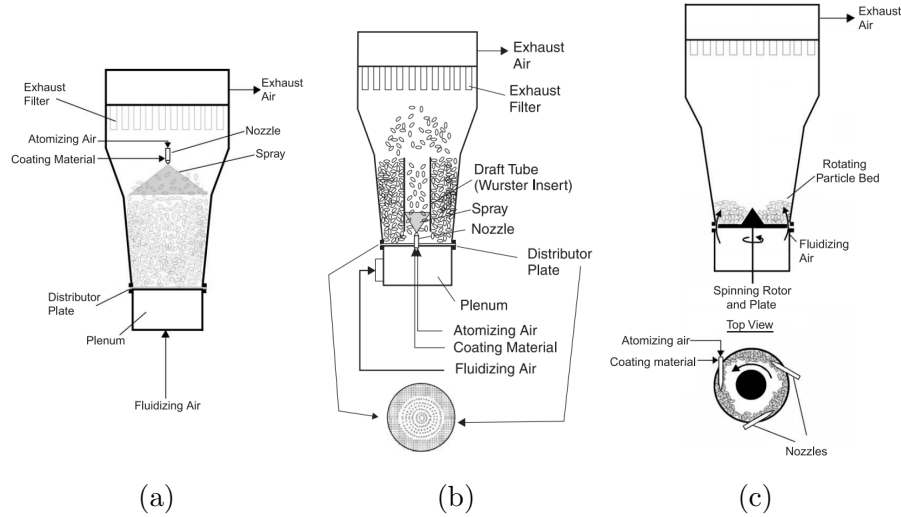


Figure 6.2: Single-cell fluidized bed coater configurations: (a) top-sprayed, (b) bottom-sprayed (Wurster) and (c) tangentially-sprayed fluidized bed coater [112].

number of droplets can be spray-dried [108]. Top-sprayed coaters are consequently better suited for granulation processes [112, 113].

To avoid pronounced agglomeration, bottom-sprayed fluidized bed coaters are generally used with a co-axial draft tube and an air distributor guiding the motion of the particles. As more gas is injected in the center, the particles are flowing upwards in the inner tube and fall back in the outer annular region (spouted bed). Such so-called Wurster coaters (Figure 6.2b) offer a better coating quality because of an intense circulation, good contacts between particles and droplets and a resulting increased evaporation rate. Consequently, more uniform coated and less agglomerated particles are produced compared to top-sprayed coaters [106, 107]. The coating efficiency is also improved as spray-drying can be avoided [107, 108]. The risk of particle erosion is, however, more pronounced because of the high air velocities required for sustaining the particle circulation [112]. Another drawback of Wurster coaters is the inability to remove the spray nozzle during a run if the latter is clogged [112]. Wurster coaters are therefore used when the coating quality and uniformity are crucial whereas top-spray coaters are commonly

used for high production rates.

In tangentially-sprayed coaters (Figure 6.2c), the coating solution or suspension is injected tangentially at the bottom of the fluidized bed. Tangentially-sprayed coaters are also known as rotor or rotary fluidized bed coaters because they are generally used in combination with a rotating disk mixing the particles just above the air distributor [112]. As a result, a coating quality comparable to that of Wurster coaters with improved sphericity and density can be achieved [107, 113]. However, tangential spray coaters cannot be used for friable or crumbly material because of the high agitation created by the rotating disk [107, 113].

Single-cell fluidized bed coaters are usually operated batch-wise. The first attempt to go to fully continuous coaters was performed with a single-cell fluidized bed containing a discharge pipe located in the center of the column (Figure 6.3a). However, single-cell continuous fluidized beds cannot produce uniformly coated particles with controlled agglomeration. In a second type of continuous coater, the horizontal fluidized bed, the solids move horizontally by means of a conveyor belt. Horizontal fluidized bed coaters are generally vibrating and contain multiple top spray nozzles as shown in Figure 6.3b. Nevertheless, the length required to ensure a sufficiently high residence time can be a critical issue, especially for coating. To overcome this problem, multi-cell continuous fluidized bed coaters (Figure 6.3c) have been developed recently. They consist of a horizontal fluidized bed divided in four or five compartments connected by regulating flaps. This allows obtaining higher residence times keeping a reasonable reactor length. Multi-cell fluidized bed coaters are suited for a number of coating applications such as taste masking. The continuous production of high quality coated particles as well as dealing with a large particle size distribution remain, however, challenging [107, 113].

Fluidized bed coaters have been successfully used to coat particles with a diameter down to 100 μm [108]. Processing smaller or light, cohesive particles (Geldart-C type) is challenging, the inter-particle Van der Waals forces becoming too important to allow fluidization. Chen et al. [12, 110, 115] dry-coated the surface of 15 μm cornstarch particles with nano-size particles in order to modify the surface properties and reduce the inter-particle forces. This first step allowed fluidizing these

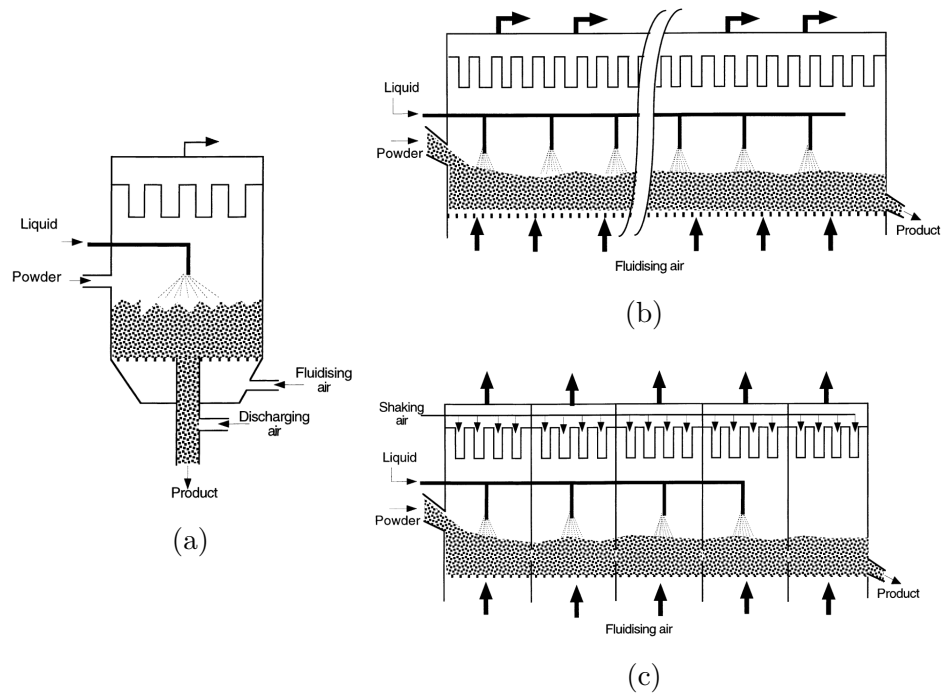


Figure 6.3: Continuous fluidized bed coater configurations: (a) single-cell, (b) horizontal and (c) multi-cell fluidized bed coater [107].

particles [12] and coating [110] or granulating [115] them with polymer films in a top-sprayed fluidized bed coater. The treatment with nano-size particles is, however, costly. Geldart-C particles can also be successfully fluidized in cylindrically shaped rotating fluidized bed (RFB) devices [54, 55]. High-G operation allows increasing the gas-solid slip velocity and intensifying interfacial momentum transfer. The importance of the inter-particle Van der Waals forces is thereby significantly reduced. Interfacial mass and heat transfers are also intensified because of the increased slip velocities [27, 53–55] leading to a faster evaporation of the solvent. Furthermore, denser and more uniform particle beds can be obtained. Rotating fluidized beds (RFBs) can be generated in rotating and static chambers. On the lab-scale, the coating and granulation of cohesive powders in a RFB with a rotating chamber have been successfully studied by Watano et al. [54, 116]. The use of a motor limits, however, commercial application of rotating chambers because of undesired mechanical vibrations and sealing problems [53–55]. The coating of particles in RFBs with a static chamber has not been reported in the literature so far.

Rotating drums (perforated pan) coaters

A rotating drum coater is a perforated cylinder containing a single or multiple spray nozzle that sprays the coating solution/suspension on the core material (Figure 6.4). As the drum is rotating, solids cascade down at the bed surface. As a result the core material particles exposed to the coating solution droplets are renewed. Hot process gas, commonly air, is flowing through the holes of the cylinder to evaporate the solvent. The main advantage of rotating drum coaters is the smooth mixing of the particles that allows the coating of erosion-sensitive materials. Rotating drums are particularly suited for particles that would require a high air velocity to be fluidized. Interfacial mass and heat transfer are, however, less efficient than in fluidized bed coaters. Moreover, a uniform product is difficult to achieve [112].

6.2.2 Solvent-free processes involving a liquid phase

Solvent-free processes have also been developed. They save the energy consumption for the evaporation of the solvent and allow coating solvent-sensitive particles. One way to avoid the use of a solvent is injecting the coating material as a hot melt through a spray nozzle. In this case, the

6.2. Particle coating techniques

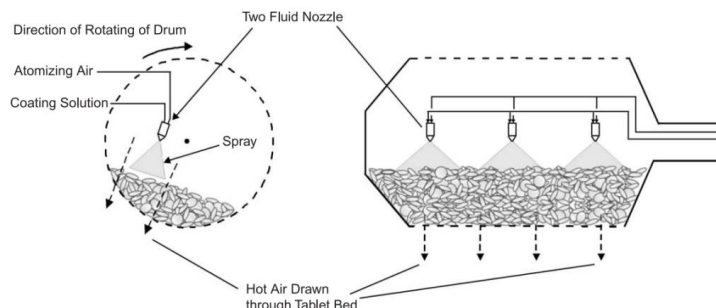


Figure 6.4: Rotating drum coater [112].

coating material is heated at 40 to 60 °C above its melting point and solidifies on the core material surface [111,117]. If a two-fluid nozzle is used, the heating of the atomization air might be required to guarantee that the coating material is molten when fed in the coater. Hot melt coating processes can be carried out in all types of fluidized bed coaters [111] but top-sprayed coaters are commonly used [112]. As no drying is required, the fluidization air doesn't need to be heated [111]. Nevertheless, hot melt coating cannot be applied with heat-sensitive coating or core materials. Moreover, very viscous molten coating material will lead to bad flow properties [111]. Similarly, the use of a liquid precursor that polymerises on the core material surface can be considered. However, its potential is limited for toxicological reasons [117].

Finally, the use of a binder or plasticizer can be considered. In a first step, the binder is sprayed on a mixture of core (host) and coating (guest) particles, leaving a deposit of guest on the host particles. In a second step, this deposit is transformed into a uniform film coating by curing, i.e. heating the particles for a short time [111].

6.2.3 Dry coating processes

Dry coating is another way to avoid the use of solvent or binder. Dry coating refers to the coating of relatively large (1-200 μm) particles (core material or host) with fine submicron particles (guest) by means of an external mechanical force [111,118]. The latter must be strong enough to break the guest particles aggregates and bring them into contact with the host particles. The guest particles being very small, Van Der Waals interactions ensure a strong adhesion on the host particles once in con-

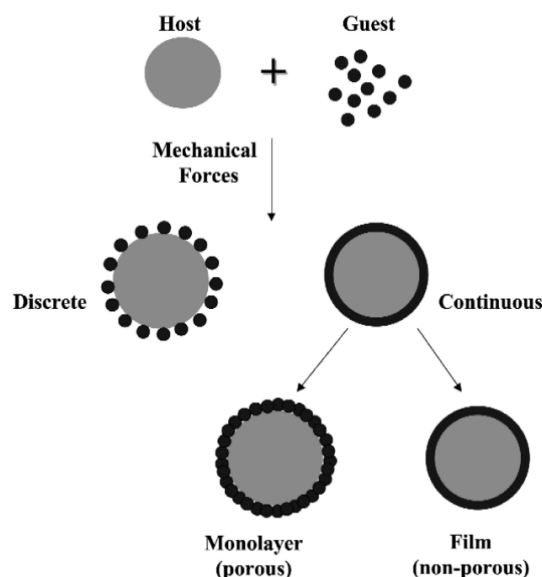


Figure 6.5: Dry coating morphologies [118].

tact with each other. The guest particles are finally randomly dispersed over the host particles because of collision between coated and uncoated host particles [111, 118]. Depending on the materials and operating conditions, discrete or continuous coating can be achieved (Figure 6.5). Continuous coating can consist in a porous pile-up of guest particles layers or in a non-porous film [118].

A number of dry coating devices have been developed. The nature of the mechanical force required for breaking the guest agglomerates and bringing them in contact with the host particles depends on the device. However, dry coating is currently more expensive than wet coating processes [108]. Furthermore, a uniform, good quality coating cannot be achieved with all types of guest and host particles.

Mechanofusion [111, 118]

Mechanofusion was the first dry coating method. The device consists of a rotating cylindrical vessel containing two stationary pieces: a scraper and a hammer (Figure 6.6). As the guest and host particles rotate with the vessel between 200 and 1600 rpm near the outer wall, they are forced

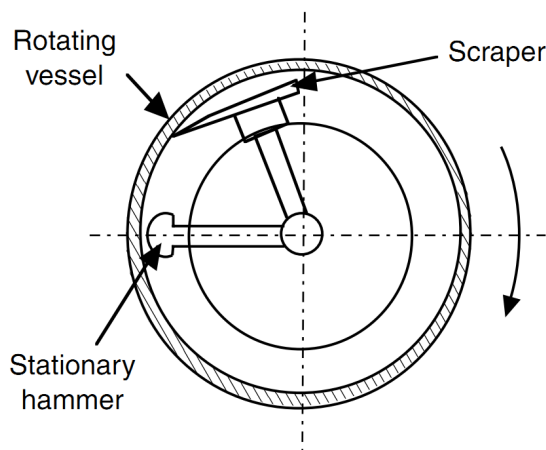


Figure 6.6: Mechanofusion device [111].

to pass through the gap between the stationary pieces and the rotating wall. Consequently, the particles are submitted to strong shearing and compressive forces that allow "fusing" the guest particles onto the surface of the host particles. The gap sizes between the wall and both the hammer and the scraper are important parameters to control the coating thickness. Because the particles undergo strong mechanical forces, mechanofusion can lead, however, to a degradation of the particle size, shape and composition if they are too soft. Moreover, mechanofusion is not suited for heat-sensitive particles as the temperature can rise to 60 °C during the process.

Dry impact blending or hybridization [111,118]

The hybridizer also generates high mechanical forces by means of a high speed rotating rotor. The blades located on the rotor force strong impact and dispersion of the particles (Figure 6.7). Electrostatic charges created by collisions between particles facilitate the attachment of the guest particles on the host particles. Sometimes, dry impact blending is referred to as mechanochemical treatment because chemical interactions can occur between the guest and the surface of the host particles. Dry impact blending presents the same drawbacks as mechanofusion when dealing with soft or heat-sensitive particles. The three dry coating techniques discussed hereunder allow to dry-coat such particles.

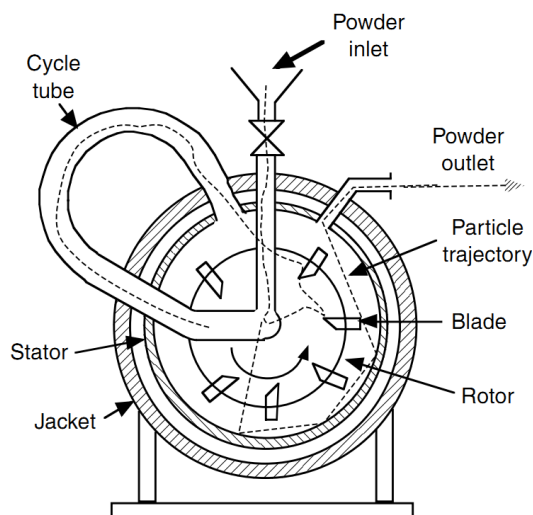


Figure 6.7: Schematic representation of a hybridizer [111].

Theta composer [111,118]

A theta composer consists of a slowly rotating vessel and a rotor rotating coaxially and more rapidly in the opposite direction (Figure 6.8). The particles are submitted to compressive and shearing forces when going through the small clearance between the vessel and the rotor. The forces generated are less intense than in mechanofusion or in dry impact blending so that the theta composer can be used to coat soft particles.

Rotating fluidized bed dry coater [111,118]

In addition to their application for wet spray coating [116], rotating fluidized beds can be used for dry coating. Indeed, as previously explained in section 2.1, high-G makes the fluidization of Geldart C-type guest particles possible. The strong centrifugal and shear forces generated by the rotation of the fluidization chamber allow breaking up the guest aggregates and attaching the guest onto the surface of the host particles. Moreover, the fluidization creates intense mixing resulting in a good distribution of the guest particles on the host.

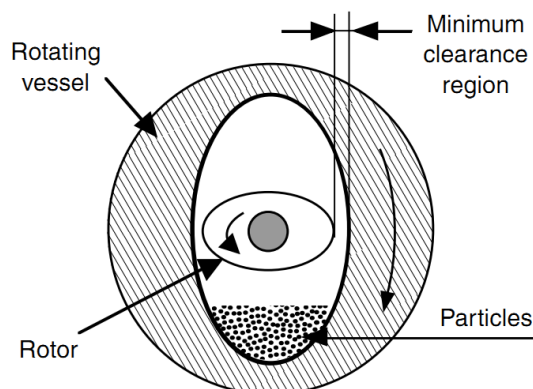


Figure 6.8: Schematic representation of a Theta composer [111].

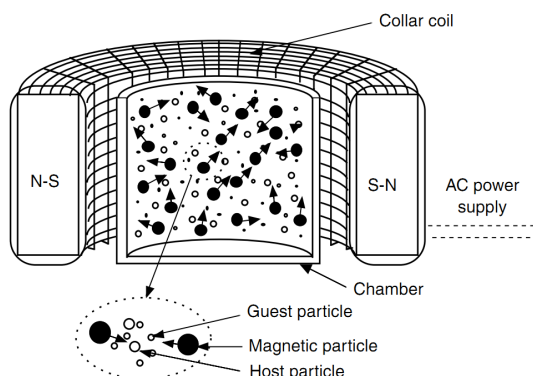


Figure 6.9: Magnetically assisted impaction coater [111].

Magnetically assisted impaction coater (MAIC) [111, 118]

The MAIC is a static vessel in which, in addition to guest and host particles, magnetic particles are fed. An external oscillating magnetic field is generated using electromagnets surrounding the vessel (Figure 6.9). As a result, the magnetic particles move vigorously in the vessel creating a pseudo fluidized bed of magnetic particles. The latter forces collisions between guest and host particles and provides sufficient mechanical forces to coat the host with the guest particles. To avoid coating contamination, the barium ferrite magnetic particles are coated with polyurethane.

High-G, low-temperature coating of cohesive particles in a vortex chamber

The content of this chapter has been published in Powder Technology, Vol. 258, pp. 242-251, 2014.

7.1 Introduction

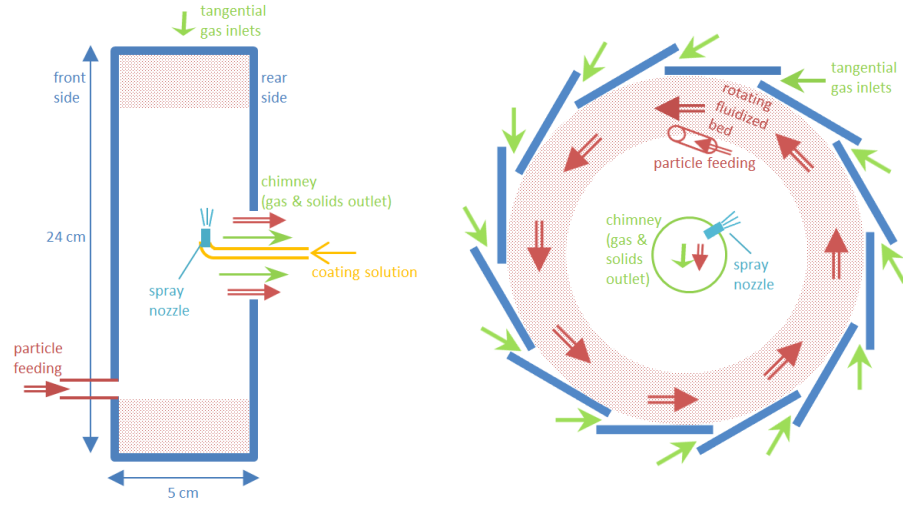
Fluidized bed coaters have been successfully used to coat particles with a diameter down to 100 μm [108]. Processing smaller or light, cohesive particles (Geldart C type) is challenging, the inter-particle Van der Waals forces becoming too important to allow fluidization. Chen et al. [12,110,115] dry-coated the surface of 15 μm cornstarch particles with nano-size particles in order to reduce the inter-particle forces. This allowed fluidizing these particles [12] and coating [110] or granulating [115] them with polymer films in a top-spray fluidized bed coater. The treatment with nano-size particles is, however, costly or may be unwanted, e.g. in the food industry. Geldart C type particles can also be successfully fluidized in cylindrically shaped rotating fluidized bed (RFB) devices [54,55]. High-G operation allows increasing the gas-solid slip velocity and intensifying interfacial momentum transfer, beside interfacial mass and heat transfer [27,53–55]. The impact of the inter-particle Van der Waals forces is as such significantly reduced. Rotating fluidized beds can be generated in rotating and static chambers. The use of a motor limits commercial application of rotating chambers because of undesired mechanical vibrations and sealing problems [53–55]. On the lab-scale, the coating and granulation of cohesive particles in such devices have

been successfully studied by Watano et al. [54,116].

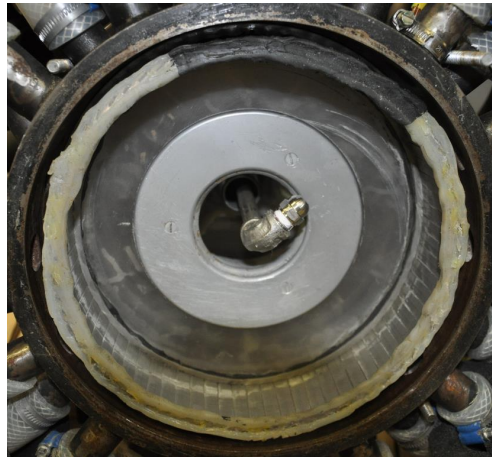
Vortex chambers allow the generation of a rotating fluidized bed in a static geometry [57,58,68]. As shown in Figure 7.1, the gas is injected tangentially through multiple inlet slots to generate a rotational flow and evacuated via a chimney, centrally positioned in one of the end plates [60,62]. Due to their inertia, particles do not follow the gas phase motion and accumulate in the vortex chamber, forming a rotating bed against the cylindrical wall. In cylindrically shaped rotating fluidized beds, the 'bed height' is in the radial direction, whereas the vortex chamber length is in the axial direction. The vortex chamber design is critical for avoiding immediate particle entrainment into the chimney by generating a centrifugal force sufficiently strong to counteract the radial gas-solid drag force. Whereas the latter depends on the radial gas velocity and, hence, on the gas flow rate and the circumferential surface area, the former depends on the tangential gas velocity and, hence, on the gas flow rate, the number of gas inlet slots and their size and inclination. The particle characteristics affect both forces. The amount of tangential momentum injected for a given gas flow rate can be increased by reducing the number of gas inlet slots and/or their size [76,79]. This logically results in an increased pressure drop over the vortex chamber. A minimum number of gas inlet slots is, however, required to ensure tangential uniformity [76,98]. Rosales-Trujillo and De Wilde [76] theoretically and experimentally showed that the slot size required for obtaining a dense and stable rotating fluidized bed decreases with decreasing particle size and/or density. This implies that for fluidizing cohesive particles, extremely small slots have to be used.

Because of the intensified interfacial mass, heat and momentum transfer, vortex chambers seem particularly suited for low-temperature drying [100], coating or granulation of fine particles. The maximum gas-solid slip velocity the terminal velocity or maximum settling velocity depends on the G-force against which particles are fluidized. In vortex chambers, the latter is generated by means of the gas flow. Hence, the terminal velocity increases with increasing gas flow rate. As shown in previous work [27], both the radial gas-solid drag force and the counteracting G-force increase roughly equally with increasing gas flow rate within a given range [60,62,79]. As a result, a vortex chamber of given design offers a high flexibility with respect to the gas flow rate. Varia-

7.1. Introduction



(a)



(b)

Figure 7.1: (a) Scheme and (b) picture of the vortex chamber with integrated spray nozzle (picture with a removed end plate).

tion of the gas flow rate to control agglomeration can then be considered.

Particle bed stability and uniformity is an important issue when using vortex chambers. Carrying out experiments with Geldart B and -D type particles, de Broqueville and De Wilde [62] showed that the solids loading is critical. At too low solids loadings, slugging and channeling may occur. The non-uniform particle and gas distribution in the vortex chamber is detrimental for gas-solid contact and related interfacial mass, heat and momentum transfer. The introduction of a dispersed coating solution may affect the particle bed stability. In particular at high coating solution concentration, particle agglomeration may become too pronounced and prevent the formation of a uniform rotating fluidized bed.

The present chapter aims at studying the fluidization and low-temperature coating of cohesive particles in a vortex chamber. The introduction of the coating solution via a centrally positioned spray nozzle directed towards the bed is considered. The influence of important operating conditions on the bed stability, agglomeration and the coating quality is then studied. The latter are analyzed by means of particle size distribution measurements, SEM imaging and an active component release test.

7.2 Experimental set-up

The experimental set-up consists of air, particle and coating solution feeding systems, the vortex chamber in which the actual particle coating is carried out, and a cyclone to separate entrained particles from the gas stream before its evacuation to the vent. The vortex chamber has a diameter of 0.24 m and a length of 0.05 m. The end plates are built in polycarbonate in order to allow the visualization of the particle bed. As shown in Figure 7.1, a polygonal design was adopted with 72 gas inlet slots to guarantee a uniform gas distribution. To generate a sufficiently high centrifugal force, the slot size was only 0.2 mm. The air is evacuated via a single, 0.1 m diameter chimney. A downstream cyclone connected to the vortex chamber uses the rotational motion of the evacuated gas to separate particles that might be entrained from the air [60].

The flow rate of the air coming from the compressor (Roots Blowers RVT70) is controlled by means of a mass flow controller. The air

7.2. Experimental set-up

is heated to 55 °C by a resistor with a PID controller. The particles are conveyed to the vortex chamber using compressed air and are fed via the front end plate of the vortex chamber. The particle feed rate is controlled using a screw and a sealing rotary valve (TBMA MAX-150). For the fluidization experiments, the air flow rate was varied between 250 and 400 Nm³/h (Normal cubic meters per hour, i.e. at 0 °C and 1 bar), the particle feed rate between 2 and 6.7 g/s. The particle bed rotation speed is measured in the axial center of the chamber and by means of a flag rotating freely around a centrally installed axis.

For the particle coating experiments, the coating solution is injected using a single-fluid pressure spray nozzle mounted via the chimney, centrally positioned and directed towards the outer cylindrical wall of the vortex chamber (Figure 7.1). The coating solution is heated to 90 °C using a Huber CC-208b thermostatic bath. The coating solution feed rate can be varied by setting manually the pressure in the solution tank. Remark that while increasing the coating solution injection pressure increases the coating solution flow rate, it simultaneously reduces the droplet size so that a higher flow rate of smaller droplets is to be dealt with. The solution tank is positioned on an electronic balance to measure the coating solution feed rate. The pressure nozzle is a Schlick Model 121 hollow cone spray nozzle with a 0.5 mm diameter orifice and a spray angle of 15 °. With the conditions imposed, it produces droplets with a diameter between 60 and 70 µm, depending on the coating solution injection pressure. A relatively large droplet size was chosen in this study to ensure that the droplets reach the particle bed and are not directly entrained by the air into the chimney.

The operating conditions are summarized in Table 7.1. The air and the particles are fed continuously. As long as no coating solution is sprayed into the vortex chamber, particle losses via the chimney compensate for the continuous particle feeding, that is, after a steady state particle bed is formed. No separate solids outlet is foreseen in the vortex chamber used. In the coating experiments, the coating solution is sprayed onto the particle bed once and during the specified coating time. At that moment, coated (less and more agglomerated) particles form. Because of their higher inertia, they do not leave via the chimney as easily as the uncoated particles. As a result, the coated particles start accumulating in the reactor and segregation of uncoated and coated par-

Chapter 7: High- G , low- T coating of cohesive particles in a vortex chamber

Operating variable		Unit	Value		
AIR	Flow rate	Nm ³ /h	250	400	
	Pressure before distributor	mbar	300-370	540-650	
	Inlet temperature	K	328.15		
PARTICLES (CORE MATERIAL)	Feed rate	g/s	2		
	Density	kg/m ³	260		
	Mean particle diameter	μm	69.71		
COATING SOLUTION (AQUEOUS)	Feed rate	g/s	1	1	1.63
	Injection pressure	bar	1.5	1.5	3
	Mean droplet diameter	μm	70	70	60
	Inlet temperature	K	363.15		
	Fraction coating material w_{cm}	wt-%	50		
	Coating material density	kg/m ³	1340		
	Coating time	s	5,15, 30,60	3,5,15, 30,60	5

Table 7.1: Operating conditions for the coating experiments.

ticles occurs. Only with a sufficiently long coating time, a new steady state would be reached where coated particles would also leave the vortex chamber via the chimney at a constant rate. With the short coating times required and applied, such a new steady state is never reached. The coated particles accumulate in the vortex chamber during the entire coating time, in contrast to uncoated particles. The coated particles are directly recovered by opening the vortex chamber after the experiment. At the end of the coating time, both the coating solution and particle feeding are stopped. The air feeding is stopped 45 s later to evacuate uncoated particles that might remain in the vortex chamber. The coated particles are analyzed and their quantity measured.

The core particles used are a Hiprotal whey protein alfa (supplied by FrieslandCampina Domo). They consist of 72% protein, 12% fat and some minor components like minerals and lactose. Their properties are summarized in Figure 7.2. According to the mean size (70 μm) and density (260 kg/m³), the core particles belong to the 1G-Geldart C group (Figure 7.2d) [17]. The SEM pictures in Figures 7.2a and 7.2b reveal particles that are non-uniform in size and generally spherical but damaged. In fact, the particles are partially fragmented and a fraction of hollow particles is observed. Smaller particles are contained in the hollow particles or found on their outer surface. For this first investigation of particle coating in a vortex chamber, particles that were chosen are extremely challenging to coat with conventional fluidized bed technology. The particle size distribution (PSD) of the particles has been measured both with sieves (Hosokawa Alpine Air Jet Sieve 200 LS-N)

7.2. Experimental set-up

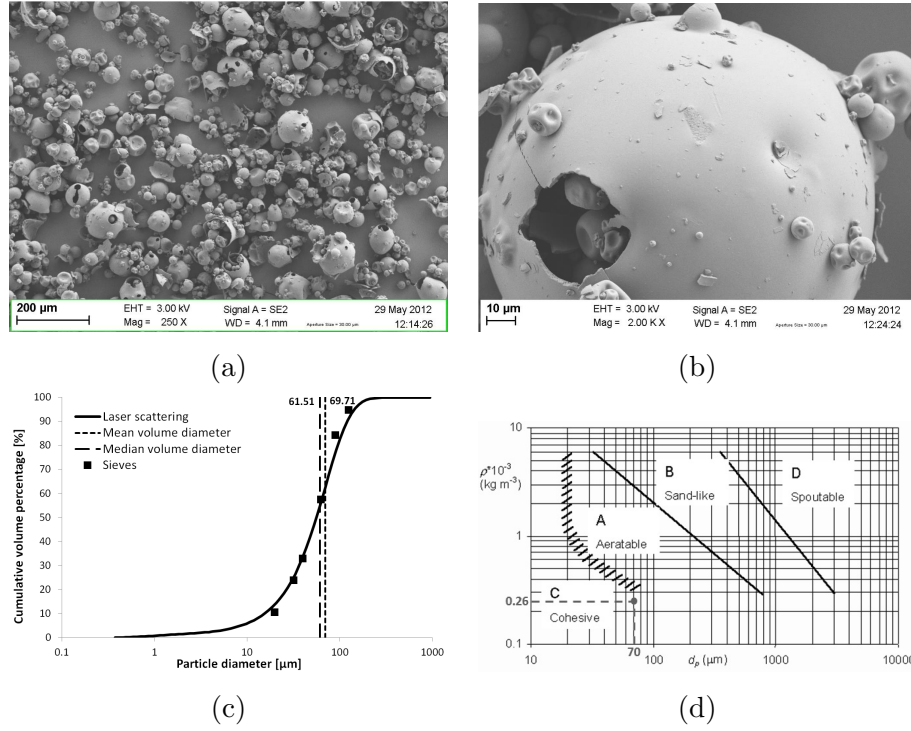


Figure 7.2: Characterization of the uncoated particles: (a) and (b) SEM pictures; (c) particle size distribution measured by laser scattering and with sieves; (d) Geldart classification [17].

and by laser scattering (Beckman Coulter LS 100Q). The cumulative volume percentage versus the particle size measured with the two methods is shown in Figure 7.2c.

The core particles were coated with a 50 wt% aqueous solution of maltodextrin 33DE (supplied by Roquette). Cochineal carmin pigment (food additive E120) was added to the coating solution to facilitate visualization of the presence of coating material on the particles and, in case, on the walls of the vortex chamber.

The particle size distribution (PSD) of the coated particles was measured by means of laser diffraction (Beckman Coulter LS 100Q) using isopropanol as liquid dispersant. After each experiment, three samples

were taken for PSD measurements. SEM pictures were taken at various magnifications to analyze the particle morphology, agglomeration and the coating quality. To quantitatively determine the coating quality, the fraction of core and coating material was measured and an active component release test carried out. The latter consists in passing the coated particles over a liquid in which one of the active components of the core material, in this case the fat, is soluble. The quantity of active component released into the liquid is then measured [109]. Repetitive experiments were carried out to verify the reproducibility of the experimental results and to determine the experimental errors.

In the next section, the fluidization of the cohesive particles in the absence and presence of a dispersed coating solution is discussed. Next, coating is addressed, focusing on the influence of some important operating conditions on agglomeration and the coating efficiency. Finally, the coating quality is analyzed in detail.

7.3 Fluidization of cohesive particles and influence of a dispersed coating solution on the bed stability

Not feeding a dispersed coating solution, the experiments showed that a rotating fluidized bed of (uncoated) cohesive particles ($< d_p > = 70 \mu\text{m}$, $\rho_s = 260 \text{ kg/m}^3$) could be obtained in a vortex chamber with 72 gas inlet slots, 0.2 mm wide each. A picture of the particle bed at given operating conditions is shown in Figure 7.3a. The use of a vortex chamber with a larger total gas inlet surface area, i.e. 24 x 3 mm slots, did not allow generating a sufficiently strong centrifugal force to avoid almost direct entrainment of the particles with the gas into the chimney. In contrast to what was observed with Geldart A, B, and D type particles in other vortex chambers [62], with the Geldart C type particles, particle losses via the chimney cannot be avoided and the retention capacity, i.e. the amount of particles remaining infinitely in the chamber after stopping the particle feeding, is negligible, even in the vortex chamber with 72 x 0.2 mm slots. Hence, continuous particle feeding is required to obtain a rotating fluidized bed. A separate solids outlet not being present in the current set-up, the rate of particle losses via the chimney equals the particle feed rate once a steady state reached. Important for the application studied is that the residence time of the particles in the vortex

7.3. Fluidization of cohesive particles and influence of a dispersed coating solution on the bed stability

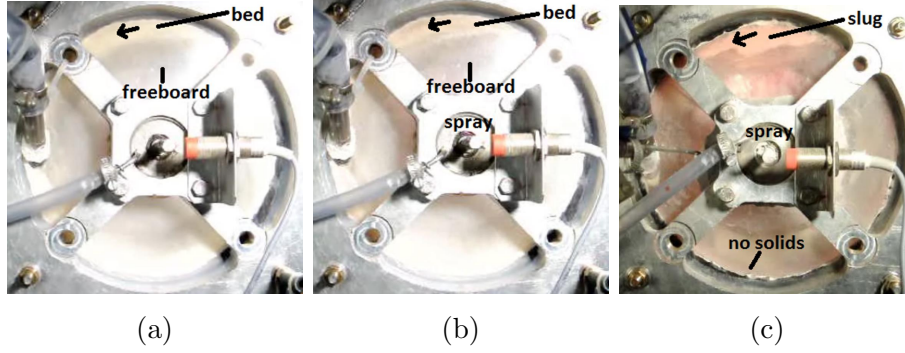


Figure 7.3: Stable rotating fluidized bed of Geldart-C type core particles (a) without injection of a coating solution and (b) while injecting the coating solution (snap-shot after $\tau_c = 20$ s). (c) Slugging due to excessive injection of coating solution. Operating conditions: air flow rate = 250 Nm³/h, core particles feed rate = 2 g/s, coating solution feed rate = 1 g/s.

chamber with 72 x 0.2 mm slots, τ_s , is sufficiently high to allow coating taking place.

Increasing the gas flow rate results in an increase of both the centrifugal force and the counteracting radial gas-solid drag force. With the applied solids loading and in the gas flow rate range studied, the increase of the centrifugal force was somewhat more pronounced, from 70G at 250 Nm³/h (particle bed rotating at 12 rps) to 200G at 400 Nm³/h (particle bed rotating at 20 rps). This results in a slight bed contraction, i.e. a denser and more uniform bed, with increasing gas flow rate. As previously shown with other types of particles [62], within a given range, the solids loading in the vortex chamber increases nearly proportionally with the particle feed rate. Hence, the particle residence time, $\tau_s = V_s/Q_s$, varies only slightly with varying particle feed rate. The flexibility in the gas and particle feed rates is a particular advantage of vortex chambers.

When introducing a dispersed coating solution, segregation between the uncoated and coated particles occurs and coated particles accumulate in the vortex chamber. The coated particles have a larger mean particle diameter, mainly due to agglomeration, and higher density and

as such a higher inertia. Hence, the coated particles concentrate towards the outer cylindrical wall of the vortex chamber. This prevents further collisions with droplets injected via the spray nozzle, the latter colliding mostly with uncoated particles in the inner region of the particle bed. In the experiments carried out, the coating time was always too small to reach a steady state during coating and coated particles kept accumulating in the vortex chamber during the entire coating time. Only minor losses of non-agglomerated coated particles and no losses of agglomerated coated particles via the chimney were observed, both during and after stopping the particle feeding. The coated particles being well retained in the vortex chamber of the given design allows them to be recovered easily after stopping the air supply. Fully continuous operation requires a separate solids outlet, typically installed in the end plate of the vortex chamber opposite the end plate through which the particles are fed.

The presence of the dispersed coating solution can affect the bed stability. Whereas in a given coating solution feed rate and coating time range the particle bed was found to remain stable and fluidized (Figure 7.3b), a too high coating time and/or rate causes slugging and defluidization of the bed, as shown in Figure 7.3c. The influence of the coating time is explained by the growing bed thickness of wetted and, hence, heavier particles in the vortex chamber due to accumulation. Because of the cylindrically shaped particle bed and the conical spray pattern, this leads to a reduced contact surface area between droplets and bed and as such a locally higher coating solution concentration in the bed. When slugging occurs particles are sticking together, producing a paste that rotates in the bed. Slugging is the result of insufficiently fast drying of wetted particles. Operation at higher gas flow rates and resulting in higher slip velocity and coefficients of mass, heat and momentum transfer allows extending the range of coating time and/or rate in which a stable rotating fluidized bed can be maintained. This is illustrated in Figure 7.4 showing data for the given vortex chamber design and operating conditions. Operation at higher gas inlet temperatures is also expected to allow operation at higher coating times/rates because of an increased evaporation rate.

7.4. Influence of the operating conditions on agglomeration and on the coating efficiency

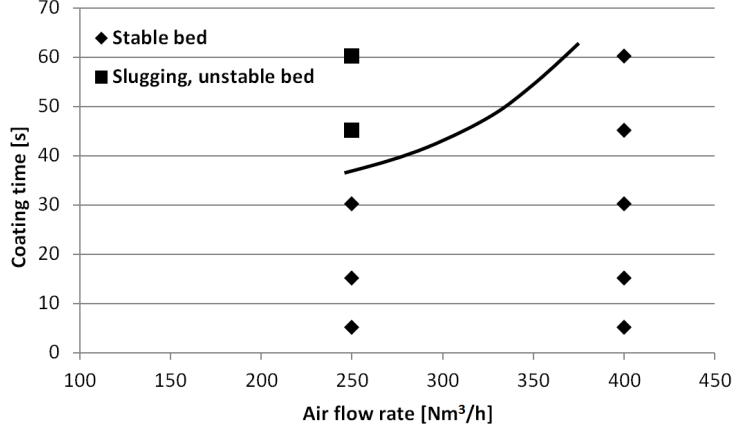


Figure 7.4: Stability of the rotating fluidized bed as a function of the air flow rate and the coating time. Operating conditions: particle feed rate = 2 g/s, coating solution feed rate = 1 g/s. The line in the figure is indicative.

7.4 Influence of the operating conditions on agglomeration and on the coating efficiency

Controlling agglomeration is a critical issue in coating processes. The process efficiency is an additional concern. For batch process, a coating efficiency factor is typically defined as the fraction of injected coating material that is deposited on the particles [119]. In the present work, the core particles are fed continuously and, accounting for the experimental procedure, an overall coating efficiency factor, η , can be defined as:

$$\eta = \frac{(\text{mass of coated particles recovered})}{\tau_c \cdot (\text{core particles feed rate} + w_{cm} \cdot \text{coating solution feed rate})} \quad (7.1)$$

with τ_c the coating time and w_{cm} the mass fraction coating material in the coating solution. The ratio is taken of the mass of coated particles recovered and the mass of particles and coating material fed during the applied coating time. Accounting for the composition of the coated particles as reported in Table 7.2 allows the separate calculation of particle and coating material efficiency factors. For the latter:

$$\eta_{cm} = \frac{(\text{mass of coating material recovered})}{\tau_c \cdot (w_{cm} \cdot \text{coating solution feed rate})} \quad (7.2)$$

With the given vortex chamber and in the range of operating conditions studied, agglomeration is seen to occur during coating and to increase somewhat less than proportionally with increasing coating time τ_c . This is shown in Figure 7.5a, focusing on the mean particle diameter $\langle d_p \rangle$, a direct measure of agglomeration, and in Figure 7.6, showing the measured particle size distributions. In what follows, "coated particles" will refer to both non-agglomerated and agglomerated coated particles, knowing that agglomeration takes place. The size of the agglomerated coated particles varies between 200 and 450 μm , depending on the operating conditions. As mentioned in the section 2.1, vortex chambers offer a particular flexibility with respect to the gas flow rate, the latter affecting both the centrifugal force and the counteracting drag force in similar ways. Operation at higher gas flow rates does not only increase the supplied heat flux. It implies higher particle bed rotational speeds and resulting increased shear in the particle bed and shorter coating time per spray passage, and higher gas-solid slip velocities and consequently intensified interfacial mass and heat transfer leading to faster evaporation of the coating solution. This allows reducing agglomeration, if desired, as illustrated in Figure 7.5a. Operation at higher gas flow rates comes, however, at the cost of higher losses of uncoated particles through entrainment into the chimney and correspondingly lower overall coating efficiencies, as shown in Figures 7.5b and 7.5c. Losses of coating solution droplets into the chimney or evaporation of coating solution droplets before reaching the bed are eventually also possible. In the range of operating conditions studied, the coating material efficiency factor is close to 100%, indicating that the droplets are sufficiently large to avoid direct entrainment into the chimney or complete evaporation before reaching the bed. The amount of coated particles recovered logically increases with the coating time. The increase is nearly proportional at lower gas flow rates, somewhat less than proportional at higher gas flow rates. The overall coating efficiency factor η is roughly between 25 and 65% in the range of operating conditions studied and is seen to go through a maximum and then decrease with increasing coating time. The relatively low values reflect the pronounced entrainment of uncoated particles into the chimney. The initial increase of the overall coating efficiency with coating time, particularly well observed at higher gas flow rates, is due to the droplets of coating solution preventing entrainment of uncoated particles into the chimney. With increasing coat-

7.4. Influence of the operating conditions on agglomeration and on the coating efficiency

ing time, the particle bed grows and its composition changes, heavier coated particles accumulating and segregation taking place. As a result, the particle bed rotational speed decreases, the particle bed expands, and losses of uncoated particles via the chimney are facilitated. A more uniform distribution of the coating solution using multiple spray nozzles could reduce losses of uncoated particles via the chimney and increase the overall coating efficiency. Because of the high coating material efficiency with the current droplet size, the use of smaller droplets could be considered.

The particle size distribution curves in Figure 7.6 indicate that applying the shortest coating time (5 s), a significant fraction of the particles remains non-agglomerated, explaining the secondary peak close to the peak of the uncoated particles (the coating layer is only about 1-2 μm thick). This secondary peak is less pronounced at higher gas flow rates, suggesting more pronounced losses of non-agglomerated particles via the chimney. Figure 7.6 again illustrates the reduced agglomeration rate at higher gas flow rate. The flexibility with respect to the gas flow rate is, hence, an important feature of vortex chambers allowing the control of agglomeration.

Figure 7.7 shows the influence of the coating solution injection pressure $p_{l,inj}$ on the amount of coated particles recovered (Figure 7.7a) and on their size distribution (Figure 7.7b) for the given operating conditions, including given coating time. The overall coating efficiency is also shown in Figure 7.7a. Increasing the coating solution injection pressure from 1.5 to 3 bar only slightly increases agglomeration, the mean particle diameter of the coated particles shifting from 205 to 209 μm . The corresponding increase of the coating solution flow rate from 1 to 1.63 g/s was expected to increase agglomeration, but is compensated by the decrease of the droplet mean diameter from 70 to 60 μm , allowing faster evaporation. Increasing the coating solution injection pressure increases, however, significantly the amount of coated particles recovered and the related overall coating efficiency (Figure 7.7a). This is explained by the increased coating solution feed rate and the presence of a higher number of smaller droplets preventing fine particles from being entrained into the chimney. To independently control the coating solution feed rate and the droplet size, the use of a two-fluid spray nozzle could be considered.

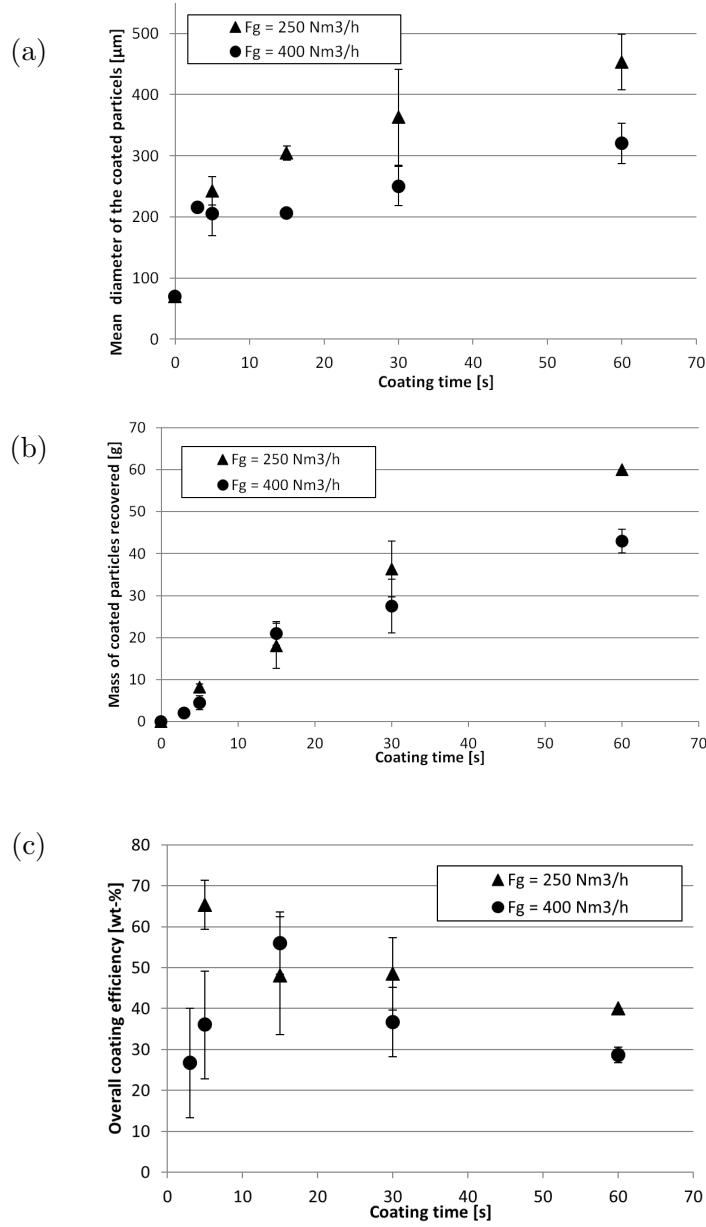


Figure 7.5: Influence of the air flow rate and the coating time on the (a) mean diameter of the coated particles; (b) amount of coated particles recovered; and (c) overall coating efficiency. Operating conditions: core particles feed rate = 2 g/s, coating solution feed rate = 1 g/s.

7.4. Influence of the operating conditions on agglomeration and on the coating efficiency

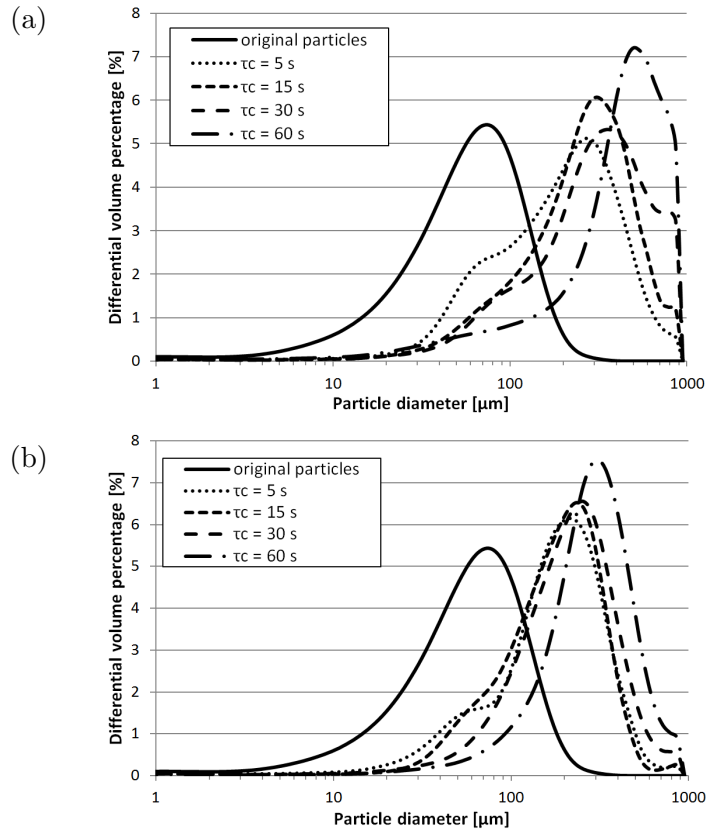


Figure 7.6: Particle size distribution of the coated particles for air flow rates of (a) 250 Nm³/h and (b) 400 Nm³/h and for varying coating time. Operating conditions: core particles feed rate = 2 g/s, coating solution feed rate = 1 g/s.

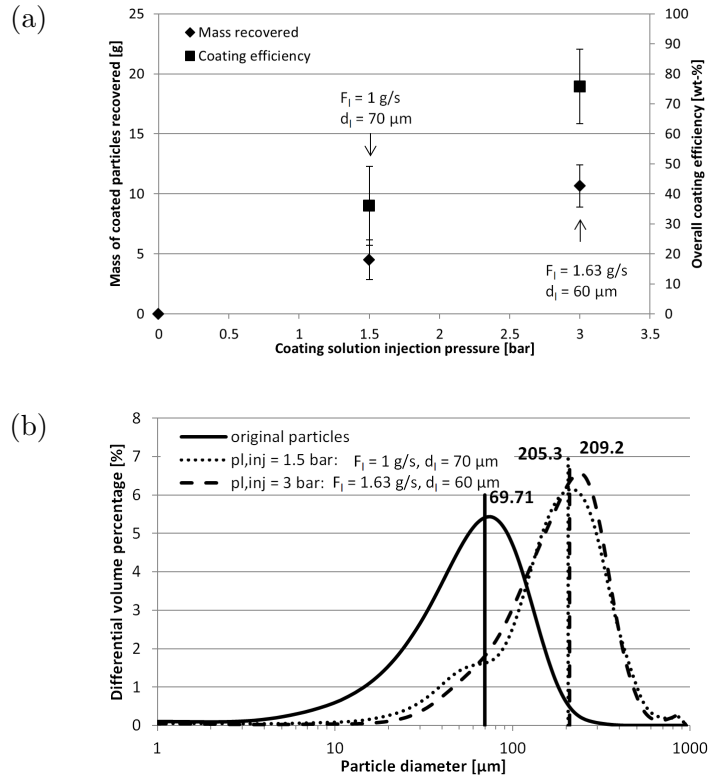


Figure 7.7: Influence of the coating solution injection pressure on the (a) amount of coated particles recovered and the overall coating efficiency; and (b) particle size distribution of the coated particles. Operating conditions: air flow rate = 400 Nm³/h, core particles feed rate = 2 g/s, coating time = 5 s.

7.5 Coating quality

The coating quality of two less and more agglomerated coated particles is compared in Figures 7.8 and 7.9 and in Table 7.2. During coating, the agglomeration was controlled by means of the gas flow rate only. Other operating conditions were identical. Figure 7.8 shows the particle size distribution (PSD) of the coated particles considered. The SEM pictures shown in Figure 7.9 confirm the agglomeration observed through the PSD measurements and involving a couple of particles, as well as a fraction of the coated particles remaining non-agglomerated. The coating quality is good, especially accounting for the nature of the particles that were coated, the low gas feed temperature and the large droplet size that was used. Only very few cracks in the coating layer were found, despite the high shear to which the particles were exposed in the vortex chamber. The coating layer is relatively thin. This is confirmed by the measurements of the fraction core and coating material in the coated particles (Table 7.2). The less and more agglomerated coated particles contain nearly the same fraction of coating material, indicating the presence of coating material in between agglomerated particles and confirming agglomeration is taking place as a result of inter-particle collisions involving a particle with an insufficiently dried coating layer. Assuming spherical particles with a diameter equal to the mean diameter ($70\text{ }\mu\text{m}$), the average coating layer thickness is found to be $1.3\text{ }\mu\text{m}$ for the given operating conditions. As shown in Table 7.2, the good quality of the coating of both the less and more agglomerated coated particles is reflected in a 60% reduction in the release of an active component.

7.6 Conclusions

High-G fluidization results in intensified interfacial mass, momentum and heat transfer, opening perspectives for low-temperature particle coating applications. The generation of a rotating fluidized bed of cohesive particles in a vortex chamber requires sufficiently small gas inlet slots. The latter allows balancing the centrifugal force and the counteracting radial gas-solid drag force and a sufficiently high particle residence time to allow coating. A stable rotating fluidized bed can be maintained when injecting a dispersed coating solution via a centrally positioned single-fluid pressure spray nozzle directed towards the bed, at least within a sufficiently wide range of coating solution feed rates and

Powder	$\langle d_p \rangle$ [μm]	$\langle \rho_s \rangle$ [kg/m^3]	Core material [wt-%]	Coating [$\text{g}_{\text{coating}}/\text{g}_{\text{core}}$]	A.c. release [%]	Reduction of a.c. release
Uncoated particles	69.7	260	100	-	60	-
Less agglomerated coated particles $F_g = 400 \text{ Nm}^3/\text{h}$	205.3	657	63.23	0.58	24	60%
More agglomerated coated particles $F_g = 250 \text{ Nm}^3/\text{h}$	242.6	660	62.96	0.59	24	60%

Table 7.2: Composition of the uncoated particles and of less and more agglomerated coated particles and data of an active component (a.c.) release test. Less agglomerated coated particles: air flow rate = $400 \text{ Nm}^3/\text{h}$; More agglomerated coated particles: air flow rate = $250 \text{ Nm}^3/\text{h}$. Other operating conditions: core particles feed rate = 2 g/s , coating solution feed rate = 1 g/s , coating time = 5 s .

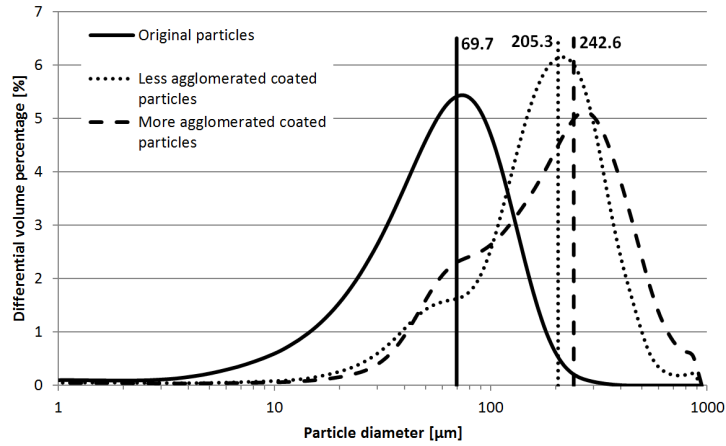


Figure 7.8: Particle size distribution of the uncoated particles and of less and more agglomerated coated particles. Less agglomerated coated particles: air flow rate = $400 \text{ Nm}^3/\text{h}$; more agglomerated coated particles: air flow rate = $250 \text{ Nm}^3/\text{h}$. Other operating conditions: core particles feed rate = 2 g/s , coating solution feed rate = 1 g/s , coating time = 5 s .

7.6. Conclusions

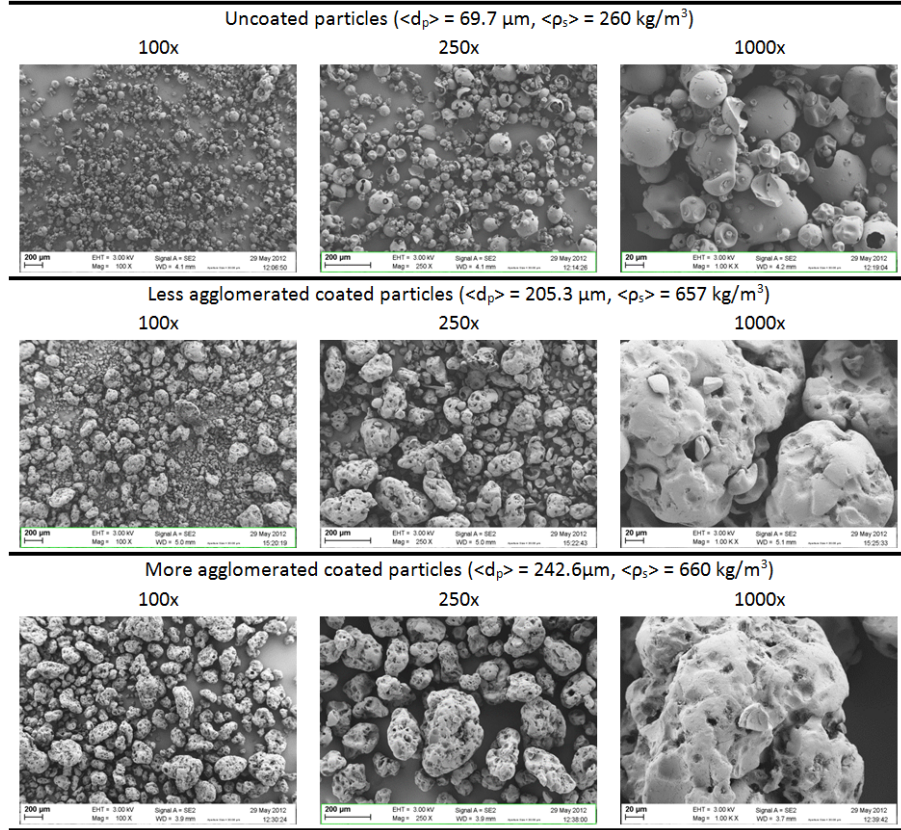


Figure 7.9: SEM pictures of the uncoated particles and of less and more agglomerated coated particles. Less agglomerated coated particles: air flow rate = $400 \text{ Nm}^3/\text{h}$; More agglomerated coated particles: air flow rate = $250 \text{ Nm}^3/\text{h}$. Other operating conditions: core particles feed rate = 2 g/s , coating solution feed rate = 1 g/s , coating time = 5 s .

coating times. Relatively large droplets were used to avoid direct entrainment into the chimney. The presence of an excess coating solution in the particle bed leads to defluidization and slugging. Coated particles can be produced at low gas feed temperatures, thanks to the relatively high gas flow rates and high gas-solid slip velocities that can be achieved in the vortex chamber. Agglomeration is observed and related to the use of relatively large droplets. Vortex chambers being particularly flexible with respect to the gas flow rate, the use of the latter to control agglomeration is demonstrated. Increasing the gas flow rate intensifies gas-solid mass and heat transfer and increases the shear in the rotating particle bed, significantly reducing agglomeration. Within the range studied, the coating solution injection pressure is seen to have only a minor effect on agglomeration and on the very high coating material efficiency, but a significant effect on the overall coating efficiency. The generation of a higher concentration of smaller droplets prevents entrainment of uncoated particles by the gas into the chimney. SEM images and an active component release test show a good quality coating, both for less and more agglomerated coated particles. The short coating time required in a vortex chamber type particle coater and the segregation between coated and uncoated particles imply that for fully continuous operation, a separate outlet for the removal of coated particles is to be foreseen.

Nomenclature

d_l	m	mean diameter of the coating solution droplets
d_p	m	mean particle diameter
F_g	Nm ³ /s	air (mass) flow rate (Normal m ³ /s, i.e. at 0 °C and 1 bar)
F_l	kg/s	coating solution feed rate
F_s	kg/s	core particles feed rate
$p_{l,inj}$	Pa	coating solution injection pressure
Q_s	m ³ /s	core particles volumetric feed rate
V_s	m ³	particle bed volume
w_{cm}	-	mass fraction of coating material in the coating solution

Greek letters

η	-	overall efficiency factor
η_{cm}	-	coating material efficiency factor
ρ_s	kg/m ³	particle density (core material)
τ_c	s	coating time
τ_s	s	residence time of the core particles in the vortex chamber

Outlook on vortex chamber spray drying

8.1 State of the art

8.1.1 Introduction

Spray drying consists in transforming a pumpable liquid feed (solution, emulsion, suspension, etc.) into dry particles. When dealing with a solution, the liquid feed comes into contact with a hot drying medium - in most applications, air - which evaporates the solvent creating dry particles of solute. Because the evaporation allows maintaining a low droplet temperature, high drying medium temperature can be used without damaging the product, even if the latter is heat sensitive. The size of the spray-dried particles can range from 100 nm to 2-3 mm depending on the operating conditions. The spray drying process involves the atomization of the liquid, the contact between the so-formed droplets and the hot drying medium, the evaporation of the solvent/drying of the particles and finally the separation between the product and the exhaust drying medium. When the feed is a solution, a precipitation/nucleation step is also implicated. Liquid atomization can be performed via a pressure nozzle, a two-fluid nozzle or a rotary atomizer. Contacting the droplets and the drying medium can be done through different ways depending on the type of spray dryer. Depending on the application, the process should be designed to produce spray-dried particles with desired size (and PSD), density (compact or hollow), texture, flow properties, solubility, etc. The final spray-dried product varies from a regular powder to agglomerated particles depending on the liquid feed, the atomization technique, the dryer type and the operating conditions [120–122].

It should be stressed that spray drying differs from spray congealing and freeze drying. Spray congealing consists in transforming a liquid melt into dry particles by a phase change (solidification). In freeze drying, a solution is frozen and then placed under vacuum. This vacuum aims to sublimate the solvent (most of the time water) without turning into a liquid phase. Consequently, spray congealing involves only heat transfer while spray drying and freeze drying involve both heat and mass transfer [121].

Spray drying is currently widely used in the pharmaceutical industry for preparation of tableting constituents, vaccines, etc. and in the food industry to facilitate transport and storage of various products [120,121].

8.1.2 History

Spray dryers started to be developed in the 1870's for the drying of egg products. The atomization was performed through a spray nozzle but the poor design of these first spray dryers prevented a continuous operation. As a result, no commercial application was developed up to the 1920s when spray drying started to be widely used in the dairy industry. The production of milk powder is still today a major application of the spray drying process.

Spray drying was then successfully developed for pharmaceutical and food-processing applications in the 1940s and 1950s. In particular, spray drying played an important role in the weight reduction in food transportation.

From the 1960s up to now, both the drying chamber and the atomization techniques have been continuously improved to allow the successful spray drying of a wide range of products with well-defined size, shape and porosity. Multi-stage spray dryers have also been developed to allow handling heat sensitive materials. [121,123]

8.1.3 Mechanism

The spray drying mechanism is shown in Figure 8.1. When a droplet first contacts the hot drying medium, evaporation rapidly occurs at its external surface leading to the formation of a thin shell of solute material (phase 1). Next, the evaporation rate becomes limited by the water

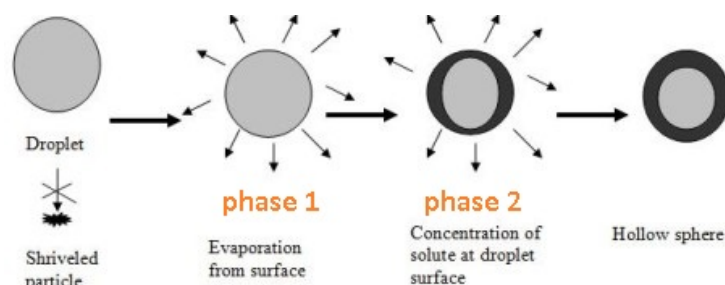


Figure 8.1: Schematic representation of the spray drying mechanism [121].

diffusion from the inner liquid core to the outer surface of the sphere (phase 2). Because of the evaporation, phase 1 leads to a shrinkage of the droplets so that the size of the produced particles is smaller than the droplets size. Depending on the materials used and the operating conditions, phase 2 can lead to the formation of cavities inside the particle [120,121].

As the latent heat required for the evaporation is supplied by the hot drying medium, the temperature of the latter decreases sharply at the location where phase 1 occurs. However, it should be remarked that spray drying is a low thermal efficiency process because a small part of the drying medium comes effectively into contact with the droplets to be dried. In some cases, recycling the drying medium is considered [121].

An important issue in spray drying is the droplet/particle temperature. Indeed, this temperature has to be kept as low as possible to avoid thermal degradation of the product. In phase 1, the fast evaporation rate ensures a low droplet temperature. However, phase 2 comes with an increase of the droplet/particle temperature as the evaporation rate decreases because of diffusion limitations. Consequently, when working with heat sensitive solids, spray dryers are designed to feed co-currently the hot drying gas and the droplets. This ensures that phase 2 occurs when the droplets/particles are in contact with the drying medium near its outlet temperature. Co-current spray drying thus makes sure that the produced solids particles are never heated above the drying medium outlet temperature (45-80 °C) even if the inlet temperature of the drying medium can be much higher (150-220 °C). Counter-current operation of-

fers a better energy efficiency as the fresh drying medium is in contact with the driest particles but cannot be used with heat-sensitive particles. In some cases, a mixed flow, between co- and counter- current, can be used [120,121,124].

Ideally, the contact time between the particles and the drying medium is only a few seconds. However, the turbulence required to achieve fast drying rates can cause some particles to be recirculated. These particles can consequently be exposed to higher temperatures and can be degraded in case they are heat sensitive [121].

8.1.4 Liquid pre-concentration

Before the atomization, the liquid is generally concentrated as much as possible to reduce the amount of water to be evaporated and the related energy consumption. The pre-concentration comes generally with increased particle density and size. Indeed, the droplets shrinkage is reduced and less voids are generated by evaporation. However, the pre-concentration of the liquid solution also increases its viscosity, which makes the atomization more difficult.

8.1.5 Atomization techniques

The atomization aims at breaking a bulk liquid into small droplets. The smaller the droplets, the higher their specific surface in contact with the hot drying medium and the faster the related mass and heat transfer rates. Typically, the atomized droplets have a mean size ranging between 10 and 300 μm , generating a surface area between 20 and 600 m^2 per litre solution fed [123]. The characteristics of the droplets also affect the shape, size and density of the produced particles. The droplets size distribution has to be narrow to ensure that every drop is equally dried. The three main atomization techniques are the pressure nozzle, the two-fluid nozzle and the rotary atomizer. Ultrasonic energy and vibrations atomizers have also been tested but are currently not used in commercial applications. The choice between the different atomization techniques is based on the feed properties, the type of dryer and on the desired product properties and capacity [120–122,125].

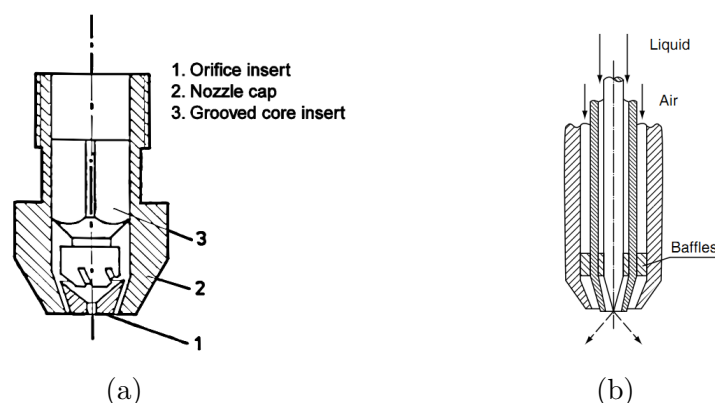


Figure 8.2: Scheme of a (a) pressure nozzle [125] and (b) a two-fluid nozzle [122]

Pressure nozzles

The pressure nozzle or single-fluid nozzle (Figure 8.2a) makes use of the pressure energy. The liquid is forced to rotate by an inner swirl chamber before reaching the outlet orifice of the nozzle. As a result, the pressure energy is converted into a rotational energy which leads to a centrifugal hollow cone thin film of rotating liquid. The film then breaks into small droplets after overcoming surface tension. The smaller the orifice and the higher the injection pressure, the smaller the produced droplets. The pressure nozzle is the most commonly used atomization device. However, the small size of the orifice, 0.5 to 3 mm, limits the liquid feeding rate to about 750 kg/h. Moreover, while working with a specific nozzle, the droplets size and the liquid feeding rate cannot be controlled independently as the latter is determined by the injection pressure. Therefore, pressure nozzles offer a limited flexibility [121–123, 125, 126].

Two-fluid nozzles

In a two-fluid or pneumatic nozzle (Figure 8.2b), compressed gas - usually air - is injected in addition to the liquid to be atomized. The liquid can be fed under pressure or by gravity. The compressed gas is injected at high velocity and its kinetic energy generates high frictional shearing

forces that break the bulk liquid into small droplets. Two-fluid nozzles are suited to produce very fine droplets or to atomize highly viscous liquid feeds. They also allow the independent control of the liquid feeding rate and the droplets size through the air-liquid ratio. However, commercial use of two-fluid nozzles is limited because of the high cost related to the compressed air consumption [121, 122, 125].

Rotary atomizers

A rotary atomizer (Figure 8.3) consists of a high-speed rotating wheel at the center of which bulk liquid is introduced. The wheel has typically a size ranging between 0.05 and 0.5 m and rotates between 2000 and 30000 rpm. The centrifugal force generated by the rotation of the disk forces the liquid to move radially outward, generally through vanes in the wheel. When the thin liquid film on the disk reaches the periphery, it breaks into small droplets with a spray angle of 180° . The droplets size can be controlled by the disk rotational speed. Rotary atomizers allow producing very fine droplets without involving high pressure and present a large flexibility with respect to the liquid flow rate. The capacity ranges from 1 kg/h for the smallest, pneumatically driven lab scale units to 200 tons/h for the largest commercial scale units driven by 1000 kW motors. Moreover, no clogging like in spray nozzles can occur. Rotary atomizers are relatively expensive and integration of a high-speed moving part in a commercial unit is challenging. This explains the continuing popularity of spray nozzles in the industry [121–123, 125, 126].

8.1.6 Dryer type

The drying chamber has to be designed to ensure efficient mixing and contact between the hot drying gas and the droplets/particles and related high evaporation/drying rates. A gas distributor, i.e. a perforated plate, is typically used upstream to the drying chamber to feed the gas uniformly. Typically, the drying chamber consists in a vertical or horizontal cylindrical vessel. Its dimensions are chosen to meet two criteria. First, its volume has to ensure an adequate contact time between the droplets and the drying medium. For that reason, spray dryers height can reach up to 10 m. Secondly, the height and diameter of the drying chamber have to be designed as to minimize the contacts between not sufficiently dried droplets and the walls of the vessel. As a result, rotary atomizers are typically used in large diameter and small height vessels

8.2. Spray drying of milk powder in vortex chamber: experimental proof of concept

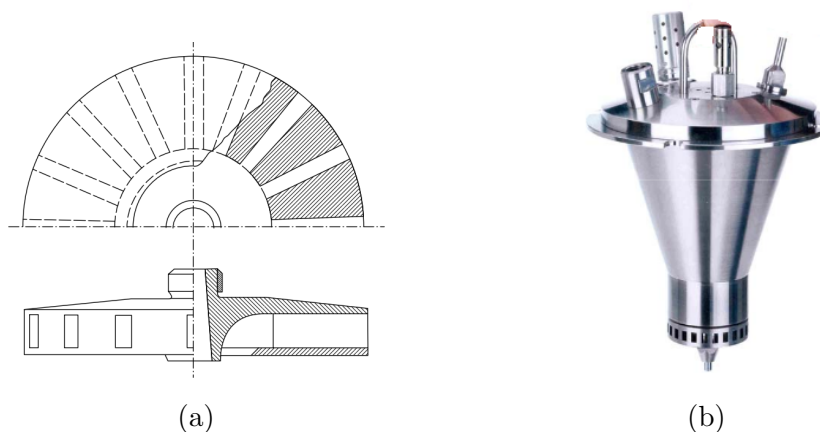


Figure 8.3: (a) Scheme of a rotating wheel [122] and (b) picture of a Niro rotary atomizer

while spray nozzles are operated in tall, small diameter vessels. The bottom of the spray dryer is typically conical, facilitating the recovery of the produced particles. The drying medium leaves the vessel through a side tube above this cone. A cyclone and a filter can be used downstream to separate entrained particles from the exhaust drying medium. Figure 8.4 shows a schematic representation of a spray drying process that makes use of a co-current drying chamber [120, 121, 124].

When dealing with highly heat-sensitive particles, using a two stage dryer may have to be considered. In this case, the first stage produces particles that present a moisture content 10-15 % higher than desired. These particles are then transferred to the second stage dryer - typically, a fluidized bed - for the final drying [120, 121].

8.2 Spray drying of milk powder in vortex chamber: experimental proof of concept

The work presented in this section is the object of the patent EP 12190243.1 submitted to the Netherlands patent office.

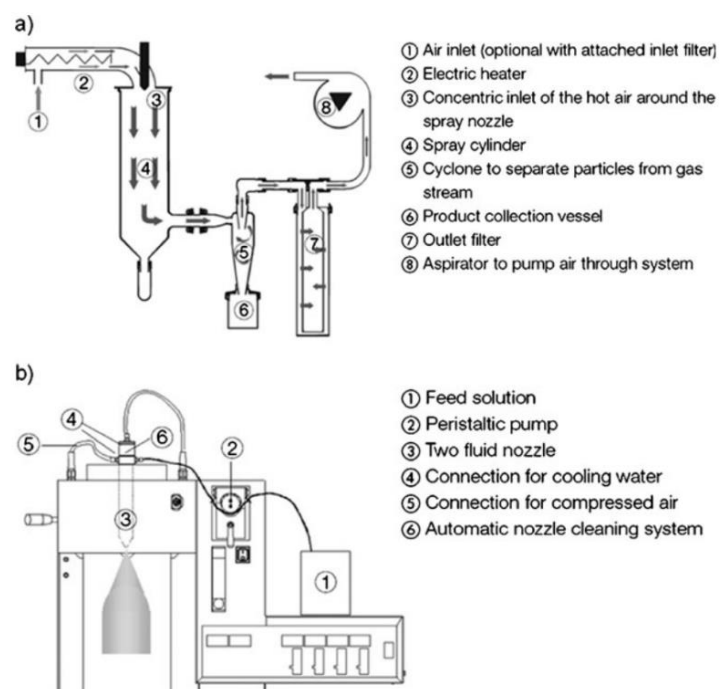


Figure 8.4: Schematic representation of the spray drying process: (a) air flow; (b) liquid feeding [124].

8.2.1 Introduction

This section reports on the development of a vortex chamber type spray dryer. The experiments aimed at demonstrating the feasibility of spray drying in a vortex chamber. Some important design aspects were studied. The major difference with respect to the other applications studied in this thesis, i.e. particle drying and particle coating, is the absence of a particle bed. Hence, a free vortex type flow pattern can be maintained and extremely high gas velocities achieved. Interfacial mass and heat transfer can as such be significantly intensified with respect to other technologies (see section 8.1), opening perspectives for the development of spray dryers of reduced size.

8.2.2 Experimental set-up & operating conditions

Spray drying experiments with a milk powder solution have been carried out in a hot flow unit, build only from steel and aluminium (Figure 8.5). Process gas is fed to the latter via a 30 kW resistor, allowing operation at temperatures up to 300 °C. Axial multi-zone operation is also possible with separate air inlets on the front and rear (chimney) side of the vortex chamber. With dual-zone operation, the spray nozzle is installed in the hot zone, located at the front side of the vortex chamber. The cold zone is located at the rear side. The milk solution is fed using a Schlick Model 970 Form 0 S4 two-fluid nozzle, allowing to independently control the milk feeding rate and the droplets size.

The vortex chamber design parameters and operating conditions used in the different experiments are summarized in Table 8.1. Estimates of the maximum allowable milk feeding rate and droplets size followed from calculations of the water evaporation capacity in the vortex chamber. The calculations are made with the assumption of either (i) completely mixed conditions (CSTR) for the droplets, (ii) the air and the droplets flowing co-currently or (iii) the air and the droplets flowing counter-currently. Counter-current operation represents the best case as the fresh air contacts the driest droplets, with highest surface to volume ratio. By-passing of droplets by the gas was neglected. An additional hypothesis was that the droplets make one rotation before leaving the vortex chamber. The vortex chamber characteristics used for the calculations were similar to those used in the present study. The results for an air flow rate of 600 Nm³/h and an inlet droplet size of 30 µm

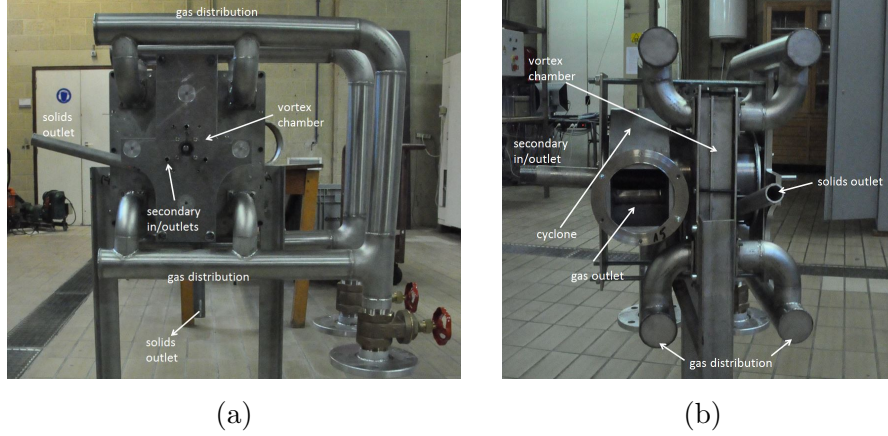


Figure 8.5: (a) Front view and (b) side view of the hot flow unit.

are given in Table 8.2.2 for different air inlet temperatures. For more details, see Annex A.

In a first set of experiments (exps. 1-5), operation with the spray nozzle located inside the vortex chamber was tested, as shown in Figures 8.6a and 8.6b. This implies the insertion of two tubes inside the vortex chamber to feed the solution and the atomizing air to the spray nozzle. Both single-zone (exps. 1-4) and dual-zone (exp. 5) operation were studied. In the latter case, the hot zone is 2 cm long and contains the spray nozzle, whereas the cold zone is 3 cm long. To fit the spray nozzle into the hot zone, it was moved by about 1 cm towards the front and installed against the front end plate. As shown in Figure 8.6c, a static disc is positioned inside the vortex chamber to separate the two zones when using the dual-zone design. Flow from the hot to the cold zone is only possible via a gap between the disc and the outer cylindrical wall of the vortex chamber. This allows preventing direct entrainment of droplets/particles into the chimney and forces the droplets to move to the cold zone via the hot zone.

The following procedure was followed. First air was fed to the vortex chamber until a steady state was achieved. Next, the milk solution was fed. When stopping the feeding of the milk solution, cold air was fed to the entire vortex chamber by means of a bypass installed on the

8.2. Spray drying of milk powder in vortex chamber: experimental proof of concept

Table 8.1: Vortex chamber design parameters and operating conditions.

		1	2	3	4	5	6	7	8	9
Experiment										
Chamber design	Diameter [cm]	24								
	Length [cm]	5								
	Number of slots	36								
	Slot size [mm]	0.5								
	Number of zones and lengths [cm]	2 zones: 2/3								
Gas inlet	Front end plate (hot zone)	1 zone: 5			2 zones: 2/3			2 zones: 3/2		
	Rear end plate (cold zone)	-			-			SD, CAI		
	Position (inside/outside the vortex chamber)	inside (axially non-central)			inside (hot zone)			SD, CAI SD, CAI SD, CAI, SO SO, CAI outside (hot zone)		
Operating conditions										
Air	Total flow rate [Nm ³ /h]	200	180	240	300	220/±70	120	250/±70	315	300/±70
	Feed temperature [°C] (zone1/zone2)				210			130	145	
Milk solution	Solids outlet (cold side) temperature if regulated [°C]				-					
	Injection pressure [bar]				0.5			1	0.5	
	Atomization pressure [bar]					4			6	
	Inlet temperature [°C]						ambient			
	Measured feed rate [kg/h]	5.4	5.04	3.6	3.96	6.264	5.148	3.996	1.908	3.12156
	Theoretical feed rate [kg/h]				6.6			9.3	6.6	
	Theoretical air atom. feed rate [kg/h]									
	Theoretical (Air atom. FR/Liq FR)				0.979545455	6.465		0.69516129	9.6975	
	Theoretical mean droplet size [μm]				25-30			30-40	1.469318182	
	Milk powder concentration [wt-%]				30				20-25	
	Feeding time [s]	300			20			180	40	300
SD = 140mm diameter freely spinning disc SO = solids outlet CAI = secondary cold air injection in the hot zone										

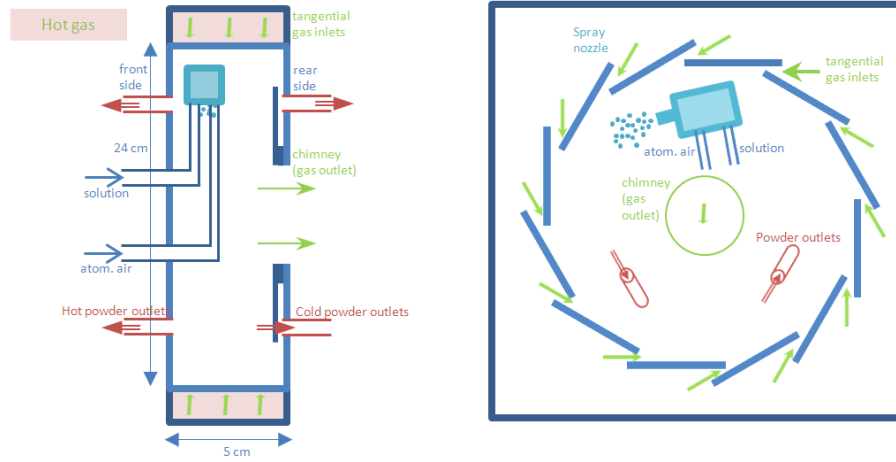
<i>Air inlet temperature [° C]</i>	80	130	200	300
Evaporation capacity [ml/s]				
CSTR	4	8.4	14.6	24.4
Co-current	3.1	7.6	14.1	24
Counter-current	4	8.4	14.6	24.4
Air outlet temperature [° C]				
CSTR	35.3	39.7	47.4	55.9
Co-current	46	48.3	53	59.1
Counter-current	35.4	39.2	47.4	55.9
Air relative humidity at the outlet [%]				
CSTR	66.1	93.3	99.5	99.99
Co-current	30.4	52.9	72.95	84.78
Counter-current	65.7	96.3	99.5	100
Air absolute humidity at the outlet [g_{H₂O}/kg_{dry air}]				
CSTR	24	44.4	73.7	119.7
Co-current	19.4	40.5	71	118
Counter-current	24	44.6	73.7	119.7
Air to water mass ratio [kg_{dry air}/kg_{water}]				
CSTR	52.7	25.4	14.5	8.72
Co-current	69.5	28.1	15.1	8.85
Counter-current	52.7	25.2	14.6	8.74

Table 8.2: Evaporation capacity of the vortex chamber at different air inlet temperature under different hydrodynamics assumptions. Air flow rate = 600 Nm³/h, inlet droplet size = 30 μm. For more details, see Annex A.

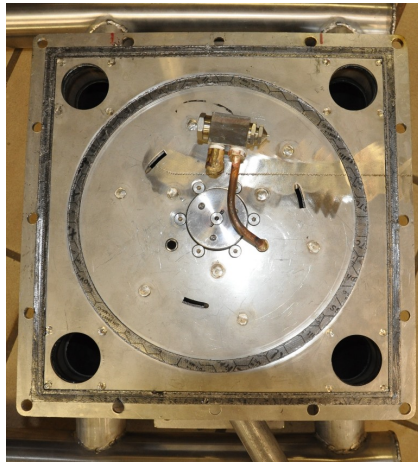
8.2. Spray drying of milk powder in vortex chamber: experimental proof of concept

air heater. In the first set of experiments, the solids outlets foreseen in the front and rear end plates of the vortex chamber were closed, forcing the particles or entrained droplets to leave the vortex chamber via the chimney. They are separated from the gas stream in a cyclone and filter.

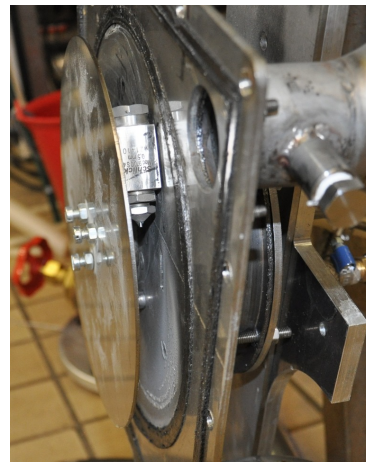
For a second set of experiments (exps. 6-9), the configuration was modified as shown in Figure 8.7. To avoid the generation of turbulence by the presence of the spray nozzle, the latter was located outside the vortex chamber. A perforation in one of the plates of the outer cylindrical wall of the vortex chamber allows feeding the droplets radially inwards resulting in a co-current flow of the air and the droplets. Dual-zone operation was applied. A separating disc was not installed as the droplets cannot by-pass the hot gas with this specific spray nozzle configuration. To prevent the liquid droplets sticking on the front wall, a 14 cm diameter freely rotating disc was installed against the front side end plate of the vortex chamber. In exps. 7-8-9, cold air is also injected from the center of the vortex chamber through the front and rear end plates. This secondary injection of cold air allows a rapid cooling of the particles in the hot zone of the vortex chamber. The rear side solids outlets were opened allowing the solids to leave the vortex chamber via these outlets or via the chimney. The solids leaving the chamber via the chimney are separated from the gas stream by means of a cyclone and filter. An additional pressure drop in the chimney outlet was introduced to control the distribution of the flow over the chimney outlet and the solids outlets. The experimental procedure was also modified. To prevent the spray from clogging, the atomizing air was fed from the beginning to the spray nozzle. The hot air was first injected at its pre-defined temperature while all the cold air inlets (including the cold air slots) were closed with a valve. Next, water was injected and evaporated in the chamber until a steady state temperature was obtained. At that time, the cold air valve was progressively opened to reach a pre-defined temperature at the solids outlet / chimney. Once this pre-defined outlet temperature reached, a valve was switched to start feeding the milk solution instead of water. As in previous experiments, cold air is fed to the entire vortex chamber after stopping the milk solution feeding.



(a)



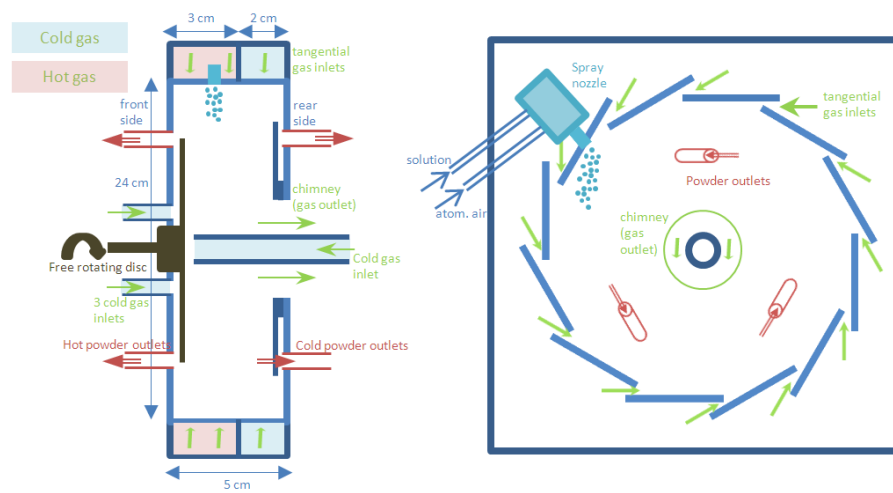
(b)



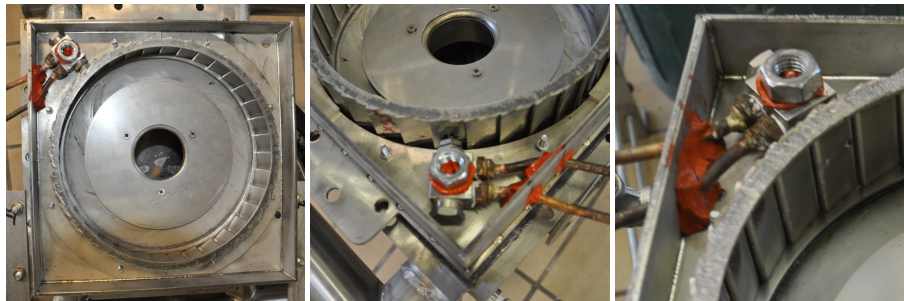
(c)

Figure 8.6: (a) Scheme and (b) pictures of the single-zone vortex chamber with an integrated spray nozzle. (c) Dual-zone operation with a static disc separating the two zones.

8.2. Spray drying of milk powder in vortex chamber: experimental proof of concept



(a)



(b)

Figure 8.7: (a) Scheme and (b) pictures of the modified set-up with the spray nozzle mounted outside the vortex chamber.

8.2.3 Experimental results and discussion

First set of experiments: In the 5 cm long vortex chamber, contacts between the droplets and the front end plate are difficult to avoid. As a result a milk paste is formed on the end plate. As seen in Figure 8.8 (exp 1), the accumulation of liquid on the front end plate can be very significant. Increasing the air inlet temperature and reducing the milk solution feeding time allow reducing liquid accumulation on the front end plate, as seen in Figure 8.8 (exp. 2: $T_{g,in} = 180^\circ\text{C}$, exp. 3: $T_{g,in} = 210^\circ\text{C}$, exp. 4: $T_{g,in} = 240^\circ\text{C}$). Increasing the air inlet temperature, however, also leads to a higher fraction of the milk being burned. In exps. 3 and 4, a small quantity of burned milk powder was found in the chimney and in the filter.

In order to cool down the powder and prevent it from burning, dual-zone operation was tried in exp. 5 imposing a comparable air injection temperature in the hot zone than in exp. 4. This implies the amount of energy supplied to the system per unit time is lower than in exp. 4. The liquid solution feed rate was, however, similar. Powder was recovered in the chimney and the filter and this powder was significantly less burned. Some more accumulation of the liquid milk solution in the hot zone was observed, suggesting that a higher gas injection temperature in the hot zone or lower liquid solution feed rates / droplet size is required. Burning the milk powder is then, of course, an issue. The increased milk accumulation on the front end wall is, however, also explained by the small length of the hot zone and the spray angle, giving a limited amount of time for a fraction of the droplets to dry before encountering the end plates of the hot zone, in particular the front end plate against which the spray nozzle is mounted. Turbulence contributes to the dispersion of the droplets. In this context, the integration of the spray nozzle inside the vortex chamber can be questioned. Indeed, the required tubing causes excessive turbulence and is detrimental for the formation of a free vortex.

Second set of experiments: To deal with the combination of above mentioned challenges, the set-up was modified as explained in Section 8.2.2. In exp. 6 the droplet size and liquid feed rate were comparable to previous experiments. As shown in Figure 8.9, the accumulation of liquid on the end plates was significantly reduced, despite a milk feeding time of 3 min instead of 20 s in the previous experiments. The milk deposition was also significantly less burned. No significant

8.2. *Spray drying of milk powder in vortex chamber: experimental proof of concept*

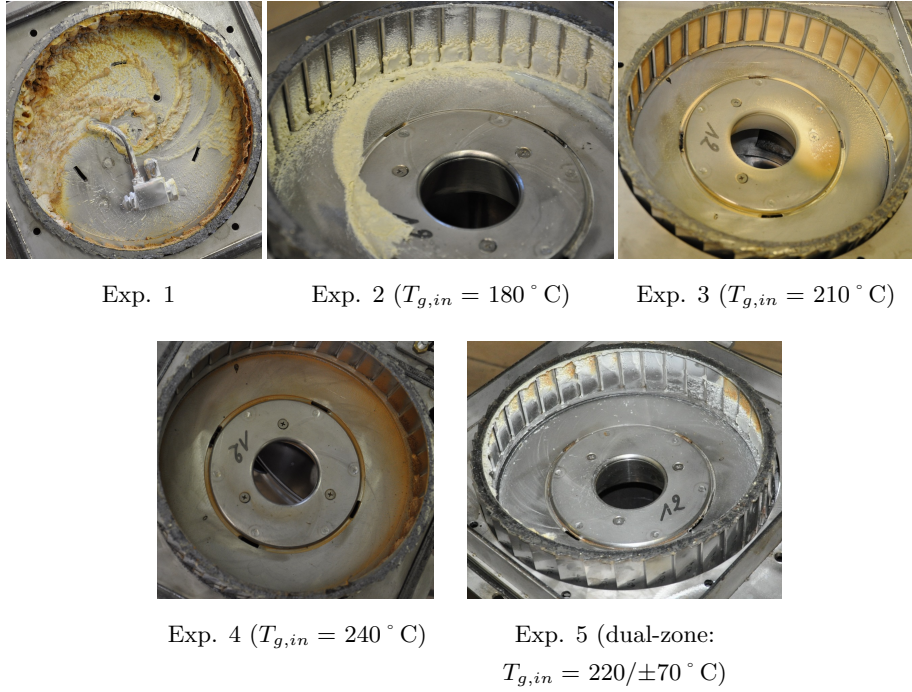


Figure 8.8: Pictures of the vortex chamber after spray drying experiments with the spray nozzle mounted inside the vortex chamber. Vortex chamber design and operating conditions: see Table 8.1.

amount of milk powder was, however, found. In exp. 7, the outlet temperature was increased to 130° , while increasing the droplet size. Again, the accumulation of liquid on the end plates was reduced, but somewhat more accumulation on the plates constituting the outer cylindrical wall of the chamber was observed. Therefore, in exp. 8 and 9, the air feed temperature and the solids outlet temperature were increased, the latter to 145° , and the liquid droplet size reduced by increasing the spray nozzle atomization pressure. In exp 8., liquid solution deposition on the end plates remained small, while the accumulation and deposition on the plates constituting the outer cylindrical wall of the vortex chamber were almost completely eliminated. In exp. 9, the higher feeding time resulted in more accumulation, both on the outer cylindrical wall and on the end plates. Nevertheless, in both exps. 8 and 9, a significant amount of fine milk powder was recovered from the chimney, cyclone and filter (Figure 8.10). This powder is not or only very slightly burned. Some powder was found downstream of the filter (Figure 8.10c) indicating that very fine particles are produced. A picture of the produced powder from exp. 8 is shown in Figure 8.11. SEM pictures of the produced powder from exp. 9 show that the powder consists in aggregates of spherical particles between 3 and 10 μm .

8.2.4 Conclusions

Continuous spray drying in a vortex chamber is possible, but requires a careful design and careful choice of the operating conditions. The latter have to be chosen as to avoid at the same time liquid accumulation in the vortex chamber and burning the produced powder. To achieve these goals, the spray dryer design has been optimized:

- Dual-zone operation allows injecting a hot gas in the zone where the liquid is injected while injecting cold gas in the zone where the final drying of the particles is carried out and from where the particles are evacuated.
- Secondary cold air injection in the hot zone allows controlling the burning of the particles produced in this zone.
- Positioning the spray nozzle outside the vortex chamber reduces turbulence and allows a co-current injection of droplets and gas in the vortex chamber. Deposition on the end walls can as such be minimized.

8.2. *Spray drying of milk powder in vortex chamber: experimental proof of concept*

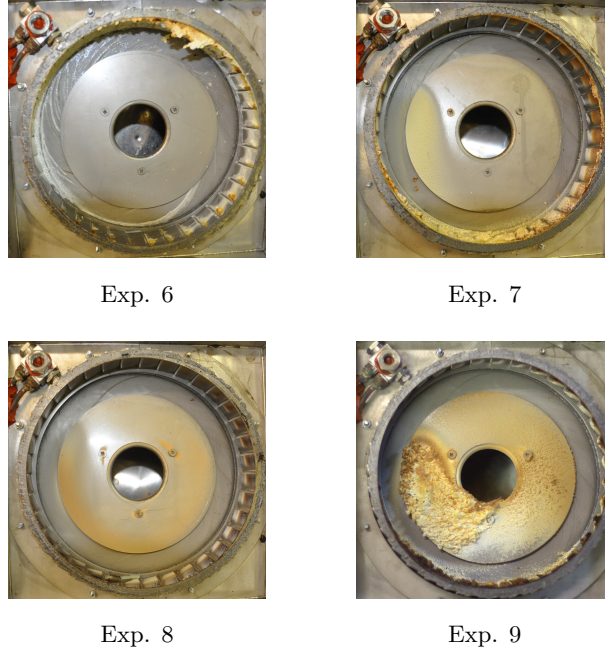


Figure 8.9: Pictures of the vortex chamber after spray drying experiments with the spray nozzle mounted outside the dual-zone vortex chamber. Vortex chamber design and operating conditions: see Table 8.1.

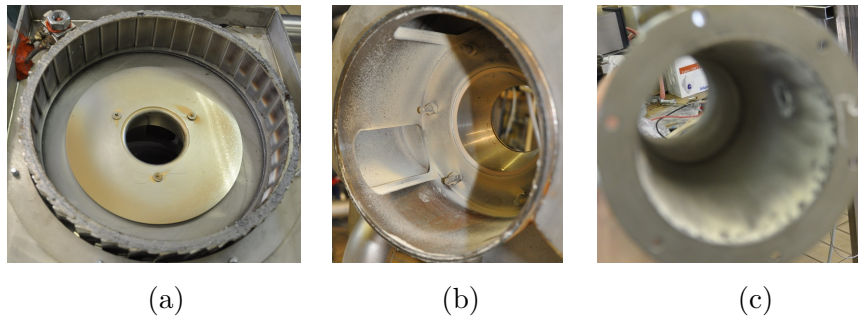


Figure 8.10: Pictures of the (a) vortex chamber, (b) cyclone, and (c) tube downstream of the filter after spray drying exp. 8 with the spray nozzle mounted outside the dual-zone vortex chamber. Vortex chamber design and operating conditions: see Table 8.1.



Figure 8.11: Powder produced in during spray drying exp. 8 with the spray nozzle mounted outside the dual-zone vortex chamber. Vortex chamber design and operating conditions: see Table 8.1.

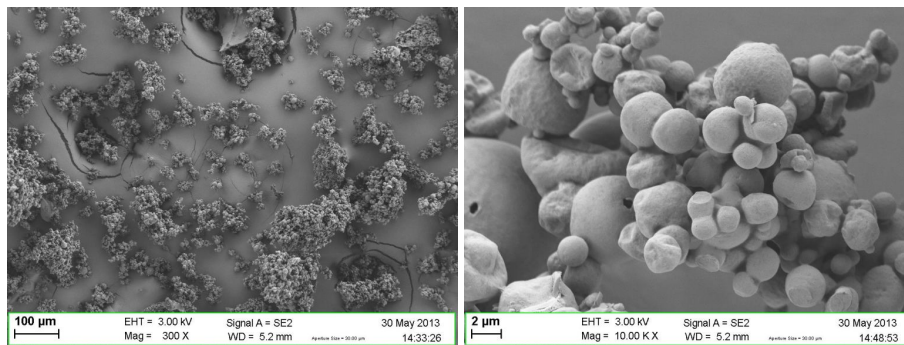


Figure 8.12: SEM pictures of the powder produced in during spray drying exp. 9 with the spray nozzle mounted outside the dual-zone vortex chamber. Vortex chamber design and operating conditions: see Table 8.1.

8.2. *Spray drying of milk powder in vortex chamber: experimental proof of concept*

- A freely spinning disc allows minimizing liquid accumulation on the front end plate (hot zone).

However, the spray drying chamber tested in this work is limited in length (5 cm). For the previously studied coating and granulation of powders, this was sufficient and necessary with the available air flow rates. For spray drying applications, integration of the spray nozzle and dual-zone operation is challenging.

8.2.5 Outlook

Based on the observations made in this work, some modifications with respect to the spray drying chamber design and operating conditions could be made to improve the spray drying process. First, to reduce liquid accumulation on the end plates, the length of the spray drying chamber could be increased. A connection piece to join two vortex chambers could allow experiments with a doubled vortex chamber length, i.e. 10 cm. Hence, the spray nozzle can be positioned at a somewhat higher distance from the front end plate. Moreover, cross-current operation can be considered in a longer vortex chamber, the liquid droplets being injected through the front end plate. The use of adapted vortex chambers is also crucial to reduce liquid accumulation on the outer cylindrical wall. In particular, the number of slots and their size have to result in an appropriate injection velocity and related centrifugal force for the given operating gas flow rate. Indeed, a too high or too low centrifugal force can lead to collisions between the droplets and the cylindrical wall as shown in Figure 8.13.

The configuration used in this study led to a poor thermal efficiency. Indeed, the main part of the air by-passes the droplets before leaving the reactor through the chimney. To use more efficiently the heat supplied with the air flow, the liquid solution has to be distributed more uniformly in the vortex chamber. That can be done by either using multiple spray nozzle distributed over the outer cylindrical wall or by using a rotary atomizer. The latter, however, implies the presence of a rotating wheel in the vortex chamber, which could be detrimental for the flow pattern.

Separately testing the influence of the different operating condition parameters is necessary. Separately controlling/measuring air flow rates

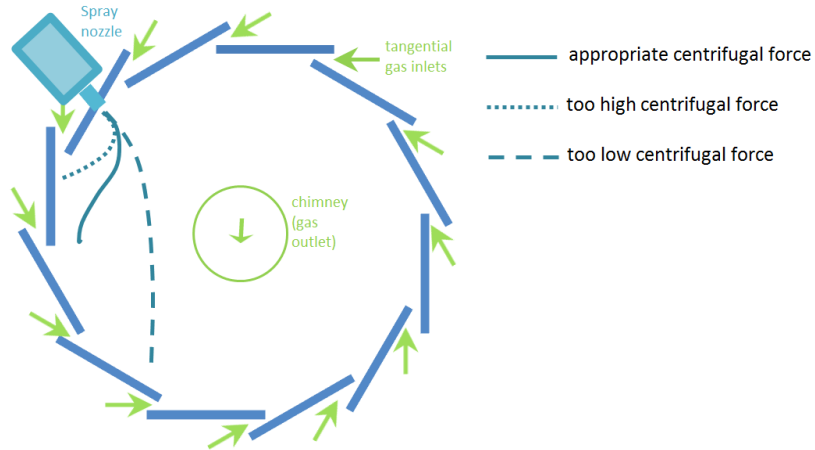


Figure 8.13: Droplets trajectory as a function of the centrifugal force.

injected in the cold and hot zones of the vortex chamber is then a must. This requires additional equipment to be installed. Finally, an optimization of the gas-solids separation is to be considered in order to recover easily the produced powder.

Conclusions

The aim of this thesis work was to demonstrate the intensification of mass, heat and momentum transfer in vortex chambers through the study of potential applications. Applications involving the drying of a liquid phase were chosen as their efficiencies are determined by mass and heat transfer rates. Biomass drying, coating of cohesive particles and spray drying in vortex chambers were experimentally studied. In addition to heat and mass transfer, momentum transfer was also found to be a key issue because of its influence on the hydrodynamics. In addition to the experimental work, a model for simulating biomass drying in vortex chambers was developed and validated.

The focus of Chapter 4 was the continuous drying of woody biomass in vortex chambers. The comparison of experimental results in both a conventional fluidized bed (FB) and a vortex chamber, i.e. a rotating fluidized bed in a static geometry (RFB-SG), allowed demonstrating a intensification factor of one order magnitude using the given vortex chamber in the range of operating conditions studied. This intensification factor was seen to arise from a combination of three factors: (i) higher heat and mass transfer rates, (ii) a densification and (iii) an improved uniformity of the particle bed. As expected, the advantage of the vortex chamber compared to the FB was more pronounced for the initial drying stage during which no intra-particle diffusion limitations occur. However, the cost related to the use of vortex chambers is the high air consumption. Nevertheless, it was shown that the air usage efficiency was similar in both the FB and the vortex chamber. This work also demonstrated that the product quality is not uniform because of the intense mixing and related CSTR behavior of the particle bed. The scale-up of the technology is therefore challenging.

In Chapter 5, modeling of biomass drying in vortex chamber was addressed. Model discrimination and estimation of the parameters were made based on batch drying experiment data. A good agreement between experimental and numerical data was obtained accounting for interfacial heat transfer and intra-particle diffusion limitations. Next, the scaling up of the technology was focused on. The beneficial ef-

fect of compartmenting the vortex chamber on the product uniformity was demonstrated. Furthermore, the latter was shown to provide an additional intensification of the drying process in the regime where intraparticle diffusion is limiting, i.e. when reaching low biomass moisture contents. To deal with the high air consumption related to the use of vortex chamber, air recycling was considered. The latter requires, however, re-heating the air between the different vortex chambers or compartments in order to achieve a reasonable outlet biomass moisture content. Dual chimney design allows taking advantage of both compartmenting and air recycling.

Chapter 7 presents the results of an experimental study of low temperature coating of cohesive particles in a vortex chamber. The characteristics of these particles makes them non-fluidizable in conventional fluidized beds because of the predominance of van der Waals forces. In vortex chamber, momentum transfer is intensified and, as a result, lighter and smaller particles can be fluidized compared to FBs. In the first part, the fluidization of whey cohesive particles in a vortex chamber is demonstrated while injecting or not a dispersed liquid phase. The successful fluidization of such particle requires, however sufficiently small gas inlet slots. In the second part of Chapter 7, the coating of these particles with maltodextrin is demonstrated. Intensified mass and heat transfer in vortex chambers allow obtaining a product with limited agglomeration despite the relatively low temperature and large droplet size. The unique flexibility of vortex chambers with respect to the gas flow rate was seen to allow controlling the agglomeration. Finally, SEM imaging of the produced particles and an active component release test revealed a good quality coating with only few cracks.

Finally, Chapter 8 is a proof of concept for spray drying in vortex chambers. The vortex chamber design as well as the operating conditions have to be carefully chosen such as avoiding burning the product and liquid accumulation in the vortex chamber. Within a certain range of operating conditions, successful spray drying of a 40w-% milk solution was demonstrated in a dual zone vortex chamber in which: (i) the spray nozzle was positioned outside the vortex chamber to avoid the disturbance of the flow pattern, (ii) hot air was injected in the zone including the spray nozzle and (iii) cold air was injected in the zone near the chimney to cool down the produced particles.

Conclusions

To conclude, the intensification of mass, heat and momentum transfer using vortex chambers was demonstrated by the experimental study of potential applications. Vortex chambers allow as such reducing the residence times and the development of smaller and safer units. Intensified reactors also allow new processing routes and severe operating conditions compared to the conventional technologies.

APPENDIX A

Calculation of a vortex chamber evaporation capacity

These calculations aim at estimating the evaporation capacity of water droplets in a given vortex chamber. The vaporisation rate and the related evaporation capacity depends on the droplet size, the air inlet temperature, the vortex chamber design and the air and water droplets flow rates.

A.1 Model

The water evaporation rate is determined by mass and heat transfer rates, which have to be coupled to an hydrodynamic model.

A.1.1 Mass Transfer

The molar flux of water at the gas-liquid interface can be written as

$$j = k_g(C_{int} - C) \quad (\text{A.1})$$

where k_g is the gas-liquid mass transfer coefficient and C_{int} and C are the water vapor concentration in the gas phase, respectively at the gas-liquid interface and in the bulk air. The gas-liquid mass transfer coefficient can be calculated through the Ranz-Marshall correlation [127]:

$$Sh = \frac{k_g d_l}{D_{w,g}} = 2 + 0.6 Re_l^{1/2} Sc^{1/3} \quad (\text{A.2})$$

with the droplet-based Reynolds number $Re_l = (u_{ig,r} - u_{il,r}) \rho_g d_l / \mu_g$. The air viscosity and the vapor diffusivity in air are calculated with the temperature using respectively Sutherland's [101] and Bolz & Tuve's [102]

equations. The air density is estimated from the ideal gas law.

The interfacial vapor concentration in the gas phase is assumed equal to the vapor saturation concentration. The latter depends on the interfacial temperature T_{int} according to

$$C_{int} = C_{sat}(T_{int}) = \frac{p_{sat}(T_{int})}{RT_{int}} \quad (\text{A.3})$$

$$p_{sat} = \exp \left[46.784 - \frac{6435}{T_{int}} - 3.868 \ln(T_{int}) \right] \cdot 133.322 \quad [\text{Pa}] \quad (\text{A.4})$$

A.1.2 Heat transfer

Assuming quasi-stationarity of heat transfer, the heat required for water evaporation is supplied by the interfacial heat transfer:

$$(-\Delta H_{vap})M_w j = h_f (T_{int} - T) \quad (\text{A.5})$$

with ΔH_{vap} the latent heat of vaporization, h_f the gas-liquid heat transfer coefficient and T_{int} and T the air temperature, respectively at the gas-liquid interface and in the bulk air. The heat transfer can be calculated by an analogous relation to equation (A.2) [127]:

$$Nu = \frac{h_f d_l}{\lambda_g} = 2 + 0.6 Re_l^{1/2} Pr^{1/3} \quad (\text{A.6})$$

A.1.3 Mass and heat balances

Mass and heat balances were calculated assuming either (i) a CSTR behavior of the droplets, (ii) co-current flow or (iii) counter-current flow between the air and the droplets. In each case, the gas phase is flowing radially inward. Assuming plug flow for the gas phase leads to the following mass and heat balances:

$$\begin{aligned} \frac{d(Q_g C)}{-dr} &= j \cdot a_v \varepsilon_w \Omega \\ Q_g C \left(r = \frac{D}{2} \right) &= Q_{g,in} C_{in} \end{aligned} \quad (\text{A.7})$$

$$\begin{aligned} \dot{m}_g c_{pg} \frac{dT}{-dr} &= h_f (T_{int} - T) \cdot a_v \varepsilon_w \Omega \\ T \left(r = \frac{D}{2} \right) &= T_{in} \end{aligned} \quad (\text{A.8})$$

with

- $\Omega = 2\pi r L$, the cross section "seen by the gas",
- Q_g and $\dot{m}_g$ the volumetric and mass flow rate of the gas phase,
- c_{pg} the heat capacity of the gas phase,
- $a_v = 6/d_l$ the mean surface to volume ratio of the droplets,
- ε_w the droplets volume fraction in the vortex chamber and
- T_{int} calculated from equation (A.5).

The droplets volume fraction is evaluated from the residence times and volumetric flow rates of the gas phase and the droplets:

$$\varepsilon_w = \frac{Q_w \tau_w}{Q_w \tau_w + Q_g \tau_g} \quad (\text{A.9})$$

Experimental data in the range of operating studied by the following simulations showed that droplets rotates at about 300 rpm in the vortex chamber. If the droplets are assumed to make one turn before leaving the chamber, a droplets residence time of 0.2 s is found. As $\varepsilon_w \ll 1$, the gas phase residence time τ_g can be determined from

$$\tau_g = \frac{V_r}{Q_g} \quad (\text{A.10})$$

(i) CSTR model

Assuming a perfect mixing of the droplets, the mass balance is given by

$$Q_{w,out} = Q_{w,in} - \frac{M_w}{\rho_w} \bar{a}_v \int_{V_r} \varepsilon_w j \, dV \quad (\text{A.11})$$

with $Q_{w,in}$ and $Q_{w,out}$ the water droplet flow rate at the inlet and outlet of the vortex chamber. The mean droplet surface to volume ratio

Appendix A: Calculation of a vortex chamber evaporation capacity

$\bar{a}_v$ is related to the mean droplet diameter. The latter is calculated via the following equation, introducing $\bar{Q}_w = (Q_{w,out} + Q_{w,in})/2$:

$$\bar{d}_l = \sqrt[3]{\frac{\bar{Q}_w}{\pi \dot{N}/6}} \quad (\text{A.12})$$

where $\dot{N}$ is the number of droplets fed per second

$$\dot{N} = \frac{Q_{w,in}}{\pi \bar{d}_{l,in}^3/6} \quad (\text{A.13})$$

$\bar{Q}_w$ is also used to determine ε_w .

Remark that for the CSTR model, the radial velocity of the droplets is assumed to be zero. Consequently, given that $\varepsilon_w \ll 1$:

$$Re_l \simeq \frac{(\dot{m}_g/\Omega) d_l}{\mu_g} \quad (\text{A.14})$$

(ii) Co-current flow model

In the co-current flow model, the droplets are injected at the outer cylindrical wall and flow radially inward with the air. Assuming plug flow, the mass balance for the droplets is written as

$$\begin{aligned} \frac{dQ_w}{-dr} &= -\frac{M_w}{\rho_w} j \cdot a_v \varepsilon_w \Omega \\ Q_w \left(r = \frac{D}{2} \right) &= Q_{w,in} \end{aligned} \quad (\text{A.15})$$

The droplet diameter profile is given by

$$d_l(r) = \sqrt[3]{\frac{Q_w(r)}{\pi \dot{N}/6}} \quad (\text{A.16})$$

The droplets interstitial radial velocity is related to the gas radial velocity by:

$$u_{il,r} = u_{ig,r} \frac{\tau_g}{\tau_w} \quad (\text{A.17})$$

Consequently, the droplet-based Reynolds number becomes

A.2. Input data

Vortex chamber diameter	$D = 0.24 \text{ m}$
Chimney diameter	$D_{\text{chimney}} = 0.12 \text{ m}$
Vortex chamber length	$L = 0.05 \text{ m}$
Vortex chamber volume	$V_r = 1.7 \cdot 10^{-3} \text{ m}^3$

Table A.1: Dimensions of the vortex chamber used in the simulations.

$$Re_l \simeq \frac{\left(\dot{m}_g / \Omega \left[1 - \frac{\tau_g}{\tau_w} \right] \right) d_l}{\mu_g} \quad (\text{A.18})$$

(iii) Counter-current flow model

In the counter-current flow model, the droplets are injected at the chimney and flow radially outward. Assuming plug flow, the mass balance for the droplets is written as

$$\begin{aligned} \frac{dQ_w}{dr} &= -\frac{M_w}{\rho_w} j \cdot a_v \varepsilon_w \Omega \\ Q_w \left(r = \frac{D_{\text{chimney}}}{2} \right) &= Q_{w,in} \end{aligned} \quad (\text{A.19})$$

The droplets interstitial radial velocity is related to the gas radial velocity by:

$$u_{il,r} = u_{ig,r} \frac{\tau_g}{\tau_w} \quad (\text{A.20})$$

Consequently, the droplet-based Reynolds number becomes

$$Re_l \simeq \frac{\left(\dot{m}_g / \Omega \left[1 + \frac{\tau_g}{\tau_w} \right] \right) d_l}{\mu_g} \quad (\text{A.21})$$

A.2 Input data

A.2.1 Vortex chamber dimensions

The vortex chamber dimensions used in the simulations are similar to those of the experimental set-up. They are summarized in Table A.1.

A.2.2 Operating conditions

The simulations were carried out with an air mass flow rate $\dot{m}_g$ of 0.2155 kg/s (corresponding to 600 Nm³/h) and an inlet droplet size $d_{l,in}$ of 30 μ m. The inlet air absolute humidity y_{in} was set to $5 \cdot 10^{-3}$ kg_w/kg_{dry air}. The relation between y_{in} and C_{in} is given by

$$C_{in} = \frac{p}{RT_{in}} \frac{y_{in}/M_w}{1/M_a + y_{in}/M_w} \quad (\text{A.22})$$

Simulations aim at estimating the evaporation capacity of the vortex chamber as defined in Table A.1 for different air inlet temperature and the different models. The evaporation capacity is the maximum water flow rate for which the droplets are completely evaporated before leaving the reactor.

A.3 Simulation Results

The results are shown in section 8.2.2.

Nomenclature

a_v	$\text{m}_{\text{int}}^2/\text{m}_{\text{w}}^3$	mean surface to volume ratio of the droplets
$c_{p,i}$	$\text{J}/(\text{kg}\cdot\text{K})$	specific heat of i
C	$\text{mol}_{\text{w}}/\text{m}_{\text{g}}^3$	water vapour concentration in the gas phase
D	m_{r}	vortex chamber diameter
D_{chimney}	m_{r}	chimney diameter
$D_{w,g}$	m^2/s	molecular diffusivity of water vapour in air
d_l	m_{w}	droplets diameter
ΔH_{vap}	$\text{J}/\text{kg}_{\text{w}}$	latent heat of vaporization
h_f	$\text{W}/(\text{m}_{\text{int}}^2\cdot\text{K})$	heat transfer coefficient
j	$\text{mol}_{\text{w}}/(\text{m}_{\text{int}}^2\cdot\text{s})$	water vapour molar flux at the gas-liquid interface
k_g	$\text{m}_{\text{g}}^3/(\text{m}_{\text{int}}^2\cdot\text{s})$	mass transfer coefficient
L	m_{r}	vortex chamber length
$\dot{m}_g$	kg/s	gas phase mass flow rate
M_i	kg/mol	molecular weight of i
$\dot{N}$	$1/\text{s}$	number of droplets fed per second
Nu	-	Nusselt number
p_{sat}	Pa	water vapour saturation partial pressure
p	Pa	total pressure
Pr	-	Prandtl number
Q_i	$\text{m}_{\text{i}}^3/\text{s}$	volumetric flow rate of i
r	m_{r}	spatial coordinate in the radial direction of the vortex chamber
R	$\text{J}/(\text{mol}\cdot\text{K})$	ideal gas law constant
Re_l	-	droplet-based Reynolds number
Sc	-	Schmidt number
Sh	-	Sherwood number
T	K	gas phase temperature
$u_{ij,r}$	$\text{m}_{\text{r}}/\text{s}$	radial interstitial velocity of j
V_r	m_{r}^3	vortex chamber volume
y	$\text{kg}_{\text{w}}/\text{kg}_{\text{dry air}}$	absolute air humidity

Greek letters

ε_w	$\text{m}_w^3/\text{m}_r^3$	water volumetric fraction in the vortex chamber
λ_i	$\text{W}/(\text{m.K})$	thermal conductivity of i
μ_i	Pa.s	viscosity of i
ρ_i	kg/m^3	density of i
τ_i	-	residence time of i in the vortex chamber
Ω	m_r^2	surface normal to the gas flux

Subscripts

a	air
g	gas phase
int	gas-liquid interface
in	at the inlet
l	droplets
r	reactor
sat	saturation
w	water

APPENDIX B

Commercial scale vortex chamber dryer simulations: supplementary results

In this annex are presented the simulations of C and D-design (without air reheating) commercial scale vortex chamber dryers.

Appendix B: Commercial scale vortex chamber dryer simulations:
supplementary results

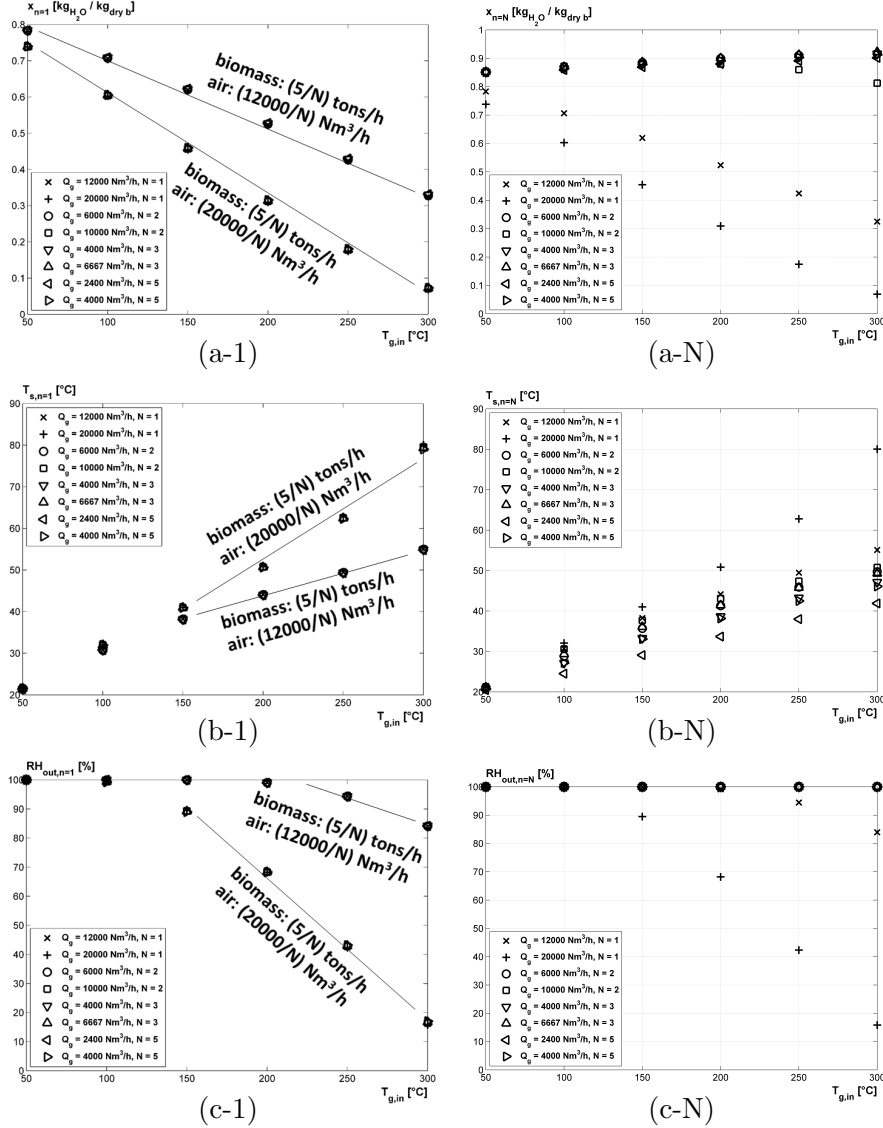


Figure B.1: Simulations of commercial-scale biomass drying for Design C with N equally sized parallel vortex chambers for different total air flow rates (design and other operating conditions: Table 5.2 and Figure 5.12). Biomass: (a) moisture content (dry basis) and (b) temperature; Air: (c) relative humidity at the outlet as a function of the gas inlet temperature. (...-1) indicates for the first chamber, (...-N) for the last. Simulations for 1, 3 and 5 vortex chambers.

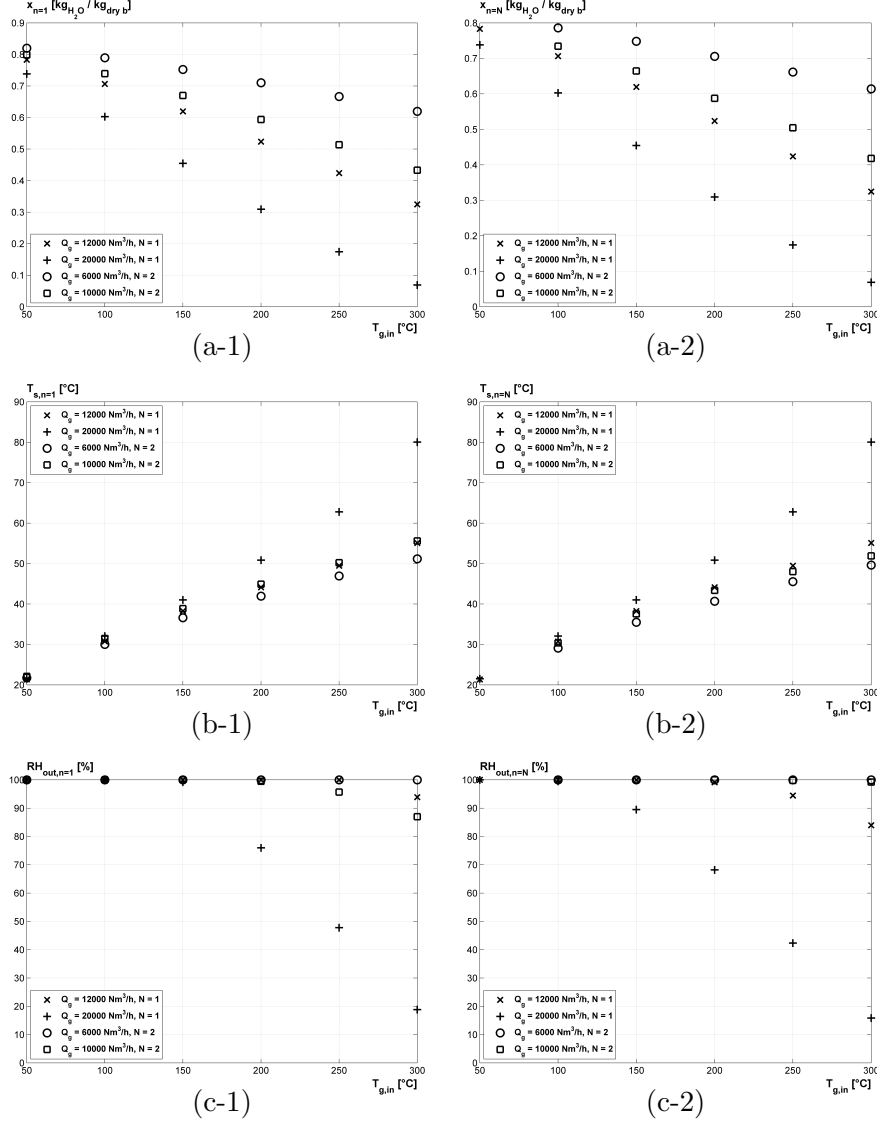


Figure B.2: Simulations of commercial-scale biomass drying in a Design D vortex chamber with 2 equally sized compartments for different total air flow rates (design and other operating conditions: Table 5.2 and Figure 5.12). Biomass: (a) moisture content (dry basis) and (b) temperature; Air: (c) relative humidity at the outlet as a function of the gas inlet temperature. (...-1) indicates for the first compartment, (...-2) for the second. Comparison with one-compartment vortex chamber shown.

*Appendix B: Commercial scale vortex chamber dryer simulations:
supplementary results*

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