Fundamentals of the high-shear pelletisation process
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Chapter 3

Down-scaling of the high-shear pelletisation process

Abstract

For the predictive modelling of the high-shear pelletisation process it is necessary to have a better understanding of the underlying mechanisms. Therefore, the high-shear pelletisation process was down-scaled and experiments were performed in a coffee grinder. As a reference, the Gral 10 was used for the so-called ‘large’-scale experiments. A toroidal flow-pattern was observed and described for both apparatus. There was no effect of an increased impeller speed on the velocity of pellets in the torus; only a change in angle with the tangential direction was observed. Moreover, the size of pellets decreased rapidly to a new value. This size increased again when the impeller speed was reduced to the initial (low) value. This suggested the existence of fast break-up and growth, resulting in a dynamic equilibrium depending on the impeller speed. Tracer experiments were used to investigate the turnover times of different sieve fractions. First order conversion rate constants for the pellets of different sieve fractions, calculated from colour concentrations at several processing times, showed to be equal for both apparatus when plotted against the dimensionless pellet diameter ($d/d_{50}$). These results are a strong indication for the use of small-scale experiments to retrieve a rapid insight in relevant mechanisms of growth of high-shear pelletisation.

This chapter is based on:
3.1. Introduction

High-shear pelletisation processes are well established in the pharmaceutical industry. Control of these processes is to a large extent still based on trial and error. What is needed, as recently mentioned by Ennis\(^1\), is a much better mechanistic understanding of agglomeration phenomena. Only then we will be able to master this process to such an extent that we are, for example, able to generalise the conditions that allow us to make narrow pellet size distributions as obtained with microcrystalline cellulose (MCC).

To establish real mechanistic models one has to take into account what happens at the scale of particles involved, and how this relates to transport processes, mechanical properties of excipients and operational variables like specific power input \((P/V)\), relative swept volume \((RSV)\), moisture content, etc.

High-shear pelletisation of mixtures containing MCC is an interesting process because very narrow particle size distributions can be obtained (80% within a range of 0.1 mm relative to an average value of 0.95 mm)\(^2\). Furthermore, a very typical toroidal flow pattern is present\(^2\). This leads to the question whether or not there is a relation between these two phenomena, and to what extent MCC does influence it. In the future, this should lead to conditions for high-shear pelletisation processes in which similar results (in particular a narrow particle size distributions) can be obtained with a broader range of excipients than MCC alone.

Because it always will be difficult to predict the behaviour of high-shear pelletisation entirely from first principles (particularly for the influence of the mechanical and surface properties of the materials involved), a combination of a mechanistic approach with small-scale experiments that are representative for production-scale conditions is necessary. In our case, a high-shear coffee grinder was used for small-scale experiments.

3.2. Theory

3.2.1. Previous work

Growth of pellets can be divided into four stages: initial nucleation with minor growth (which can be characterised by the formation of primary nuclei and the subsequent breakage into secondary nuclei), exponential growth by coalescence, linear growth, and equilibrium with zero net growth (chapter 2). During kneading, the mean pellet size increases until a dynamic equilibrium is reached in which zero net growth occurs. The rates of growth and break-up of pellets become equal. In other words, although the mean pellet diameter is not changing anymore, growth and break-up of individual pellets still occur. The time of onset on the equilibrium depends on many variables like moisture content, relative swept volume and liquid distribution.

The relative swept volume, \(RSV \ (s^{-1})\), is defined as the fractional volume of the particles which is displaced by the impeller per second, and given by:
Here, $D$ is the diameter of the bowl, $h_\perp$ the projected height of the impeller, $n_{\text{arm}}$ the number of impeller arms, $\Phi_{\text{fill}}$ the filling degree of the bowl, $V$ the total bowl volume, and $N$ the impeller rotational speed. Because $V \sim D^3$, and $h_\perp \sim D$, the relative swept volume is proportional to the impeller rotating speed, $N$.

3.2.2. Critical Froude number

Dimensionless numbers, like the Froude number, can be very useful to investigate scale-up issues\(^3\). Two forces acting on the pellets are of major importance for the flowing profile of the pellets. These two forces are the centrifugal force ($mv^2/R$) and the gravitational force ($mg$), where $m$ is the mass of one pellet, $g$ the gravitational acceleration constant, $v$ the velocity of the pellet calculated as a fraction of the impeller tip velocity ($e\pi ND$), $R$ the radius of the bowl (½$D$), $D$ the diameter of the bowl, and $N$ the impeller rotational speed. The relationship between the centrifugal force and the gravitational force is given by the Froude number:

$$Fr = \frac{N^2D}{g}$$

$$Fr_c = \frac{mv^2/R}{mg} = \left(\frac{e\pi ND}{\sqrt{1/D}}\right)^2 \frac{1}{2g} = 2\pi^2 e^2 \frac{N^2D}{g}$$

where $Fr$ is the Froude number, and $Fr_c$ is the critical Froude number, and $e$ is the coefficient of restitution based on linear velocity differences (as will be discussed in greater depth in chapter 4).

Above a critical Froude number of 1, the centrifugal force exceeds the gravitational force. At critical Froude number of 1, the centrifugal force and the gravitational force are equal. Above a critical Froude number of 1, the pellets are moving in a more fluidised pattern (e.g. torus profile), due to the higher centrifugal force acting on the pellets.

3.2.3. Impeller as a rotating disk

The impeller of the high-shear mixer like the coffee grinder acts as a rotating disk. This can be illustrated by the next comparison. The time one pellet (with a mean diameter of 0.5 mm) requires to fall along the side of the impeller (height of the impeller is 0.5 mm) is about 0.014 seconds:

$$d_p + h_\perp = \frac{1}{2} gt_{\text{fall}}^2 \Rightarrow t_{\text{fall}} = \sqrt{\frac{d_p + h_\perp}{\frac{1}{2} g}} = 0.014 \text{ s}$$

The time between the two passages ($t_{\text{pass}}$) of the both impeller arms is, at an impeller speed of 82 rps, is 0.006 seconds (figure 3.1):
\[ t_{\text{pass}} = \frac{1}{n_{\text{blades}} N} \]  

(3.5)

The same calculations can be made for the Gral: the time a pellet requires to fall along the impeller is \((h_\perp = 2 \text{ cm})\) 0.06 seconds. The time between two passage of the impeller arm (three arms in total, 450 rpm) is 0.04 seconds. This means that the impeller arm of the Gral also acts largely as a rotating disk.

This means, for most pellets, that the time between two passages of the impeller arm is too short for the pellet to fall along the impeller. So, at an impeller speed of 82 rps, the impeller arm of the coffee grinder acts as a rotating disk.

The critical velocity of the impeller to act as a rotating disk occurs at \(t_{\text{fall}} = t_{\text{pass}}\). This velocity is equal to 35 rps for the coffee grinder, and 5 rps (=300 rpm) for the Gral 10 (see figure 3.1).

![Figure 3.1. Comparison between the \((-\rightarrow) t_{\text{pass}}\) and \((-\cdots) t_{\text{fall}}\) of a pellet in the coffee grinder.](image)

Taking the relative swept volume as a measure for the process intensity is somewhat questionable. Because an increase of the relative swept volume should result in a higher process intensity. This is not always valid, because the impeller can sometimes be considered as a rotating disk, because the time between two passages of the impeller is too short for pellets to fall along the impeller. This results in a pellet-impeller collision just before the pellet is able to fall along the impeller and reach the bottom of the bowl.

Above the critical velocities of the impeller arm, the relative swept volume is not a valid measure for the process intensity, because the velocities of the pellets are supposed to be less influenced by the impeller speed as expected based on the relative swept volume.

The aim of this study is to show the advantages of the coffee grinder as a down-scaled high-shear mixer, which is representative for the large-scale production. This will be done by elucidating the flow pattern of pellets and a number of mechanisms involved in growth and break-up of pellets in a high-shear pelletisation process.
3.3. Materials and methods

3.3.1. Materials

Mixtures with equal amounts of microcrystalline cellulose (Pharmacel PH101, DMV, Veghel, The Netherlands) and \(\alpha\)-lactose (Pharmatose 200 mesh, DMV, Veghel, The Netherlands) were used as starting materials. Amaranth, not having any effect on pellet growth, was added as colouring agent during the tracer experiments (1 g/l binder liquid for the coffee grinder experiment, and 2 g/l binder liquid for the Gral experiment). Demineralised water was used as binder liquid.

3.3.2. Equipment

A Moulinex coffee grinder (Moulinex 980, Ireland) was used for small-scale experiments. The impeller speed was made continuously adjustable until 300 rps using a transformer. A Gral 10 high-shear mixer (Machines Collette, Wommelgem, Belgium) was used for the preparation of pellets. The impeller speed of the Gral was continuously adjustable until 660 rpm, the chopper speed could be varied between 0, 1500, and 3000 rpm. The temperature of water in the double-sided wall of the bowl was 25°C during the entire process. Binder liquid was added under gravity. The mean diameter of the liquid drops was 5 mm.

The geometry of both apparatus and some other relevant properties are given in table 3.1, whereas the relative swept volume is calculated according to equation 3.1.

<table>
<thead>
<tr>
<th></th>
<th>Gral 10</th>
<th>coffee grinder</th>
</tr>
</thead>
<tbody>
<tr>
<td>volume bowl (l)</td>
<td>8</td>
<td>0.250</td>
</tr>
<tr>
<td>diameter bowl (cm)</td>
<td>24.5</td>
<td>8</td>
</tr>
<tr>
<td>impeller radius (cm)</td>
<td>11.5</td>
<td>3.3</td>
</tr>
<tr>
<td>impeller height (h_\perp) (cm)</td>
<td>2</td>
<td>0.1</td>
</tr>
<tr>
<td>impeller speed (rps)</td>
<td>6.7</td>
<td>82</td>
</tr>
<tr>
<td>batch size (g)</td>
<td>800</td>
<td>24</td>
</tr>
<tr>
<td>binder liquid (ml)</td>
<td>400</td>
<td>12</td>
</tr>
<tr>
<td>relative swept volume (s^{-1})</td>
<td>12</td>
<td>14</td>
</tr>
</tbody>
</table>

The sieve analyses were carried out with a set of sieves with openings of 150, 212, 300, 425, 600, 850, 1080, 1180, 1400, and 2000 \(\mu\)m. An UV/VIS-spectrophotometer (Philips PU8720, England) was used for colour concentration measurements during the tracer experiments.
3.3.3. Methods

Photos were taken from the pellets in the torus during the dynamic equilibrium stage. A stroboscope was used for the illumination of the process. The frequency of the stroboscope was adjusted to obtain two or more images of a pellet at one photo. The distance and the deviation of the pellets from the tangential line (= angle) were measured. The radial velocity ($v_r$), the tangential velocity ($v_t$), and the total velocity ($v$) were calculated from these data (see figure 3.2).

![Diagram of velocities](image)

*Figure 3.2. Measurement of the velocities at the top of the torus. $v_r$ is radial velocity; $v_t$ is tangential velocity and $v$ is total velocity.*

Preparation of pellets in the Gral 10 was as follows: after 5 minutes of premixing, a precisely determined amount of binder liquid was added under gravity in 1 minute, and the mass was kneaded for four minutes. Subsequently, the wall-addition was scraped from the wall, and the mass was kneaded for another 15 minutes. The total processing time was 25 minutes (including 5 minutes premixing). During the whole experiment, the impeller speed was set at 400 rpm. The chopper speed was set at 3000 rpm during the liquid addition and the first four minutes of the kneading stage; the chopper was turned-off during the last 15 minutes of kneading stage. All pellets were tray-dried at 60°C for 24 hours.

In the coffee grinder, pellets were prepared in a similar way, but the time-scales were shortened to 30 seconds of premixing, 30 seconds of liquid addition and 120 seconds of wet massing. The impeller speed was set at 82 rps and at 187 rps for two different experiments. The liquid (12 ml) was added using a syringe.

During the tracer experiments five percent (Gral experiment) or ten percent (coffee grinder experiments) of the pellets was replaced by freshly prepared wet coloured pellets. All coloured pellets used in the tracer experiments, were prepared in the coffee grinder at a higher impeller speed (187 rps). The concentration of amaranth in each sieve fraction was measured at several processing times. After dissolving 300 mg of pellets in 3 ml water, the colour concentrations were measured spectrophotometrically at 522.5 nm.
3.4. **Results and Discussion**

3.4.1. **Flow profile**

The flow profile of the pellets, during the dynamic equilibrium, was investigated by measuring the displacement of pellets photographically. A torus of pellets was observed in the coffee grinder (see figure 3.3) and the Gral. This torus can be characterised by the helical shape\(^2,4\). The pellet-pellet, pellet-impeller, and pellet-wall collisions are responsible for the fact that all pellets stay in this stable torus: even without the cover, the particles stay in the coffee grinder. No such a stable flow pattern was observed for dry pellets. Most likely, this flow pattern is the result of visco-elastic properties of microcrystalline cellulose (chapter 4).

![Figure 3.3](image)

*Figure 3.3. Photo of a part of the top-view of the torus in a coffee grinder. The arrows indicate some of the traceable pellets.*

The total velocity of pellets in the torus in a coffee grinder is shown in figure 3.4. The velocity of pellets was lowest near the wall of the bowl. Further from the wall, the total velocity of the pellets is increased. No pellets are observed in the centre of the mixer; the bottom of the bowl could be seen clearly (see figure 3.3).

Surprisingly, no differences in velocities of the pellets (at the surface of the torus) were measured at a higher impeller speed. The angle (tangential difference) of the pellets is shown in figure 3.5. Near the wall, the direction of the pellets was almost equal to the tangential line. Near the centre, the angle increases and the direction of the pellets was more inside (radial) orientated. At a higher impeller speed the differences in the tangential direction became lower, while the velocity of the pellets remained unchanged. Only the velocities of the pellets at the upper side of the torus were measured, because no method was available to measure the velocity of pellets at the bottom or in the torus. Minor changes were observed in the height
and width of the torus by different impeller speeds.
As a consequence of the higher impeller speed more energy was dissipated. Because no change in pellet velocity was observed, more energy dissipation should lead to faster growth and destructions of all pellets, which will be more emphasised in the next part about the tracer experiments.
The torus showed to be very stable. Disturbances due to the turned-off chopper (or little obstacles) only produced local changes in the direction of the pellets, and minor changes in velocity. The changes in the directions of the pellets caused more collisions between pellets. So, the turnover of the pellet mass will be faster due to the action of the chopper.
3.4.2. Pellet diameter

The pellet diameter, measured as pellet size distribution and volumetric mean diameter \((d_{50})\), depends on several process variables like moisture content, impeller speed and processing time. The influence of the impeller speed on the final pellet size distribution is shown in figure 3.6. Several authors have shown the effect of the impeller speed on the pellet size distribution\(^5\)\(^-\)\(^7\). In contrast to their results, this figure shows a decrease of pellet diameter at an increased impeller speed. We also found a dynamic equilibrium in growth, which has not been found by these authors. A combination of a high impeller speed during the process and a low filling degree of the bowl, resulting in a high relative swept volume probably cause these two differences.

![Figure 3.6](image)

Figure 3.6. Cumulative pellet size distributions at several impeller speeds for experiments performed in the coffee grinder. The impeller speed was set at \((\bigdiamond) 20\) rps; \((\square) 82\) rps; \((\triangle) 138\) rps; \((\times) 187\) rps; and \((+) 354\) rps.

The shift of the pellet size distribution towards a smaller mean diameter at an increased impeller speed can be explained by the higher destructive forces (higher tip-velocity) of the impeller. This causes two effects: first, the number of collisions is increased, and second the chance of successful collisions for coalescence is decreased, because of the low contact-time between the pellets. Break-up of pellets becomes more important compared to growth of pellets. In other words, the equilibrium between growth and break-up shifts towards a higher break-up, which leads to a lower mean pellet size.

The mean pellet size can be controlled by the impeller speed. After preparation of a batch of pellets in the coffee grinder at 82 rps, the impeller speed was increased to 187 rps. This resulted in a 119 \(\mu m\) decrease of the mean pellet diameter \((d_{50})\). A decrease of the impeller speed to the starting value (82 rps) resulted in an increase of the mean pellet size with 188 \(\mu m\). This indicated the existence of a dynamic equilibrium between growth and break-up of
pellets. Furthermore, the effect of a changed impeller speed on the mean pellet size is partly reversible. Therefore, the impeller speed can be used to control the mean pellet size.

The linear relationship between the impeller speed of the coffee grinder and the mean pellet diameter ($d_{50}$) is shown in figure 3.7. This relationship is levelled-off at impeller speeds above 150 rps, but still remains linear. The transition-point of the relationship between the impeller rotational speed and the mean pellet size at 150 rps, may be caused by the more turbulent flow at higher impeller speeds. The critical Froude number, calculated with the mean pellet velocity at the whole torus (using $e = 0.3$, see chapter 4), exceeds the critical value of 1 at an impeller speed of 70 rps theoretically (see equation 3.3). This can not explain the practically observed transition-point at 150 rps in figure 3.7.

A better explanation can be found if we take a more careful look at the impeller, and consider the concept of the impeller as a rotating disk. As said before (section 3.2.3), the critical impeller speed of the coffee grinder at which the impeller can be considered as a rotating disk is 35 rps, which not exactly corresponds to the transition-point of 150 rps as given in figure 3.7. This can partly be explained as follows.

The impeller rotational speed of the coffee grinder has to be controlled with a transformer, and has been measured as an unloaded impeller speed (e.g. without any material in the bowl). The loaded impeller speed, thus during processing, is actually much lower: less than half of the unloaded impeller speed. The loaded impeller speed, however, changes continuously during pelletisation, and therefore is not a constant representative value. The relationship between a loaded impeller speed and the mean pellet size is also given in figure 3.7 (open symbols). Using the loaded impeller speed, the impeller speed at the transition-point corresponds to the moment at which the impeller starts to act as a rotating disk. Thus at

Figure 3.7. Mean pellet diameter ($d_{50}$) as a function of the impeller speed for the coffee grinder experiments; (●) unloaded impeller speed, (◇) loaded impeller speed.
unloaded impeller speeds (corresponding to a loaded impeller speed of about 35 rps) above 150 rps, the effect of the impeller speed on the mean pellet size is less pronounced, which can be explained by the rotating-disk-action of the impeller.

In spite of the large differences between the unloaded and the loaded impeller speed of the coffee grinder (this difference has not been found for the Gral), we decided to represent all coffee grinder data with the unloaded impeller speeds. The main reason is that the unloaded impeller speed is a constant value and the loaded impeller speed is a variable value during mixing depending on the consistency of the wet mass.

3.4.3. Tracer experiments

During equilibrium, the pellet size distribution was narrow and did not change with time. Nevertheless, growth and break-up of pellets still occurred due to the interactions between pellets in the torus, and between the impeller (or wall) and the pellets. This can be explained by equal growth and break-up velocities, which result in zero net growth and a dynamic equilibrium. To prove this mechanism and to elucidate the rate at which growth and break-up occurs, experiments with coloured small pellets were performed. Figures 3.8 and 3.9 give the colour concentration of the different sieve fractions at several processing times for the coffee grinder and the Gral, respectively.

![Graph showing colour concentration for different sieve fractions at different processing times after colouring the small pellets in the coffee grinder at 82 rps](image)

*Figure 3.8. Colour concentration for different sieve fractions at different processing times after colouring the small pellets in the coffee grinder at 82 rps. Sieve fractions (●) 0-150 µm; (•) 150-212 µm; (◇) 212-300 µm; (★) 300-425 µm; (x) 425-600 µm; (△) 600-850 µm; (□) 850 µm-1.18 mm; (◇) 1.18-2 mm. The lines are the models according to eq. 3.6.*

Figures 3.8 and 3.9 show a fast exponential decay of the colour concentration of the smallest sieve fractions, and a fast increasing colour concentration of the largest sieve fractions. All concentrations converged to a constant value. The colour concentration-curve for the Gral experiment was similarly shaped, only the time-scales were prolonged. The colour
distribution changed within thirty seconds in the coffee grinder-experiment (and within sixty seconds for the Gral-experiment) from a heterogeneous state (only the smallest pellets were coloured) into a homogeneous state (all pellets were equally coloured). This indicated total turnover of the whole pellet population within thirty seconds for the coffee grinder experiment and sixty seconds for the Gral experiment.

Figure 3.9. Colour concentration for different sieve fractions at different processing times after colouring the small pellets in the Gral. Sieve fractions (■) 0-150 µm; (+) 150-212 µm; (○) 212-300 µm; (★) 300-425 µm; (▲) 425-600 µm; (△) 600-850 µm; (□) 850 µm-1.18 mm; (◇) 1.18-2 mm. The lines are the models according to eq. 3.6.

For each sieve fraction, the colour concentration as a function of time could be described with:

\[
\frac{C - C_\infty}{C_0 - C_\infty} = \exp[-k \cdot t]
\]  

(3.6)

where \( C \) is the colour concentration (mg/g), \( k \) is the conversion rate constant (s\(^{-1}\)) and \( t \) is the processing time (s). Figure 3.10 shows the conversion rate constants for the small- and large-scale experiments as a function of the dimensionless pellet diameter \((d/d_{50})\), whereas \( d_{50} \) is volume mean diameter of the whole pellet mass, and \( d \) is the mean diameter of one sieve fraction.

First, the Gral experiment and the 82 rps coffee grinder experiment are discussed. The conversion rate constants of the smallest pellets were five to seven times higher compared to the larger ones. This indicated fast growth of small pellets, and a fast formation of small pellets by break-up of larger pellets. The conversion rate constants of the larger pellets \((>d_{50})\) were almost constant. This indicated a similar mechanism of growth and break-up for those pellets at both scales.
Because of the different geometry of both apparatus it was not possible to prepare pellets with exactly the same pellet size distribution. The conversion rate constant of the pellets however, showed to be independent of the scale of operation. So in this respect, both scales can be compared to each other.

An additional tracer experiment was performed in the coffee grinder at a higher impeller speed (187 rps). The colour concentration distributed similarly, only the time until homogeneous colour distribution was shortened from 30 seconds (at an impeller speed of 82 rps) to 7 seconds (at 187 rps). The conversion rate constants for the sieve fractions of larger particles (figure 3.10) were almost ten times higher, compared to the lower impeller speed experiment. This corresponds to the ratio of the power numbers for both processes in the coffee grinder:

\[ P \propto N^3 D^5 \]

\[ \frac{P_{\text{high}}}{P_{\text{low}}} \propto \left( \frac{187}{82} \right)^3 \approx 10 \]  

(3.7)

The velocities of pellets in the torus of the coffee grinder (figure 3.4) were almost the same at 82 rps and 187 rps. Additional power input at a higher impeller speed caused faster break-up of pellets. Although the velocity of the impeller was increased with a factor 2.3, the measured velocities of the pellets at the top of the torus remained unchanged. This indicated larger differences in velocity between pellets and the impeller. So, higher impacts between impeller and pellets occurred, which caused higher conversion rate constants for higher impeller speed experiments and therefore faster onset of equilibrium.
3.5. Conclusions

The flow profile of pellets during the equilibrium stage was characterised as a torus. As soon as pellets were formed this torus was seen in the Gral and the coffee grinder. The velocity of pellets at the surface of the torus was equal at different impeller speeds; the difference of the trajectory of the pellets from the tangential line was increased at a higher impeller speed. Because the velocity of the pellets at the top of the torus did not change at higher impeller speeds, the increased differences in velocity between the impeller and pellets resulted in more impact and higher conversion rate constants due to faster growth and breakage.

An increased impeller speed resulted in a decrease of the mean pellet size. This effect showed to be partly reversible. Therefore, the impeller speed can be used to control the mean pellet size during high-shear pelletisation.

Comparison of the time needed for a pellet to fall along the impeller arm, with the time between two passages of the impeller arm, showed that the impeller hits the pellets before reaching the bottom of the bowl. From these results is was concluded that the impeller arms of the high-shear mixer acts as a rotating disk (like in the centrifugal equipment).

Growth and destruction of pellets from different sieve fractions could be measured with tracer experiments. Conversion rate constants were derived from the exponential decay of the colour concentration at different processing times for each sieve fraction. Compared to the dimensionless diameter, similar conversion rate constants were found in the coffee grinder and the Gral. The conversion rate constants of the smallest pellets were higher (compared to the larger ones), which indicated faster growth of the smaller pellets due to the destruction of the large pellets.

At a higher impeller speed, the velocity of pellets at the upper side of the torus was unchanged. The velocity differences between impeller and pellets were increased, which resulted in faster turnover of the whole pellet mass. The conversion rate constants at 187 rps were about ten times higher compared to those found for larger pellets at 82 rps. This ratio is equal to the third power of the ratio between impeller speeds, suggesting that $P/V$ is important.

The fact that similar phenomena are observed in coffee grinder and Gral suggested that experiments with a coffee grinder can be a cheap and rapid method to elucidate relevant mechanisms in high-shear pelletisation.
3.6. Nomenclature

\( C \)  \( \) colour concentration (mg/g)
\( d \) \( d \)  diameter pellet (m)
\( D \)  \( D \)  diameter bowl (m)
\( e \) \( e \)  coefficient of restitution based on linear velocity differences (-)
\( Fr \) \( Fr \)  Froude number (-)
\( g \) \( g \)  gravitational acceleration constant (m/s\(^2\))
\( h_\perp \) \( h_\perp \)  projected height of the impeller (m)
\( k \) \( k \)  conversion rate constant (1/s)
\( m \) \( m \)  pellet mass (kg)
\( n_{arm} \) \( n_{arm} \)  number of impeller arms
\( N \) \( N \)  impeller rotational speed (1/s)
\( P \) \( P \)  power (Watt)
\( R \) \( R \)  radius bowl (m)
\( t \) \( t \)  time (s)
\( v \) \( v \)  velocity (m/s)
\( V \) \( V \)  bowl volume (m\(^3\))

Greek symbols
\( \Phi_{\text{fill}} \) \( \Phi_{\text{fill}} \)  filling degree of the bowl (-)

Subscripts
\( c \) \( c \)  critical
\( p \) \( p \)  pellet
\( r \) \( r \)  radial
\( t \) \( t \)  tangential
3.7. References