Cross-flow deep fat frying and its effect on fry quality distribution and mobility

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Abstract Conventional industrial frying systems are not optimised towards homogeneous product quality, which is partly related to poor oil distribution across the packed bed of fries. In this study we investigate an alternative frying system with an oil cross-flow from bottom to top through a packed bed of fries. Fluidization of rectangular fries during frying was characterised with a modified Ergun equation. Mixing was visualized by using two coloured layers of fries and quantified in terms of mixing entropy. Smaller fries mixed quickly during frying, while longer fries exhibited much less mixing, which was attributed to the higher minimum fluidization velocity and slower dehydration for longer fries. The cross-flow velocity was found an important parameter for the homogeneity of the moisture content of fries. Increased oil velocities positively affected moisture distribution due to a higher oil refresh rate. However, inducing fluidization caused the moisture distribution to become unpredictable due to bed instabilities.

Keywords French fries · Frying · Quality distribution · Mixing · Fluidization · Rectangular particles

Introduction

Deep fat frying is one of the more important unit operations in the food processing industry. As for other technologies, also for frying technology there is a continuous need for innovation in terms of process efficiency and product quality (Blumenthal and Stier 1991). The fried product with the highest market share is without doubts French fries with a worldwide production of 11 million tonnes in 2003 (Huffaker 2003), a number expected to increase with the increase in world population and spreading of western culture.

One of the major challenges with large scale frying systems is to obtain products of consistent quality. Variations in raw materials, but also frying conditions are sources of quality variation (Stier 2004). For state-of-the-art industrial frying operations, especially the oil flow conditions are not well defined. This may be obvious since industrial fryers work at high production capacity and hot oil is continuously injected, while the cooled oil is removed for reheating. Depending on the exact design details, variations in frying conditions may occur as a result of the characteristic flow of the oil, which can affect the local temperature, convective heat transfer and the residence time in the frying unit. Inefficient oil flow conditions can give rise to an undesired distribution of final fry quality.

The best way to ensure homogenous heat transfer from the frying oil to the fries would be to subject each fry to exactly the same amount of oil with a steady temperature. If the fries are transported horizontally, then the most appropriate way to do this is by applying a vertical oil flow. This is analogous to continuous fluidized bed drying. Fluid bed drying can be used for the uniform drying of powders with high heat and mass transfer rates (Senadeera et al. 2000). Additionally, fluidization

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enhances the mixing of the particles ensuring more homogeneous conditions (Nitz and Taranto 2007; Srinivasakannan and Balasubramanian 2008). An important parameter to control is the horizontal transport rate of particles in the continuous fluidised bed. In industrial fryers the horizontal transport is carried out with a conveyor belt that moves the French fries slowly through the oil (Vitrac et al. 2000). With increasing fluidization the horizontal transport rate would be reduced such that the residence time becomes uncontrollable.

The objective of this study is to quantify and control the effect of convective oil flow on the quality distribution of French fries, which is achieved by applying an upward oil flow through a packed bed of fries at increasing flow rates. First, the influence of flow conditions is studied in a model packed bed system to find the minimum fluidization velocities as function of the geometry. Subsequently, a newly developed pilot-scale frying unit with upward convective oil flow is used to study the influence of hydrodynamic conditions and particle geometry on fluidization behaviour and quality distribution in terms of moisture content. Finally, the influence of hydrodynamic conditions on packed bed mobility under different oil flow velocities is investigated.

Materials & methods

Fluidization of fry shaped particles

Initial fluidization experiments were carried out with bar-shaped (fry-shaped) objects fluidised in a tall column filled with water. The system consists of the column (10 cm internal diameter), a centrifugal pump (Iwaki MDH-401) to control the flow speed, a rotameter to measure the flow rate and a water reservoir. The minimum fluidization velocity was determined by changing the flow speed of the water while measuring variations in the. Since the dimensions of the column and particles are known, the void fraction can be directly determined from the measured bed height. Varying void fractions and particle sizes were used to validate calculations on the minimum fluidization velocity.

Bar shaped polycarbonate (PC) with a density of 1125 kg/m³ and polyvinylchloride (PVC) with a density of 1375 kg/m³ with different dimensions were used to establish the fluidization behaviour of fry shaped particles. PC and PVC were chosen for their ease of fabrication and their difference in density. The materials used and their ratios are summarized in Table 1. The different combinations were chosen to observe the fluidization behaviour for different geometries and different densities.

Ergun equation

The superficial velocity at which the pressure drop across a packed bed is just high enough to support the entire weight of the bed is called the minimum fluidization velocity. This velocity can be obtained by relating the pressure drop over a packed bed to the pressure drop across a completely fluidized bed. The latter can be obtained by balancing the net weight of the bed against the upward force exerted on the bed:

$$\frac{\Delta P}{L} = (\rho_s - \rho_f)(1 - \varepsilon)g \quad (1)$$

In which ΔP is the pressure drop (Pa) across the fluidized bed height L (m), ρ_s the density of the fluidized material (kg/m³), ρ_f the density of the fluid phase (kg/m³), ε the void fraction of the fluidized bed and g the gravitational constant (9.81 m/s²).

A frequently applied semi-empirical relation to characterise the pressure drop across a cylindrical packed bed of particles is the Ergun equation (Eq. 2) (Ergun 1949). This relation is mostly consistent with experimental data, where the pressure drop is a continuous function of the superficial velocity and is as follows:

$$\frac{\Delta P}{L} = 150 \frac{(1 - \varepsilon)^2}{\varepsilon^3} \frac{\eta u}{d^2} + 1.75 \frac{(1 - \varepsilon)}{\varepsilon^3} \frac{\rho_f u^2}{d} \quad (2)$$

Where ΔP is the pressure drop (Pa) and L is the length for the pressure drop measurement (m). The first term on the right hand side represents the viscous loss while the second term represents the inertial loss. In this expression 150 and 1.75 are the Ergun constants, η the dynamic viscosity of the fluid phase (Pa·s), u the superficial velocity of the fluid phase (m/s) and d the diameter of the fluidized particles (m).

Since the Ergun equation describes the pressure drop across a packed bed of spherical particles, it cannot be directly applied to a packed bed of non-spherical particles. A generally accepted approach to modify the Ergun equation for this situation is to replace the diameter d in eq. 2 by the equivalent diameter and multiplying it with the sphericity of the particles (Li and Ma 2011; Ozahi et al. 2008). The equivalent diameter of a particle is defined as the diameter of a sphere of equal volume to the particle while its sphericity is defined as the ratio of the surface area of the equivalent-volume sphere to the surface area of the actual particle (Hewitt 2002). As mentioned before, at the minimum fluidization velocity the pressure drop across the fixed bed will provide the onset of fluidization. This means that at the minimum fluidization velocity eq. 1 will

Table 1 Overview of the different particle geometries and combinations (in number of particles) used during the packed bed flow experiments. PC indicates use of polycarbonate particles, while PVC indicates polyvinylchloride particles

Run	16 × 4 × 4 mm PC	20 × 4 × 4 mm PC	24 × 4 × 4 mm PC	20 × 4 × 4 mm PVC
1	503	0	0	0
2	0	395	0	0
3	0	0	387	0
4	0	0	0	477
5	254	0	170	0
6	0	203	0	280

equal eq. 2 and an expression for the minimum fluidization velocity can be obtained:

$$u_m = \frac{150(1 - \varepsilon_m)}{3.5} \frac{\nu}{\phi d_p} \left(\sqrt{1 + \frac{7}{150^2} \frac{\varepsilon_m^3}{(1 - \varepsilon_m)^2} \frac{(\rho_s - \rho_f)}{\rho_f} \frac{\phi^3 d_p^3 g}{\nu^2}} - 1 \right) \quad (3)$$

With

$$\phi = \sum_{i=1}^n x_i \varphi_i \quad \text{and} \quad \varphi_i = \frac{\pi^{\frac{1}{3}} (6V_{pi})^{\frac{2}{3}}}{A_{pi}} \quad (4)$$

$$d_p = \sum_{i=1}^n x_i d_{pi} \quad \text{and} \quad d_{pi} = \frac{6V_{pi}}{\varphi_i A_{pi}} \quad (5)$$

$$\bar{\rho}_s = \sum_{i=1}^n x_i \rho_{si} \quad (6)$$

Where u_m is the minimum fluidization velocity (m/s), ε_m the void fraction of the packed bed, ν the kinematic viscosity (η/ρ_f , m²/s), φ the sphericity of the particles, d_p the equivalent particle diameter (m), ρ_s the particle density (kg/m³), x_i the volume fraction of particle fraction i , V_{pi} the volume of a particle from particle fraction i (m³) and A_{pi} the surface area of a particle from particle fraction i (m²).

The expression can be extended for materials of mixed size and/or composition by averaging the material properties by volume fraction, as shown in eqs. 4–6 (Asif 2011). Thus the minimum fluidization can be calculated as a function of the initial void fraction if the fluid properties and particle dimensions are known.

Frying experiments

The frying experiments were carried out in a newly designed frying unit that allows accurate control of the upward oil flow (Fig. 1). The frying unit has a close-coupled pump (KSB Etabloc SYT) to regulate the flow speed, a flanged immersion

heater (Cetal) to control the oil temperature and a Coriolis mass flow meter (Rheonik RHM-20) to measure the flow rate. Additionally, the frying compartment is equipped with two glass windows to observe the French fries during frying. A video camera is positioned in front of one window to monitor the initial fluidization behaviour of the French fries. Sunflower oil was chosen as the frying medium, as this is mostly used in industry.

Alexia potatoes purchased at a local supermarket were cut to a thickness of 0.8 by 0.8 cm and three different lengths; 3, 4 and 5 cm. For mobility frying experiments, half of the fries were coloured red by mixing these in an aqueous solution of Rhodamine B (0.01 % w/w) for 10 min while the other fries were treated for the same time with normal tap water. Afterwards tissue paper was used to remove the excess of surface water from the fries, after which they were placed in the frying basket to create two separate layers of coloured fries (see Fig. 2). A product to oil ratio of approximately 1 to 6 was used. The fries were fried for 1 min at 180 °C, as these are common pre-frying conditions in industry, with oil flow velocities varying between 0.007 and 0.07 m/s.

Quality distribution

In defining the quality of French fries, the most important parameters are moisture content, oil content, colour and texture. Since the water content and total oil content are to large extent determined by the method of frying, these two quality parameters were used as an indication of the quality distribution of the fries after frying. After frying, 10 fries were taken from the frying basket for further analysis. These fries were transferred to aluminium trays, weighed and dried to constant weight in an oven at 105 °C. The recorded loss in weight was used to represent the remaining water after frying. The dried fries were then submerged in petroleum ether (20 mL per fry) for 3 days to remove all the oil from the fries. Although this method of oil extraction differs from the more generally applied soxhlet extraction, it reduces the amount of solvent wasted. Moreover, it was examined in preliminary experiments that this method extracted the same amount of oil as conventional soxhlet extraction. The final weight after the oil extraction was used to represent the fat-free dry matter mass of the fry. After this the water content and total oil content after frying could be calculated by dividing the water remaining after frying and the oil removed from the fries by the fat-free dry matter mass.

Degree of mixing

Before and after frying a picture of the top view was taken from the frying basket. The obtained pictures were cropped to obtain only the packed bed with fries and divided into grid cells (Fig. 2).

Fig. 1 **a** Cross-flow frying unit with pump (1), heat exchanger (2) and frying compartment (3). **b** Schematic drawing of the frying compartment with two glass windows; the arrows indicate the direction of the circulating oil flow

An in-house developed routine in Matlab (version 7.10.0 with image processing toolbox) was used to determine the amount of red and yellow pixels in each grid cell. To quantify the degree of mixing the mixing entropy was used (Schutyser et al. 2001). The mixing entropy for a single grid cell was calculated using:

$$S(i, j) = p_r(i, j) \log(p_r(i, j)) + p_y(i, j) \log(p_y(i, j)) \quad (7)$$

Where $S(i, j)$ is the mixing entropy in cell (i, j) , $p_r(i, j)$ the fraction of red pixels in cell (i, j) and $p_y(i, j)$ the fraction of yellow pixels in cell (i, j) . The total mixing entropy was then calculated using:

$$S_{tot} = \frac{1}{-\log 2} \sum_{i,j} S(i, j) \left(\frac{N_{cell}(i, j)}{N} \right) \quad (8)$$

With S_{tot} being the total mixing entropy, $N_{cell}(i, j)$ the amount of red and yellow pixels in cell (i, j) and N the amount of red and yellow pixels in the entire image. The ratio between the coloured pixels per cell and total amount of coloured

Fig. 2 Division of image used for image analysis into 16 grid cells to determine mixing entropy

pixels is used to correct for cells with a relatively low amount of coloured pixels. The $-\log(2)$ factor is the mixing entropy at maximum mixing and is used to scale the calculated mixing entropies from 0 to 1. When the grid cell size is too small then the total mixing entropy will remain 0, as only one fry will fit each cell, whereas when the grid cell size is too large then the entropy will be already close to 1 at the beginning of the mixing. Using a square grid it was found that 16 grid cells gave the best result, with less cells causing values to converge towards 1, while more cells caused values to move towards zero.

Results & discussion

Predicting the minimum fluidization velocity

The flow rate was varied while the bed height was monitored to obtain the minimum fluidization velocities for different materials. The obtained void fractions as a function of the superficial velocity are presented in Fig. 3. It can be seen that the void fraction remains constant until a certain point after which the bed starts to expand. This point represents the transition from a packed to a fluidized bed. The minimum fluidization velocity is the superficial velocity at which the void fraction starts to increase. Run 4 clearly shows the highest minimum fluidization velocity, followed by run 6. This can be explained by the amount of PVC particles in these two runs, since the PVC has a higher density than the PC, it will require a higher fluid velocity to fluidize. Other differences are less obvious, as the other runs only represent differences in geometry and thus the fluidization behaviour depends primarily on the specific packing. Run 5 shows the lowest fluidization velocities, which may be attributed to the binary composition of the system. Mixing of two differently sized particles can result in densely packed beds (Asif 2011), which in turn decreases the fluidization velocity due to a higher pressure drop.

Minimum fluidization velocities (u_m) were also calculated using eq. 3 and depicted in Fig. 3. The Ergun equation could predict the experimentally determined minimum fluidization velocity of fry shaped particles with a difference of at maximum $\pm 10\%$. This is in agreement with other studies

Fig. 3 The void fraction of the fixed/fluidized bed as a function of the superficial velocity for different rectangular particle shapes, as listed in Table 1: run 1 (■); run 2 (×); run 3 (○); run 4 (+); run 5 (◇) and run 6 (△). The minimum fluidization velocity calculated using eq. 3 for the different runs is indicated with a corresponding filled data point

describing the minimum fluidization velocity of non-spherical particles using the adjusted Ergun equation (eq. 3) (Li and Ma 2011; Ozahi et al. 2008). Though better predictions might be obtained by fitting modified Ergun-type equations, this is always done for a specific set of materials, shapes and possible wall effects, which makes them less appropriate for extrapolation to other data sets.

Initial fluidization during frying

Batches of 3, 4 and 5 cm long fries were subjected to vertical oil flow velocities between 0.007 and 0.07 m/s at 180 °C to experimentally determine the minimum fluidization velocity during frying. Video recordings of the initial seconds, before water evaporated, showed either a fixed bed of fries or a fluidized bed of fries. By considering the largest flow rate that still showed a fixed bed and the smallest flow rate that showed a fluidized bed, a region was found that should contain the minimum flow rate required for fluidization (Fig. 4). Additionally, the minimum fluidization velocity was predicted using eq. 3 and plotted in Fig. 4. Input parameters are derived from the bed and fry dimensions, while the density was 1030 kg/m³ determined by means of weighing and the water displacement method.

The experimentally determined regions in which the minimum fluidization velocity is expected overlaps with the predicted minimum flow rate required for fluidization, taking into account a 10 % relative error. This is in accordance with the findings in section 3.1, where the validity of a modified Ergun equation was confirmed for rectangular particles.

Quality distribution

The final moisture and oil contents on fat-free dry matter base are plotted in Fig. 5 against the different oil flow velocities for

different fry lengths. The final moisture content was observed to be equal for different flow rates. This is expected since water evaporation is a function of the heat transfer rate which in turn depends on the temperature differences, fry size and frying time. Since the oil temperature, fry size and frying time are all constant within each data set, similar amounts of water loss are expected. A parameter that could cause variation in water loss is the increasing oil flow rate, which should positively affect the external convective heat transfer coefficient. However, most of the heat transfer will take place in the boiling phase during which water is evaporating and at that moment the heat transfer coefficient is dominated by the local turbulence close to the surface of the fry, induced by vapour bubble formation (Costa et al. 1999; Farkas and Hubbard 2000). Similar values for the moisture content are found in literature studies in the absence of forced convective oil flow (Krokida et al. 2000, 2001; Van Koerten et al. 2015).

Fig. 4 Minimum fluidization velocity calculated using eq. 3 plotted against the fry length (■). The dotted lines indicate a 10 % relative error. The connected blocks (■) represent the maximum flow velocity at which no fluidization was observed and the minimum flow velocity at which fluidization was observed for the fry lengths of 3, 4 and 5 cm

Fig. 5 Left side: Water content per fat-free dry matter with standard deviation plotted against the superficial oil velocity for 3 (a), 4 (c) and 5 (e) cm fries. Right side: Oil content per fat-free dry matter with standard deviation plotted against the oil flow velocity for 3 (b), 4 (d) and 5 (f) cm fries. All fries were fried at 180 °C for 1 min

The final oil content was also expected not to vary much for different flow rates. This can be observed from the results shown in Fig. 5. There is more overall variation in the oil content than in the water content. A possible explanation for this is related to the rate at which the frying basket is removed from the hot oil. At a higher removal rate, less oil remains adsorbed onto the surface of the fry, yielding a lower total oil content (Krozel et al. 2000). Still, the oil content values obtained are within the range of total oil contents as observed in other studies (Moyano and Pedreschi 2006; Ziaifar et al. 2010).

To assess the variability in the quality of the fried French fries, we use the measured variations within each data set, as indicated by the standard deviation. In Fig. 6, the standard deviations of the moisture content data are plotted versus the oil flow velocity. Additionally, the minimum fluidization velocity obtained from the initial fluidization experiments is indicated in Fig. 6. There is a decreasing trend in the standard deviation for flow velocities higher than 0.03 m/s for the 5 cm

fries. Note that this is before initial fluidization is achieved. The standard deviation also decreases for the 3 cm fries before

Fig. 6 Standard deviations of the water content at different oil flow velocities for 3 (▲), 4 (■) and 5 (◆) cm fries. The vertical lines indicate the experimentally obtained minimum fluidization velocity for the different fry lengths, where increasing superficial velocity corresponds to increasing fry length

the initial fluidization, but actually increases after this fluidization point. The latter is also observed for the 4 cm fries. However, the standard deviation of the 4 cm fries shows no significant change before fluidization is achieved.

For 5 cm fries it can be clearly observed that increasing oil flow velocity has a positive influence on the quality distribution as the standard deviation is continuously decreasing for flow velocities higher than 0.03 m/s (Fig. 6). This can be attributed to the increased oil refresh rate, which in turn will lead to lower temperature gradients inside the frying compartment. The 3 cm fries also show a decrease in distribution before the minimum fluidization velocity is reached, but this is already visible after flow velocities higher than 0.01 m/s. This suggests that the void fraction of the bed of fries can already partially increase before actual fluidization is reached, which increases the oil flow through the bed, and that this happens earlier for beds with a lower minimum fluidization velocity. However, this effect is not visible for the 4 cm fries, which would then be expected to decrease for flow velocities between 0.01 and 0.03 m/s. Additionally, a clear trend of decreasing homogeneity is observed for the 3 and 4 cm fries when they start to fluidize. It may be expected that upon fluidization more mixing is achieved and the fries would be fried more equally. One explanation for this increase is the limited height of the fluidization compartment (approximately 1.5 times the packed bed height), with air at the surface. If fluidization takes place and the bed expands, some fries are pushed to the surface and experience a smaller contact area with the oil, thus achieving different drying rates. Another reason for

the increased quality distribution might be bed instabilities. Such bed instabilities are related to non-homogenous packing and enhanced by resulting unequal upward oil flow distribution. Bed instabilities are predominant in systems with a high diameter ratio between the fluidization compartment and the fluidized particles as is the case in this work (Duru et al. 2002). Non-homogenous packing with different void fractions will result in varying exposure of fries to oil, which consequently affects the drying rates.

Mobility of French fries during frying

In Fig. 7 the mixing after frying is visualised using coloured French fries as function of the fry length and the oil flow velocity. The smaller fries mix to a greater extent and at lower flow velocities than the larger fries. The red line in Fig. 7 indicates the experimentally obtained minimum fluidization velocities for the different fry lengths.

The degree of mixing was also quantified in terms of mixing entropy, which was determined by image analyses after frying fries of 3, 4 and 5 cm at different oil flow rates. These results are plotted in Fig. 8. The entropy starts with a value of 0.074, representing the non-mixed situation, and then asymptotically moves towards a maximum degree of mixing (normalised to 1). In line with the visual observations, it can be concluded that the mixing of entropy increases less fast with increasing flow rate for the 5 cm fries.

These results are in line with section 3.2, where the fluidization velocity was shown to be lower for smaller

Fig. 7 Pictures visualizing the amount of mixing after 1 min of frying at different oil flow velocities and fry lengths. The red line indicated the location of the experimentally obtained minimum fluidization velocity for the different fry lengths

Fig. 8 Mixing entropy as calculated by eq. 8 plotted against the oil flow velocity for fry lengths of 3 (▲), 4 (■) and 5 (◆) cm. Lines are drawn to guide the eye

fries, thus more mixing is expected. However, Fig. 7 also shows that for 3 and 4 cm fries, mixing already takes place before the minimum fluidization velocity is reached. This can be partly attributed to the water evaporation during frying which will cause the fries to decrease in density. As the fries decrease in water content from 4 to 3 g/g dry matter during the 1 min of frying, the density is expected to decrease by 50 kg/m^3 (Costa et al. 2001). According to eq. 3, a density decrease of 50 kg/m^3 would result in a decrease in minimum fluidization velocity of approximately 10 % for the different fry lengths. However, this does not explain the mixing at velocities lower than 90 % of the minimum fluidization velocity. In a previous study it was observed that before fluidization of the entire bed already some particles were fluidised at the top of the bed, while the bottom of the bed remained stagnant (Duru et al. 2002). Since emerging vapour bubbles increased turbidity in the oil above the bed, only the lower part of the bed could be clearly monitored for initial fluidization experiments. This means that there might already be minor bed movement before any observed fluidization takes place. This confirms the suspected increase in void fraction at velocities below the minimum fluidization velocity discussed in section 3.3.

The faster observed mixing of the shorter fries, already at low flow rates, can also explain why the standard deviations in the moisture content for these fries do not decrease consistently with increasing flow rates. The degree of mixing for the 4 cm fries is already close to that of 3 cm fries at a flow rates around 0.04 m/s. This suggests that application of even a small oil flow to the 3 and 4 cm fries provides a more homogeneous quality distribution due to a higher oil refresh rate. At higher flow velocities, when the fries start to fluidize, the quality variation becomes subject to the bed instabilities at the corresponding flow velocities, and becomes independent of the oil flow velocity.

Conclusions

A modified Ergun equation was found to accurately describe fluidization of rectangular shaped particles in a water column and used to characterize initial fluidization behaviour of French fries during frying in a cross-flow frying set-up.

Subsequently, the effect of varying flow conditions on the quality distribution as represented by the moisture content and the oil content was studied. For fries of 3 and 5 cm long, an increased oil flow velocity improved the homogeneity of the fries, while this was not observed for fries of 4 cm.

The smaller fried were found to mix quickly during frying, while longer fries exhibited much less mixing. This is because of two effects: the higher minimum fluidization velocity for longer fries, due to their stronger entanglement in the bed, and the smaller surface area of long fries leading to a slower dehydration and thus later fluidization during the frying process.

The crossflow velocity was found to be an important parameter for the homogeneity of the quality of French fries. Increased oil flow velocities positively affect the quality distribution due to a higher oil refresh rate. However, inducing fluidization caused the quality distribution to become unpredictable due to the bed instabilities caused by the system dimensions.

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