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Experimental investigation of monodisperse solids drying in a gas-fluidized bed

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HIGHLIGHTS

- Fluidized bed particle drying is experimentally studied.
- Particle image velocimetry and infrared thermography techniques are applied.
- Full field solids fluxes and temperatures are measured.
- Gas-particle heat and mass transfer is characterized.
- Drying influences the solids hydrodynamics by changes in particle density.

GRAPHICAL ABSTRACT

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ABSTRACT

Fluidized bed drying is an unsteady process where particle and gas properties evolve in time. The bed hydrodynamics and its interplay with mass and heat transfer changes in time. In this study, experiments in a pseudo-2D fluidized bed setup are performed to obtain insight in this complex behavior. A coupled particle image velocimetry infrared thermography technique provides solids velocity and temperature fields in the bed. Significant hydrodynamic changes due to the evaporation of water from the porous γ-Al₂O₃ material for three superficial velocities were observed. The 1.25 umf case resulted in a fixed bed state, but after drying for 580 s a bed inversion took place. The 1.50 and 1.75 umf experiments showed an increase in bed expansion over time. Furthermore, clear asymmetrical solids volume fluxes were observed for the wet material that became more symmetrical over time when the material dried.

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Introduction

Gas-solid fluidized beds are frequently applied in industrial applications such as drying, gas phase polymerization and granulation due to their excellent solids mixing as well as their excellent heat and mass transfer characteristics. The behavior of fluidized beds is difficult to predict due to the complex prevailing gas–solid interaction and resulting flow patterns. During drying processes, this complex gas–solid interaction becomes even more pronounced. Heat and mass transport display strong interactions with the bed hydrodynamics. Evaporation causes particle mass reduction over time that impacts the hydrodynamic behavior. The
hydrodynamics in turn influences the mass and heat transfer rates and thus the rate of evaporation.

The rate of evaporation mainly determines the process efficiency, which is crucial in industrial applications since solids drying is a very energy-intensive process due to the high latent heat of water vaporization. Therefore many research efforts have been made in order to increase the understanding (Daud, 2008; Defraeye, 2014). For example, the food or pharmaceutical industry, subtle changes in the drying process can lead to large variations in the product quality (Sivakumar et al., 2016; Zhang et al., 2021). Besides, the moisture inside the granulate has a strong interaction with the material itself, leading to material shrinkage or internal moisture profiles (Sivakumar et al., 2016; Vu and Tsotsas, 2018). Both phenomena contribute to the fact that additional drying time is needed in order to reach a desired moisture content threshold, causing a lower drying efficiency and the risk for non-uniform drying rates among the particles. Findings from previous fluidized bed drying research are summarized in Table 1. It is observed that most research studies investigated the effect on the drying rate by varying general process parameters such as the drying temperature and gas flow rate.

Although these parameters certainly influence the drying rate, insufficient attention has been paid to the local hydrodynamics or local bed temperature. The local gas–solid interaction can be quantified by applying detailed optical techniques such as particle image velocimetry (PIV) while the local bed temperature can be characterized by applying advanced experimental techniques such as infra-red thermography (IRT). Salmasi et al. (2018) showed that the local solids motion had a severe effect on the heat and mass transfer rates on biomass devolatilization. This emphasizes that gas–solid interactions also clearly influence the overall process which is critical in terms of process efficiency and thermosensitive product degradation. It is hypothesized that the local bed temperature has a similar effect on the overall process. However, in the literature, this detailed information is lacking since the bed temperature is commonly measured via thermocouples. Full understanding of a drying process can only be obtained when the local gas–solid motion and the corresponding local solids temperature can be quantified.

Therefore, in this work, the detailed gas–solid interactions for fluidized bed drying are studied by making use of a combined particle image velocimetry (PIV) and infra-red thermography (IRT) technique. This combination leads to the desired detailed local hydrodynamic and thermal information by converting both data sets into valuable data using a digital image analysis (DIA) technique. The combined PIV/DIA technique was initially developed by van Buijtenen et al. (2011) and De Jong et al. (2012). The infra-red technique integration towards PIV/DIA/IR was reported by Patil et al. (2015b) building on the work of Tsuji et al. (2010) who studied heat transfer in fluidized beds. The combined technique was further extended and used for reactive systems (Li et al., 2017) and for liquid injection (Sutkar et al., 2015; Kolkman et al., 2017). Recently, Milacic et al. (2022) studied steady-state temperature distributions inside a fluidized bed.

This work is organized as follows. In the next section, the experimental method for determining the solids velocity and temperature fields by means of the PIV/DIA/IR method is described. The main results for a pseudo-2D fluidized bed with drying particles are presented in Section 3 and a final conclusion is given in Section 4.

**Experimental method**

**Setup**

The experiments were carried out in a pseudo-2D fluidized bed setup with width 8 cm, height 20 cm and depth 1.5 cm. The bed walls were made of polymethylmethacrylate (PMMA) with a thickness of 20 mm to limit the heat loss to the environment. The front wall was made of sapphire glass with a thickness of 3 mm. Sapphire glass has excellent properties for IR measurements. On the back wall, a black aluminum plate was attached to the PMMA wall to improve the contrast between the different fluidization phases (e.g. bubble, emulsion) which enhances the performance of the cameras. Nitrogen gas was uniformly fed through a porous bottom plate by making use of mass flow rate controllers. The gas entered the bed at an elevated temperature and the setup was placed in a constant environmental temperature of 18 °C. Both an optical and infra-red camera were used for capturing the fluidization behavior in the setup. A schematic overview can be seen in Fig. 1, where a high-speed IR camera (FLIR 8400sc, resolution 1024 x 648) and a high-speed visual camera from PCO (Dimax HD, resolution 1080 x 720) were used. The optical camera was placed in front of the fluidized bed and the IR camera was positioned at a slight angle. Both cameras were triggered via a Velleman PCGU 1000 controller in order to synchronize the frame rate. Besides, several sensors were incorporated in the setup of which the most important ones were the differential pressure sensor to measure the pressure drop over the fluidized bed and the thermocouples that were used to measure the gas temperatures during the experiments.

**Procedure**

The fluidized bed setup was filled with porous spherical \( \gamma – \text{Al}_2 \text{O}_3 \) particles provided by Sasol with a size of 1.81 mm. Due to the high particle porosity, the particle density was relatively low, namely, 970 kg/m\(^3\). The bed was loaded with a volume of

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**Nomenclature**

\[ \begin{align*} A & \quad \text{Area (m}^2\text{)} \\ C_p & \quad \text{specific heat capacity (J/kg/C)} \\ h_w & \quad \text{Overall heat transfer coefficient (W/m}^2/\text{C)} \\ d_{bed} & \quad \text{Depth fluidized bed (m)} \\ h_{bed} & \quad \text{Height fluidized bed (m)} \\ n & \quad \text{Number of 16 x 16 pixel windows (-)} \\ T & \quad \text{Temperature (°C)} \\ u & \quad \text{Gas velocity (m/s)} \\ u_{min} & \quad \text{Minimum fluidization velocity (m/s)} \\ w_{bed} & \quad \text{Width fluidized bed (m)} \\ c_p & \quad \text{Solids fraction field (-)} \\ \phi & \quad \text{Solids volume flux (m}^3/(\text{m}^2\text{s})) \\ \rho & \quad \text{Density (kg/m}^3\text{)} \\ \sigma & \quad \text{Standard deviation particle temperature (°C)} \\ \text{CSTR} & \quad \text{Continuous stirred-tank reactor} \\ \text{DIA} & \quad \text{Digital image analysis} \\ \text{IRT} & \quad \text{Infra-red thermography} \\ \text{IR} & \quad \text{Infra-red} \\ \text{PDF} & \quad \text{Probability density function} \\ \text{PIV} & \quad \text{Particle image velocimetry} \end{align*} \]
Slight deviation was calculated as:

\[ \Delta T_X = \frac{1}{n} \sum \left( T_{X,n+1} - T_{X,n} \right) \]

where \( n \) is the total number of 16 x 16 pixel windows. Following a similar approach, the instantaneous mean temperature was calculated as follows:

\[ T_p = \frac{\sum n e_p(n)T_p(n)}{\sum n e_p(n)} \]

where \( n \) is the total number of 16 x 16 pixel windows. Following a similar approach, the instantaneous particle temperature standard deviation was calculated as:

\[ \sigma = \left( \sum n e_p(n)\Delta T_{p,n}^2 \right)^{1/2} \]

\[ \Delta T_{p,n} = T_{p,n+1} - T_{p,n} \]

Data acquisition

In order to obtain a velocity field from PIV measurements, two consecutive images per measurement are required. The measurements were conducted with an acquisition frequency of 10 Hz where a delay of 3.4 ms was used for the two consecutive images taken by the optical camera. The acquired images need to be cropped before the data analysis could be performed. The PIV processing was performed according to the method described by van Buijtenen et al. (2011) and De Jong et al. (2012) by making use of DaVis (LaVision) 8.2.3 software where an interrogation region of 64x64 pixels was used. A multi-pass algorithm with 32x32 pixels and 50% overlap with decreasing size was adopted. The 3D solids fraction field was obtained by the method proposed by De Jong et al. (2012) where the 2D solids fraction fields from the visual camera was converted into a 3D solids fraction field. The 3D solids fraction fields were combined with the velocity vectors, resulting in the time-averaged solids volume fluxes calculated via:

\[ \phi = \nu_p \epsilon_p \]

The infra-red images were divided in the same window sizes as for the solids fraction fields. The solids fraction field was combined with the thermographic information obtained from the infra-red images in order to retrieve a particle temperature field. For more information, the reader is referred to Milacic et al. (2022). In order to correct for the solids volume fraction, the instantaneous mean temperature was calculated as follows:

\[ T_p = \frac{\sum n e_p(n)T_p(n)}{\sum n e_p(n)} \]

Before wetting the dry particles, the particle mass was measured and the particles were placed in a water bath. After this water bath, the particles were placed in an oven for a short time to remove the liquid layer around the outer particle surface. This is in order to ensure that no liquid bridges can be formed since these bridges lead to changing particle collision properties (Antonyuk et al., 2009). Subsequently, the particle weight was again measured and the water mass inside the porous particles could be determined. The material was put inside the fluidized bed setup using a gas inlet temperature of ±77 °C.

### Table 1

<table>
<thead>
<tr>
<th>Authors</th>
<th>Application/material</th>
<th>Major findings</th>
</tr>
</thead>
<tbody>
<tr>
<td>Jabboon et al. (2009)</td>
<td>Waxy rice</td>
<td>Higher drying temperatures led to higher head rice yield. However, too high temperatures had a negative effect on the granule morphology.</td>
</tr>
<tr>
<td>Srinivasakannan and Balasubramanian (2009)</td>
<td>Millet</td>
<td>The drying rate increased with increasing temperature and flow rate while it decreased with an increase in solids holdup.</td>
</tr>
<tr>
<td>Momennazadeh et al. (2011)</td>
<td>Shelled corn using microwave-assisted drying</td>
<td>The drying rate increased with increasing temperature and microwave energy level.</td>
</tr>
<tr>
<td>Silva et al. (2012)</td>
<td>Soybean meal</td>
<td>Temperature had a significant influence on internally limited soybean drying while the bed height and gas velocity did not have a significant effect on these experimental conditions.</td>
</tr>
<tr>
<td>Ozahi and Demir (2015)</td>
<td>Corn and unshelled pistachio nut</td>
<td>An increase in particle mass resulted in a higher drying time, while an increase in air temperature and air velocity led to a decrease.</td>
</tr>
<tr>
<td>Idakiev et al. (2017)</td>
<td>Inductive heating of alpha and gamma aluminium oxide</td>
<td>Inductive heating could shorten the start-up process of batch fluidized bed drying. Besides better controllable temperature profiles were obtained.</td>
</tr>
<tr>
<td>Yogendrasasidhar and Pydi Setty (2018)</td>
<td>Kodo millet grains and Fenugreek seeds in a wall heated fluidized bed dryer</td>
<td>The drying rate increased with increasing wall temperature and gas flow rate while it decreased with an increase in bed height.</td>
</tr>
<tr>
<td>Szadzisinska et al. (2019)</td>
<td>Microwave- and ultrasound-assisted convective drying of raspberries</td>
<td>Microwave and ultrasound assistance led to higher drying rates. Besides, better product quality was maintained.</td>
</tr>
<tr>
<td>Lehmann et al. (2020)</td>
<td>Fluidized bed drying using mechanical vibration</td>
<td>Vibration reduced the minimum fluidization velocity, which enables operations at lower gas velocities.</td>
</tr>
<tr>
<td>Luthra and Sadaka (2021)</td>
<td>Rough rice drying using dehumidified air in fixed and fluidized beds</td>
<td>Dehumidified air had a positive effect on moisture removal. No head rice yield difference was found between fixed and fluidized beds.</td>
</tr>
</tbody>
</table>

In order to correct for the solids volume fraction, the instantaneous mean temperature was calculated as follows:

\[ T_p = \frac{\sum n e_p(n)T_p(n)}{\sum n e_p(n)} \]

Where \( n \) is the total number of 16 x 16 pixel windows. Following a similar approach, the instantaneous particle temperature standard deviation was calculated as:
Fig. 2 shows the overall heat transfer coefficients as a function of superficial velocities and temperature driving forces (bed - surroundings). The overall heat transfer coefficient is composed of three terms: bed-to-wall heat transfer, heat conduction through the sapphire glass and wall-to-environment heat transfer. The bed-to-wall heat transfer in fluidized beds can be estimated from the proposed equation of Glicksman, Decker and Baskakov (Kennedy and Levenspiel, 1991). Depending on the local fluidization conditions, a value in the range of 150–250 W/m²/K for the used particle size is estimated. The sapphire glass conduction heat transfer coefficient is calculated via the data of Milacic et al. (2022), resulting in a very high value equal to 9070 W/m²/K. Therefore, it can be safely concluded that the heat loss is mainly controlled by the wall-to-environment heat transfer rate. However, the radiation and free-convection heat transfer coefficients for a vertical flat plate (Bird et al., 2007; Churchill and Chu, 1975) are lower than the overall heat transfer coefficient found in our system. The neglected heat loss through the PMMA walls is not the reason for this underestimation as PMMA has a thermal conductivity value equal to 0.07 W/m/K and the walls have a thickness of 2 cm, leading to a relatively low heat conduction rate. Taking into account that the setup is placed in a laboratory with a relatively high airflow due to the ventilation, it is concluded that also forced-convection is an important factor. The Richardson number substantiates this as a mixed convection region value is obtained. The small increase observed for the higher superficial velocities can be explained by the increased bed expansion that leads to a larger particle–wall contact area. The bed is heated from the bottom where the particles will experience the highest temperature. Over the axial position, the effect of the high bottom inlet becomes less causing a non-uniform sapphire glass temperature. As more bed expansion occurs, the particles will come into contact with a relatively cold sapphire glass leading to a larger cooling rate. 

Particle drying

The particle drying experiments were carried out using the described experimental procedure. Next to the steady-state heat loss characteristics described in the previous section, the initial column temperature is of importance which should be taken into account while analyzing the obtained data. The initial column temperatures and the used amount of $\gamma$-Al$_2$O$_3$ are summarized in Table 3. The drying process is further quantified by first showing the instantaneous image profiles. The used time instances are discussed in detail in the subsequent subsections, where first solids volume flux profiles are shown, and second the particle temperature profiles are examined by means of mean particle temperature and standard deviation of the particle temperature in time. The particle temperature distributions are further investigated using instantaneous normalized probability distribution functions (PDF) and an accumulative distribution (1.25$u_{in}$ only).

Instantaneous image profiles

Figs. 4–6 present the instantaneous visual and infra-red profiles for respectively the 1.25, 1.50 and 1.75$u_{in}$ drying cases. A clear difference between the lowest and the other two velocities is directly observed since the lowest superficial velocity results in a fixed bed state at the start of the experiment. The increased particle mass (displayed in Table 3) due to the addition of water inside the porous material causes a higher required minimum fluidization velocity. Due to the hot gas injection from the bottom, the bottom section particles dry faster (the top particles experience gas containing a high relative humidity) resulting in a propagating heat front. At a certain point in time, the liquid is partially evaporated.
lowering the required minimum fluidization velocity. Fig. 4C shows an inversion of the hot bottom particles to the top of the bed whereafter partial fluidization occurs due to the large particle densities in the bed. The particle density differences are not easy to measure inside this experimental setup, but previous Computational Fluid dynamics - Discrete Element Method studies (Di Renzo et al., 2008; Karimi and Dehkordi, 2015; Peng et al., 2016) have shown the effect of particle density differences on segregation and mixing inside a fluidized bed.

The two other superficial velocity cases are in a fluidizing state from the start of the experiments, where Fig. 5 presents the behavior over time for the middle superficial velocity. It can be clearly seen from the visual images that the bed expands over time indicating that the fluidizing behavior is changed. We will further quantify this observation in the subsequent sections. The temperature snapshots in Fig. 5B and C indicate a wet-bulb regime due to the externally limited drying. We will further analyze these drying regimes in the particle temperature discussion. The particle bed is in a quite uniform temperature state which is not the case in the fluidized bed, time-averaged solid fluxes are presented in Fig. 7–9 for the 1.25, 1.50 and 1.75umf cases respectively. We only analyze the solids volume fluxes for the 1.25umf case after the inversion started at 580 s (see Fig. 4), where the bottom part of the bed is fixed while the top is in a fluidized state. The time-averaged solid fluxes were obtained by averaging the flux data for 20 s (i.e. 200 frames) in order to obtain a representative data set. It should be noted that this averaging is permitted due to the hydrodynamic and drying time scales (i.e. the particle mass is not drastically changed in this time frame).

Similar to the observation in the snapshots, the 1.25umf case leads to a partially fixed and partially fluidized state. The wet material in the bottom is still below the minimum fluidization velocity, while the already dried material is above this threshold. In time, the fluidization degree is increased, leading to slightly larger solid volume fluxes. The solids flux profiles for the two other cases also clearly show the transition in the fluidization regime by means of the increase in velocity profile and the bed expansion. It is further noted that the wet material caused a clear rotational solids motion inside the bed which is not observed for the dry material. This asymmetrical behavior could be enhanced due to the limited bed depth compared to the particle diameter. The flow pattern can also be explained as a gulf stream phenomenon. This gulf stream is caused by non-uniform bubble flow. One part in the bed has an upward solids flow, while the other part has a downward flow. Merry and Davidson (1973) suggested that this phenomenon could be introduced by uneven gas distribution while Verma et al. (2015) also observed this behavior inside larger fluidized beds with uniform gas distribution. The continuous mass reduction due to drying leads to non-uniform solid densities, which in principle can be explained as a local difference in gas velocity in excess of the required minimum fluidization velocity. Therefore, the bed density distribution can enhance this asymmetrical rotational solids motion. In a later stage when the drying proceeds (see the 1.75umf data for a more clear distinction), again a regular pattern is observed.

The 1.75umf case shows this dry material behavior in Fig. 9D where an upward solids motion inside the middle of the bed is observed whereas the particles are moving downwards near the walls. As observed earlier, it is also noted that the higher velocity results in faster drying of the particle material.

In order to further quantify the hydrodynamic changes caused by the particle drying, the vertical solids volume flux is shown at a height of 5.47 cm for the 1.50 and 1.75umf drying cases. It can be clearly seen in the early drying stages that the rotational solids motion results in a non-symmetrical vertical solids volume flux
Furthermore, the magnitude of the solids volume fluxes increases over time indicating the earlier mentioned fluidization regime transition. Fig. 10B shows a more symmetrical shape at 750–770 s which was also observed earlier. This indicates that the material is relatively dry.

**Particle temperature**

The particle temperature will also have an influence on particle drying. The ability of the gas phase to store a certain vapor content is correlated with the gas temperature. Drying can be captured in different regimes which will affect the solids temperature. These regimes and corresponding evaporation rates will also apply to γ-Al₂O₃ as reported by Koptyug et al. (2000). Dependent on the gas and particle temperatures at the start, initial heating or cooling is observed until the wet-bulb temperature is reached. In an adiabatic system, the wet-bulb temperature can be computed using the dry-bulb temperature and the relative humidity. However in our case, the heat losses to the environment and the continuously changing reactor wall temperatures will influence this value. The wet-bulb temperature indicates also the start of the second regime where the drying rate is externally limited, called the constant drying regime. This regime will hold until a certain point in time where the mass transfer rate from the solid to the gas becomes larger than the internal rate of liquid supply to the surface where the evaporation takes place. At that moment, the drying becomes internally limited and the energy required for the evaporation becomes less due to the reduced evaporation rate. Therefore, still some energy is left to heat up the material.

![Instantaneous visual (top) and infra-red (bottom) profiles for the 1.25um case.](image)

The liquid content caused a particle mass increase leading to a superficial velocity lower than the required minimum fluidization velocity. Hence, the fixed bed heats up from the bottom where the hot gas is injected, causing a particle inversion starting at 580 s. Thereafter, the bottom part of the bed is fixed while the top is in a fluidized state.
Fig. 11 shows the mean and standard deviation for the particle phase calculated according to Eq. 2 and 3. All experiments show a temperature increase at the start. As shown in Table 3, the initial column temperature was slightly higher for the 1.25μm case, which explains the higher obtained maximum value. This column temperature is of high importance for the obtained particle temperatures, since the bed wall is acting as a heat supply or removal during the whole experiment. Please note that the PMMA and sapphire glass walls have different thermal properties. In order to avoid additional effects due to the pretreatment procedure by means of putting the particles inside the oven, an additional investigation was conducted (not shown here). It was concluded from these experiments that the initial moisture content (as listed in Table 3) does not play an influential role in the temperature increase which is observed in the experiments between the start and 100 s. Therefore, we conclude that this effect is fully caused by the bed wall temperature.

We will now make a clear distinction between the analysis of the fluidizing and fixed bed cases. The fluidizing cases of 1.50 and 1.75μm feature similar behavior. Both experiments showed a temperature drop starting around 40 s, after the initial heating. This can be explained by the continuously changing wall temperatures causing a wet-bulb temperature change. The minimum temperature is reached around 500–600 s where after the solid material slowly starts to heat up again indicating the start of an internally limited drying regime. As discussed in
the previous section, the solids flux profiles become more uniform profile above 600 s which could substantiate this internally limited drying regime since the particle material density is somewhat more uniform. One would expect that the temperature increase at 1.75umf would be larger than at 1.50umf. However, at this point in time the bed walls are acting as a heat sink limiting the temperature increase. The 1.75umf case shows a higher standard deviation in the particle temperature compared to the 1.50umf case. This is expected due to the larger inaccuracy of the infra-red image capturing process for the bubble phase, see Milacic et al. (2022). The case with the highest fluidization velocity also displays more bed expansion due to the formation of larger bubbles. Besides, Fig. 6 shows a larger area with a higher solids temperature compared to Fig. 5 in the bottom zone. This is due to the larger gas volumetric flow rate. Both factors are contributing to a larger particle temperature standard deviation.

The 1.25umf case clearly showed a different temperature behavior due to the initial fixed-bed state. Fluidized beds are known for their good gas-particle heat transfer characteristics compared to fixed beds. After the initial temperature increase, the mean particle temperature slowly increases. This is due to the fact that two regions can be distinguished. In the bottom, particles are already in the internally limited drying regime, while the particle drying is totally externally limited due to the saturated moisture content in the gas in the top region. The bottom region expands over time, causing a mean particle temperature increase until 580 s where the

Fig. 6. Instantaneous visual (top) and infra-red (bottom) profiles for the 1.75umf case. Over time, the bed will expand indicating the transition in fluidization regime. The bed is in a wet-bulb regime for the 240 and 480 s snapshots whereafter the granular material start to heat up again indicating the start of an internally limited drying regime.
particle inversion took place (see Fig. 4). Upon fluidization, the hot particles provide heat for the evaporation of water present inside the cool particles. The overall temperature will therefore drop fast. Fig. 11B clearly shows the different fixed-fluidized bed heat transfer characteristics since the maximum particle temperature standard deviation is one order of magnitude larger compared to the fluidizing cases. After the bed inversion, the particle temperatures homogenize due to the more extensive mixing which reduces the standard deviation. Ultimately the particle temperature standard deviation almost reaches similar values at the end (800 s) as the two higher superficial velocities.

The normalized particle temperature probability density functions (PDF) presented in Fig. 12 lead to a better understanding of the previously presented particle temperature standard deviations. Here 200 bins were used for determining the PDF while using the minimum and maximum particle temperature values as boundaries. It can be clearly seen in Fig. 12A that the fixed bed case results in a larger particle temperature range, where a sharp peak is found corresponding with the particles that are not heated up by the high gas temperature front. The bottom material is captured by the long tail, which is not easy to distinguish. One additional data point taken at 750 s corresponds with the observation of the particle temperature standard deviation. The distribution becomes more narrow, where the lowest temperature peak corresponds with the bottom part and the highest peak with the top part which is fluidizing. This is also logical taking into account the temperature drop due to the evaporation of liquid. Eventually, the temperature difference will vanish leading also to a lower particle temperature standard deviation.

Fig. 12B and C are shown using an equal temperature range in order to make a comparison between the two gas velocity cases. However, one should note that the drying regime and bed walls could be different leading to a difficult one-to-one comparison. In both cases, it is observed that a hot tail is followed by a sharp peak which is in correspondence with the average particle temperature. This hot tail is larger in the 1.75 umf case which is in agreement with the earlier observed instantaneous image profiles. Comparing the 480 and 750 s it is also noted that the bed starts to slowly heat up again indicating that the internally limited drying regime is started.

The evolving particle temperature PDF for the 1.25 umf is hard to study due to the large temperature range. Therefore, an accumulative distribution is shown in Fig. 13 where the same five experimental times are shown. The bottom part of the bed contains relatively hot particles compared to the top zone before the inversion takes place. This is clearly visible in the graph, where the bottom zone is also increasing in size. This is in good correspondence with the snapshots and the observed increase in the particle temperature standard deviation. After the inversion, starting at 580 s, the distribution starts to become more narrow again. In that case, the standard deviation decreases again until a more narrow distribution is obtained as seen at 750 s. As mentioned before, these extremely large temperature distributions were not found in the fluidized bed cases.
Conclusions

Monodisperse porous γ-Al₂O₃ particles were dried in a pseudo-2D fluidized bed setup. A combined PIV and IRT technique was used to study the local particle hydrodynamic behavior and the particle temperature. The heat losses inside the setup were accounted for. It is observed that drying has a severe effect on the hydrodynamics which was quantified by the usage of time-averaged solids volume fluxes. The 1.25umf superficial velocity case resulted in a fixed bed case and after drying the porous material, a bed inversion took place between 580 to 605 s. The particle temperature is directly influenced by the hydrodynamic behavior since the 1.25umf case showed clear discrepancies compared to the other cases in terms of a higher particle temperature mean, a higher standard deviation and wider probability density functions. This clearly indicates that the fluidized bed heat and mass transfer characteristics are superior compared to fixed beds in these drying setups.

The hydrodynamics of the 1.50 and 1.75umf studies were also heavily influenced by the solids drying as an increase in bed expansion over time was observed. This indicates that a transition towards a more vigorous fluidization regime was made due to the liquid evaporation and thus the solids density reduction. Besides, the wet material caused clear asymmetrical time-averaged solids volume fluxes which were not found for the more dry material. This non-uniform bubble flow was mainly caused by
the solids density distribution due to inhomogeneous liquid evaporation. Therefore, we showed the strong two-way interaction between the heat and mass transfer rates and the bed hydrodynamics which has an important role for even larger superficial velocities.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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