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Research Paper

# Heat transfer during forest biomass particles drying in an agitated fluidised bed

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The phenomenon of convective heat transfer between gas and biomass particles during the drying process of fluidised bed material was analysed in order to obtain the heat transfer coefficients between the gas and the particle surface. In order to promote high homogeneity of the particles suspension, the bed was mechanically stirred, to obtain a uniform temperature inside the bed. The results showed a correlation between, the Nusselt and Reynolds numbers, which predicts the surface heat transfer coefficient with a deviation of ±15%, in relation to the experimental data.



## 1. Introduction

Drying is a process where momentum, heat and mass transport, all occur simultaneously. The reliability of the fluidised bed particles drier design largely depends on the information provided, particularly on the surface heat gas-solids transport phenomena. Thus, a study of the heat transfer phenomenon occurring during the period of constant drying rate for the moisture removal process of forest biomass particles in fluidised bed, was carried out.

For the case of a spherical particle located inside of an infinite gaseous fluid with velocity $U$, one of the most well-known experimental correlations suggested in the literature, is that obtained by Ranz and Marshall (1952), using dimensional analysis for a sphere-like shape liquid droplets falling into a gas:

$$Nu_{gp} = \frac{h_{gp} D_p}{k_g} = 2.0 + 0.60\left(\frac{\rho_g U D_p}{\mu_g}\right)^{1/2}\left(\frac{C_g \mu_g}{k_g}\right)^{1/3} \tag{1}$$

where: $Nu_{gp}$ is the Nusselt number; $h_{gp}$ is the gas-particle heat transfer coefficient in W m$^{-2}$ K$^{-1}$; $D_p$ is the diameter of the equivalent-volume sphere in m; $k_g$ is the gas thermal conductivity in W m$^{-1}$ K$^{-1}$; $\rho_g$ is the gas density in kg m$^{-3}$; $U$ is the superficial gas velocity in m s$^{-1}$; $\mu_g$ is the

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**Nomenclature**

| | |
|---|---|
| $C_g$ | gas heat capacity, J $kg^{-1}$ $K^{-1}$ |
| $d_{p,m}$ | weight mean diameter of biomass particle, m |
| $D_p$ | diameter of the equivalent-volume sphere, m |
| $h_{fg}$ | heat of vaporisation of the moisture, J $kg^{-1}$ |
| $h_{gp}$ | gas-particle heat transfer coefficient, W $m^{-2}$ $K^{-1}$ |
| $k_g$ | gas thermal conductivity, W $m^{-1}$ $K^{-1}$ |
| $L_a$ | stirrer height, m |
| N | agitation speed, rev $s^{-1}$ |
| $R^2$ | explanation variance |
| $S_p$ | particle surface area per unit volume of solids, $m^2$ $m^{-3}$ |
| t | time, s |
| $T_{wb}$ | wet bulb temperature, K |
| $T_{g,o}$ | oulet gas temperature in the bed, K |
| $T_{g,i}$ | inlet temperature in the bed, K |
| U | superficial gas velocity, m $s^{-1}$ |
| $U_{mf}$ | minimum fluidisation velocity, m $s^{-1}$ |
| w | moisture content in the biomass d.b., kg $kg^{-1}$ |
| $Nu_{gp}$ | Nusselt number, $h_{gp}$ $D_p/k_g$ |
| $Pr_g$ | Prandtl number, $C_g$ $\mu_g/k_g$ |
| $Re_p$ | Reynolds number, $\rho_g$ U $D_p/\mu_g$ |
| $\Delta T_{ml}$ | logarithmic-mean temperature difference, K |
| $\mu_g$ | gas viscosity, N s $m^{-2}$ |
| $\rho_g$ | gas density, kg $m^{-3}$ |
| $\rho_{p,0}$ | dry particles density, kg $m^{-3}$ |

gas viscosity in N s $m^{-2}$; and $C_g$ is the gas heat capacity, J $kg^{-1}$ $K^{-1}$.

In the analysis of particle beds, and especially in situations of high Reynolds numbers, other dimensionless equations whose form is similar to that above are also available. For example, the value of the convective heat transfer coefficient between the particles and the fluidising gas, to coarse particles ($Re_p > 100$) and in a fixed bed condition, can be calculated from the following correlation (Kunii & Levenspiel, 1969):

$$Nu_{gp} = 2 + 1.8 Pr_g^{1/3} Re_p^{1/2} \quad (2)$$

where: $Pr_g$ is the Prandtl number; and $Re_p$ is the Reynolds number.

When this equation is extrapolated to the low range Reynolds numbers, it shows large deviations relative to the experimental data reported by many authors, therefore the exact solution is not valid for the $Nu_{gp} < 2.0$. However, various experimental results for heat transfer under low Reynolds number conditions have correlated well with the Kunii and Levenspiel equation, obtained under the assumption of a plug flow regime:

$$Nu_{gp} = 0.03 Re_p^{1.3} \quad 0.1 < Re_p < 100 \quad (3)$$

Table 1 shows some correlations reported in the specialist literature on the dependence of the Nusselt number with Reynolds number for fluidised systems. According to Vyas and Nageshwar (1999), the results of these studies often show significant discrepancies between the different researchers.

**Table 1 – Correlation equations between Nusselt and Reynolds according several authors.**

| Authors | $Re_p$ | Equation |
|---|---|---|
| In Bird, Stewart, and Lightfoot (1960) | | |
| Ranz and Marshall | $1 < Re_p$ $Pr < 6 \times 10^4$ | $Nu_{gp} = 2 + 0.6\, Re_p^{1/2} Pr^{1/3}$ |
| Leva (1959) | 8–100 | $Nu_{gp} = 0.0063\, Re_p^{1.8}$ |
| Kunii and Levenspiel (1969) | 1.1–100 | $Nu_{gp} = 0.03\, Re_p^{1.3}$ |
| | >100 | $Nu_{gp} = 2 + 1.8\, Re_p^{1/2} Pr^{1/3}$ |
| Reyes and Alvarez (1990) | 33–150 | $Nu_{gp} = 0.00116\, Re_p^{1.52}$ |
| In Ciesielczyk (1996) | | |
| Lykov | | $Nu_{gp} = 0.0087\, Re_p^{0.84}$ |
| Frantz | 8–80 | $Nu_{gp} = 0.015\, Re_p^{1.6} Pr^{0.67}$ |
| Zabrodsky | | $Nu_{gp} = 0.00195\, Re_p^{1.46}$ |
| In Vyas and Nageshwar (1999) | | |
| Kettering et al. | 10–57 | $Nu_{gp} = 0.0135\, Re_p^{1.3}$ |
| Walton et al. | 10–32 | $Nu_{gp} = 0.0028\, Re_p^{1.7} (D_p/D_d)^{-0.2}$ |
| Frantz | 1.3–2.8 | $Nu_{gp} = 0.018\, Re_p^{1.2}$ |
| Rao and Sen Gupta | 7–20 | $Nu_{gp} = 0.000075\, Re_p^{1.61}$ |
| Vyas and Nageshwar (1999) | 0.6–80 | $Nu_{gp} = 0.044\, Re_p^{1.13} Pr^{0.33}$ |

The differences between correlations are due to the type of regime assumptions, either stationary or transient (according to Vyas and Nageshwar this is the main reason), type of flow pattern assumed (plug flow or back-mix flow) and the way the gas and the particles temperatures used in the calculation equations were measured. Also, some parameters are difficult to measure and there is no certainty that the results of a particular author are applicable to all the systems.

With regard to biomass-air fluidised bed system, there is no information available in relation to the $h_{gp}$ heat transfer coefficients values. Alvarez and Shene (1994) determined heat transfer volumetric coefficients in sawdust rotatory dryers, but they should not be applied to fluid bed dryers due to the very different interaction conditions and material flows that take place. Recent studies and results (Abdel-Aziz, El-Abd, & Bassyouni, 2016; Abid, Ali, & Alzubaidi, 2011; Chen, Grace, & Golriz, 2005; Papadikis, Gu, & Bridgwater, 2010; Salve, Pande, & Khan, 2014) are available, but they are designed for very different systems than those applied for low density biomass particles.

Thus, the biomass dryers design based on the heat transfer coefficients obtained from correlations already reported, do not seem reliable because which of the correlations agree better with the particular studied system. According to Vyas and Nageshwar, the only true fact of all these studies is that the slope of the Nusselt number curves against Reynolds number varies from 1 to 1.7 and is close to 1.5, for most of the correlations.

## 2. Material and methods

### 2.1. Methods

For heat transfer coefficient studies in drying systems at the constant rate drying period, it can be assumed that the particle temperature is the same as the wet bulb temperature of the inlet air (Ciesielczyk, 1996; Reyes & Alvarez, 1990;

Zabrodsky, 1966), since the product displays a wet surface, during that period.

The analysis of the drying curves obtained by Moreno (2005), allows concluding that most of the process takes place under a constant temperature regime and at a constant drying rate. For dryer design, it is appropriate to work with mean values of $h_{gp}$ in an adiabatic regime so the heat balance equation can be written as:

$$-h_{fg}\rho_{p,0}\frac{dw}{dt} = h_{gp}S_p\Delta T_{ml} \tag{4}$$

where: $h_{fg}$ is the heat of vaporisation of the moisture in J kg$^{-1}$; $\rho_{p,0}$ is the dry particles density in kg m$^{-3}$; $w$ is the moisture content in the biomass dry base. in kg kg$^{-1}$; $t$ is the time in s; $S_p$ is the particle surface area per unit volume of solids in m$^2$ m$^{-3}$; and $\Delta T_{ml}$ is the logarithmic-mean temperature difference in K.

Thus, the $h_{gp}$ calculation can be carried out using the equation:

$$h_{gp} = \frac{h_{fg}\rho_{p,0}\left(-\frac{dw}{dt}\right)}{S_p\Delta T_{ml}} \tag{5}$$

where, the mean logarithmic temperature difference value, assuming that the particle surface temperature is equal to the wet bulb temperature of the air and can be determined as:

$$\Delta T_{ml} = \frac{T_{g,i} - T_{g,o}}{\ln\left(\frac{T_{g,i}-T_{wb}}{T_{g,o}-T_{wb}}\right)} \tag{6}$$

where: $T_{g,i}$ is the inlet temperature in the bed in K; $T_{g,o}$ is the outlet gas temperature in the bed in K; and $T_{wb}$ is the wet bulb temperature in K.

According to Eq. (5), the experimental analysis of heat transfer coefficients, requires the particle surfaces be in contact with the fluidizing gas; by means of an aerodynamics analysis of the biomass particles in contact with air, in a previous work, Moreno, Antolín, and Reyes (2009) determined the specific particle surface $S_{p,}$ used to calculate the film coefficients of heat and mass convective transfer:

$$S_p = 6737d_p^{-0.1237};\ R^2 = 0.960;\ w = 0.15\ \text{kg}\cdot\text{kg}^{-1};\ 0.89\ \text{mm} \le d_p \le 3.56\ \text{mm} \tag{7}$$

and

$$S_p = 3778\ d_p^{-0.2432};\ R^2 = 0.972;\ w = 2.0\ \text{kg}\cdot\text{kg}^{-1};\ 1.44\ \text{mm} \le d_p \le 3.56\ \text{mm} \tag{8}$$

where: $d_{p,m}$ is the weighted mean diameter of biomass particle in m; $R^2$ is the explanation of variance (coefficient of determination).

### 2.2. *Experimental work*

The experiments were carried out in a fluid bed particles dryer, shown in Fig. 1. The drying chamber had a circular section diameter of 0.3 m. The distributor size (0.071 m$^2$) places the equipment in an intermediate size between a laboratory and pilot size equipment, according to the criteria provided by Strumillo and Kudra (1986). For a bed height of 0.17 m, the chamber can be loaded with approximately 2 kg of *Pinus radiata D. Don* biomass particles.

The chamber had an inside agitator system built in steel with a vertical axis and two mixers with two blades each. The mechanism was coupled to a motor allowing changing the rotation velocity of the agitator inside the bed. In this study the mechanism was designed with a 15 mm $L_a$ height, to avoid entrapment of particles in the area near the distributor.

Bed temperatures were measured with eight type T thermocouples, placed at different points of the bed. Figure 2 shows four of the eight thermocouples ($T_1$ to $T_8$) inserted in the bed. T-type sensors were incorporated to measure the room temperature ($T_9$) as well as the air temperatures below the distributor ($T_{10}$), which corresponded to the operating temperature. The thermocouples, in addition to providing bed temperature variations, by recording temperatures and their differences with the time it could detect either possible dead zones or channeling.

Temperature sensors were installed along with a data acquisition system, consisting of a Digi-Sense Scanning Thermometer of 12 channels (Cole-Parmer Instrument Company, Model 92000-05, Barrington, IL USA), resolution 0.1 K and ±0.5 K accuracy, with a RS-232 output to a PC, to collect and to analyse the information in Windows using a ScanLink 2.0 software. Data was collected every 4 s.

The dryer temperature control system consists of a PT100 temperature sensor placed in the pre-chamber, under the distributor and connected to a fuzzy logic micro-processor controller.

Experiments to obtain the $h_{gp}$ coefficient were carried out in a biomass particles bed, where the inlet temperature data to the dryer and of the gas outlet, were recorded during drying. Therefore, it was necessary to place thermocouples at the inlet and outlet of the bed. The gas sensor on the exit of the bed or freeboard, was protected with a filter during each drying test to prevent particles influencing the outlet gas temperature measurement.

During the experiments, temperature and gas relative humidity were recorded using a Digi-Sense Temperature/Humidity Logger and ScanLink 2.0 and PCDAC Program (Cole-Parmer Instrument Company, Model 91090-00, Barrington, IL USA) Data was obtained every 10 s and transferred to the computer for analysis.

Moreover, to avoid errors caused by the heat transferred from the air to the distributor, the gas temperature was measured at the inlet of the bed, i.e., at a point just above the air distributor.

Gas temperature following its passage through the distributor had a rapid temperature drop when contacting the particles and reaching equilibrium at a height above the distributor <20 mm (Moreno, 2005), a result that is in line with other reports (Kunii & Levenspiel, 1969; Vanecek, Markvart, & Drbohlav, 1966) and confirmed with further experimental results from Temple and van Boxtel (1999) and Watano, Yeh, and Miyanami (1999).

In terms of the physical properties of the fluid presented in the dimensionless numbers, they were evaluated at the fluid film mean temperature surrounding the particle. The particle

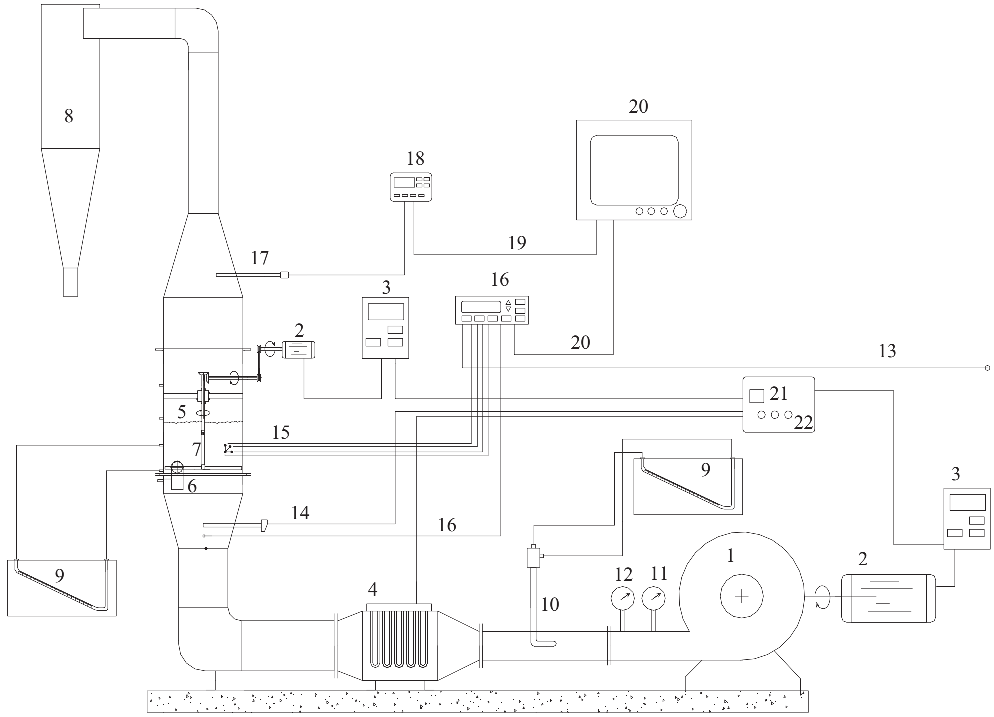


**Fig. 1 — Schematic diagram of experimental equipment: (1) air blower; (2) motor; (3) frequency converter; (4) air heater; (5) drying chamber; (6) air distributor; (7) agitator; (8) cyclone; (9) water manometer; (10) Pitot tube; (11) Bourdon manometer; (12) thermometer; (13) room thermocouple probe; (14) PT100 probe; (15) drying and bed temperature probes; (16) scanning thermometer; (17) temperature/relative humidity probe; (18) humidity logger; (19) RS-232 cable; (20) computer; (21) fuzzy-logic micro-processor, and (22) electric power.**

temperature was considered equal to the wet bulb air temperature.

To obtain dimensionless equations requires a series of experimental data obtained under different process conditions. For the experimental design it was considered that the heat transfer coefficient depend mainly on the superficial velocity and particle size; in a phenomenon of forced convection the dimensionless treatment of the heat transfer may be represented through the correlation type:

$$Nu_{gp} = \frac{h_{gp} D_p}{k_g} = a\left(\frac{\rho_g U D_p}{\mu_g}\right)^m \left(\frac{C_g \mu_g}{k_g}\right)^n \tag{9}$$

Regarding the geometric implications of this study, the system analysed here belongs to the category of called *large particle fluidized bed* ($d_p$ > 0.5 mm) with the characteristic of being subjected to greater air flows required to maintain particles in suspension.

In a previous study (Moreno, Antolín, & Reyes, 2004), it was verified that this biomass particle system can be classified as Geldart type B and D. These large particle systems are not well studied in the literature, except during the 80's by Adams (1984), and oriented towards bed-surface heat transfer and not solid-fluid heat transfer. To run a system with large particles implies that Reynolds number are ≥100; most correlations in the literature are below this limit.

It is not clear that the rotation velocity of the agitator $N$ parameter has an important influence since, according to the results obtained by Moreno, Antolín, and Reyes (2007), the effect of $N$ on the evaporation rate is very low. However, the agitator has an important effect on the uniformity of the fluidisation quality and temperature uniformity in order to minimise the typical anomalies found in the fluidisation of type B and D solids (i.e. channelling and spouting). With moisture contents above 30%, in a conventional fluidised beds, a fraction of the particles agglomerate, impoverish the air-solid contact, thereby reducing the heat transfer coefficients. In this regard, the stirrer decreases the possible agglomeration, thus minimising their effect on the heat transfer coefficient.

Thus, for the heat transfer phenomenon there are three controlling factors ($U$, $d_p$ and $N$). Experiments were carried out considering each particle size with an air superficial velocity within the range in which fluidisation occurs. Experiments for each pair ($d_p$-$U$) were performed with two agitator speeds to study possible variations of $h_{gp}$ with $N$. Each of the 10 experiments with their respective levels to determine the heat transfer coefficient, are shown in Table 2. The operating temperature in all tests was set at a constant value of 423 K.

Because operating air velocities during drying are related to the particle size, the classical $2^n$ experimental design was not carried out, because this will drag out the smallest particles from the drying chamber. To select the values of the superficial velocity, the minimum fluidisation velocity for each particle size was obtained experimentally in a

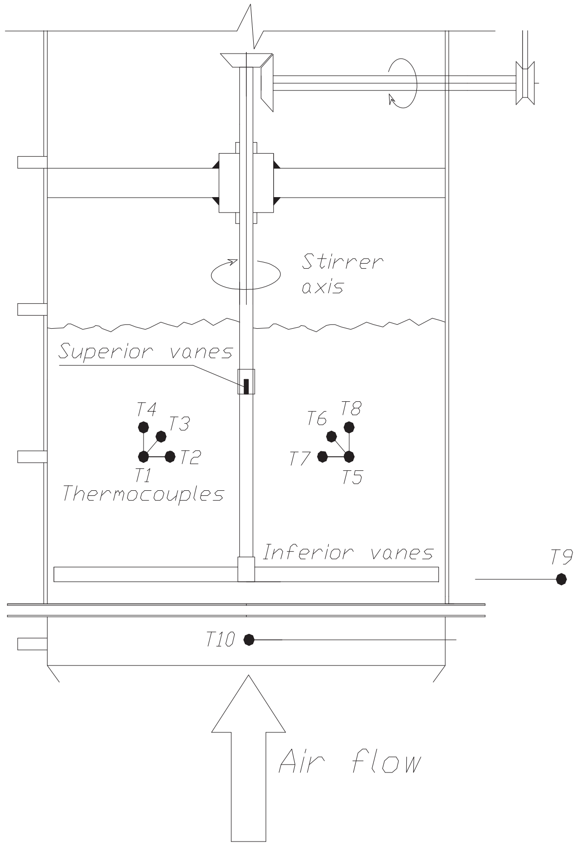


**Fig. 2 – Drying chamber with agitation system and temperature probes.**

stirred fluidised bed and they are shown in the last column of Table 2.

## 3. Results and discussion

The high uniformity of bed temperature was checked in curve temperature evolution during drying (Fig. 3), obtained with eight sensors placed in the bed, as shown in Fig. 2. From the point of view of kinetics drying, this translates in a bed of high quality with a very regular drying curve, as shown in Fig. 4. This regularity of the drying curve allows to obtain highly reliable data with a drying kinetics coefficient $R^2 = 0.997$.

**Table 2 – Experiments for determining the coefficient $h_{gp}$.**

| Trial | Randomness | $d_p$ (mm) | U (m $s^{-1}$) | N ($s^{-1}$) | $U_{mf}$ (m $s^{-1}$) |
|---|---|---|---|---|---|
| 1 | 6 | 0.89 | 0.71 | 1 | 0.46 |
| 2 | 8 | 0.89 | 0.71 | 2 | 0.46 |
| 3 | 9 | 1.44 | 0.77 | 1 | 0.56 |
| 4 | 2 | 1.44 | 0.77 | 2 | 0.56 |
| 5 | 5 | 1.85 | 0.81 | 1 | 0.66 |
| 6 | 10 | 1.85 | 0.81 | 2 | 0.66 |
| 7 | 3 | 2.18 | 0.82 | 1 | 0,80 |
| 8 | 7 | 2.18 | 0.82 | 2 | 0.80 |
| 9 | 1 | 3.56 | 1.02 | 1 | 0.95 |
| 10 | 4 | 3.56 | 1.02 | 2 | 0.95 |

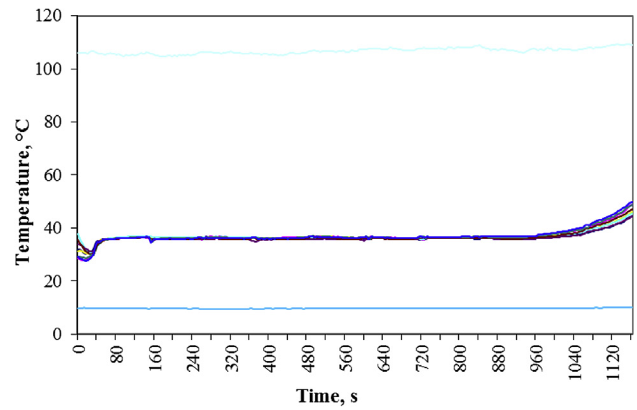


**Fig. 3 – Evolution of temperature at various locations in the bed during biomass drying in an agitated fluidised bed.**

Table 3 shows the results obtained in this study, for each of the tests performed. The $R^2$ value indicated in the table corresponds to the correlation coefficient used to obtain the evaporation rate $dw/dt$ from the experimental drying curve, which is used together with Eq. (5) to determine the $h_{gp}$ coefficient.

Figure 5 shows the correlation obtained by the adjustment between the Nusselt and Reynolds numbers. It has been confirmed that the rotational speed factor (N) does not influence the convective heat transfer coefficient. The variations found in the $h_{gp}$ coefficient with $N$ are at random as shown in Table 3, thus confirming the results reported in a previous study (Moreno et al. 2007).

Therefore, a correlation to calculate the heat transfer coefficient in forest biomass fluidized bed is proposed, as follows:

$$Nu_{gp} = 0.003 Re_p^{1.28}; \quad R^2 = 0.95 \text{ for } 100 < Re_p < 250 \tag{10}$$

or by rearranging terms as,

$$h_{gp} = 0.003 \frac{k_g}{D_p} \left( \frac{\rho_g U D_p}{\mu_g} \right)^{1.28} \tag{11}$$

Of all the previously published correlations analysed in this study, the ones that come the closest to the one proposed here, as seen in Fig. 6 are those from Reyes and Alvarez (1990) and Zabrodsky (as reported in Ciesielczyk (1996)).

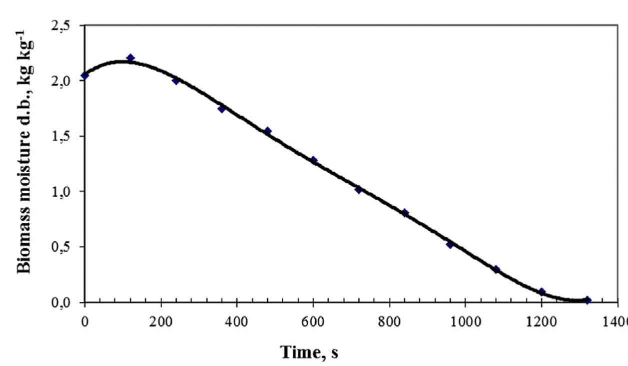


**Fig. 4 – Biomass drying curve in an agitated fluidised bed.**

**Table 3 – Experimental results in determining the coefficient $h_{gp}$.**

| Trial | $d_p$ (mm) | $D_p$ (mm) | $Re_p$ | dw/dt (kg kg$^{-1}$ min) | $R^2$ | $h_{gp}$ (W m$^{-2}$ K$^{-1}$) | $Nu_{gp}$ |
|---|---|---|---|---|---|---|---|
| 1 | 0.89 | 2.44 | 102 | 0.067 | 0.99 | 15.4 | 1.33 |
| 2 | 0.89 | 2.44 | 103 | 0.565 | 0.994 | 13.0 | 1.1 |
| 3 | 1.44 | 2.7 | 124 | 0.0553 | 0.986 | 13.2 | 1.26 |
| 4 | 1.44 | 2.7 | 124 | 0.0586 | 0.997 | 14.5 | 1.39 |
| 5 | 1.85 | 2.84 | 136 | 0.064 | 0.985 | 15.7 | 1.57 |
| 6 | 1.85 | 2.84 | 134 | 0.0631 | 0.996 | 14.9 | 1.5 |
| 7 | 2.18 | 3.13 | 148 | 0.0773 | 0.973 | 18.2 | 2 |
| 8 | 2.18 | 3.13 | 150 | 0.0645 | 0.996 | 14.4 | 1.6 |
| 9 | 3.56 | 4.3 | 256 | 0.0957 | 0.985 | 23.2 | 3.52 |
| 10 | 3.56 | 4.3 | 257 | 0.0992 | 0.99 | 25.7 | 3.9 |

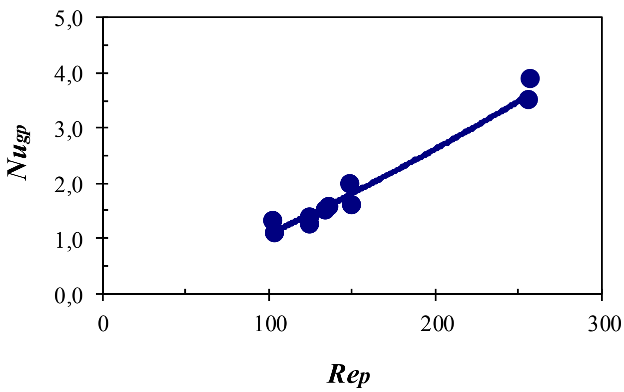


**Fig. 5 – Nusselt–Reynolds correlation for a forest biomass fluidised bed according to Eq. (10). $R^2 = 0.9473$.**

Figure 7 shows the degree of reliability of the correlation proposed to determine the heat transfer coefficient $h_{gp}$. It is worth noting that the values found in the $h_{gp}$ coefficient in this study that fluctuate between 13 and 25.7 W m$^{-2}$ K$^{-1}$, as shown in Table 3 are within the range 6 and 23 W m$^{-2}$ K$^{-1}$ reported by Botterill (1975) and other similar fluidised bed data reports. The use of other correlations found in the literature provides heat transfer coefficient values with large deviations from those found, experimentally. Renström and Berghel (2002) published

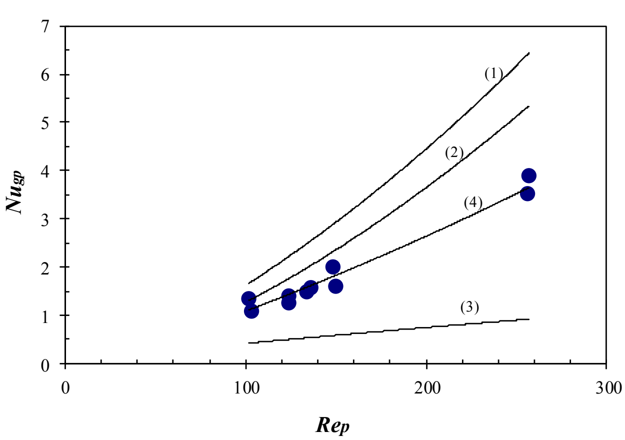


**Fig. 6 – Comparison of the results with those from other authors. (1) Zabrodsky (in Ciesielczyk, 1996); (2) Reyes and Alvarez (1990); (3) Lykov (in Ciesielczyk, 1996); (4) Eq. (10).**

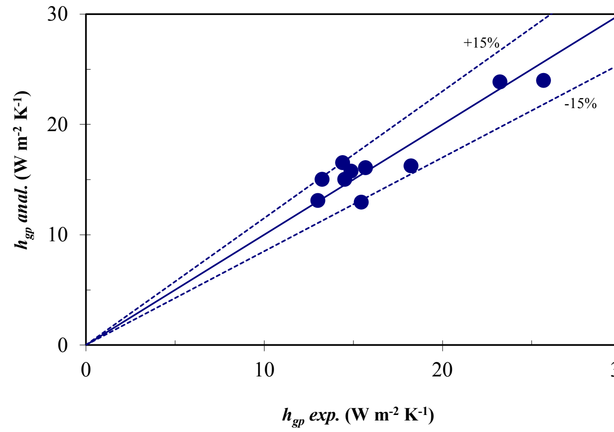


**Fig. 7 – Comparison of the experimental $h_{gp}$ values with values calculated with the proposed correlation.**

drying coefficients results for forest biomass particles, with values ranging between 10 and 60 W m$^{-2}$ K$^{-1}$. However, these were obtained in a bed dryer and using overheated steam with temperatures up to 513 K. In a recent study (Salve et al. 2014), higher $h_{gp}$ coefficient values (80–220 W m$^{-2}$ K$^{-1}$) were provided; however they were obtained for higher superficial velocities (3–11 m s$^{-1}$) and working with sand.

The large discrepancies found, as already discussed by others, are often due to the nature of the solids, to the technique for measuring the temperature differential in the solid-fluid interface and the choce of flow model (plug flow or whole mix). In this particular case, it can be stated that the fluidised system studied here, consisting of forest biomass particles, corresponded to a large particle size system with a higher Reynolds number range. Most correlations published in the literature, as shown in Table 1, were obtained for $Re_p < 100$ and in our case, the imposed flow conditions were within the range $102 < Re_p < 257$.

## 4. Conclusions

The convection heat transfer coefficient between the fluidising and the solids, was experimentally determined and varied between 13 and 25.7 W m$^{-2}$ K$^{-1}$. Based on these results, a dimensionless correlation between the Nusselt and Reynolds numbers was suggested to calculate the heat transfer coefficient of fluidised forest biomass. The sudden change in the drying air temperature at the distributor as the air tries to reach equilibrium, as well as the small differences between the gas and particle temperature (the latter assumed to be equal to the wet-bulb), have confirmed the existence of a whole mixture flow pattern, which was expected to occur due to the use of a fluidised bed.

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