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A KINETIC MODEL FOR DRYING OF MUSTARD SEEDS IN BATCH FLUIDIZED BED

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Abstract: Mustard one of the popular oil seeds is investigated for drying in Batch fluidized beds. Experiments were conducted to assess the kinetics of drying for the variation in the inlet air temperature, the inlet air flow rate and the solids holdup in the fluidized bed. The drying rate was found to increase significantly with increase in temperature and with flow rate of the heating medium, while decrease with increase in solids holdup. The duration of constant rate period was found to be insignificant, considering the total duration of drying. The drying rate was compared with various exponential time decay models and the model parameters were evaluated. The page model was found to match the experimental data very closely with the maximum root mean square error (RMSE) of less the 2.0%. The experimental data were also modeled using Fick's diffusion equation and the effective diffusivity coefficients was found to be within $1.69x10^{-11}$ to $3.26x10\text{-}11m^2/s$ for the range of experimental data covered in the present studywith RMSE less than 4%.

Keywords: Fluidized Bed, Drying Kinetics, Food Grain Drying

I. INTRODUCTION

Mustard (Brassica juncea) is one of the principal oil crops in India cultivated over seven million hectares annually. Mustard oil is one of the popular vegetable oils widely used in many parts of India traditionally. Plants are generally harvested before fruits are fully ripe to reduce shattering. Conventionally, entire plant is pulled out, tied into small sheaves and dried in the sun for 4–10 days [1]. The moisture content of the harvested mustard oil seeds must be reduced to less than 10% moisture content for safe storage.

Fluidized beds find increasing application in the drying of agricultural materials although they are being widely in use, in industries for drying of fertilizers, chemicals, pharmaceuticals and minerals. The increasing application of fluidized bed drying for agricultural materials is due to the evolving designs of fluidized bed, for fluidization of coarse material, which are rather difficult to fluidize. Fluidized beds as compared to other modes of drying offer advantages such as high heat capacity of the bed, improved rates of heat and mass transfer between the phases and ease in handling and transport of fluidized solids.

Knowledge of drying kinetics is essential for the design of dryers. The complex hydrodynamics and process calculations are material, dryer specific, rendering development of numerous mathematical models for drying kinetics. These range from analytical models solved with a variety of simplified assumptions to purely empirical models, often built by regression of experimental data.

The drying rates in fluidized beds are strongly influenced by the characteristics of the material and the conditions of fluidization. Drying of solids is generally understood to follow two distinct drying zones known as constant rate period and falling rate period demarcated with critical moisture content. The critical moisture content is reported to vary with operating parameters and with the type of drying equipment. The constant rate period is understood to have maximum drying rate, which remains constant until the critical moisture content with the resistance for moisture transfer in the gas phase. The rate of diffusion of moisture to surface of solids becomes the limiting factor for moisture transfer as far as the falling rate period is concerned. The extent of drying zones are decided based on the type of material, with materials like sand, ion exchange resin, glass beads etc. reported to have larger duration of constant rate period and short linear falling rate period while the fibrous material or food grains are reported to have a very short duration constant rate period and a longer curvy-linear falling rate period [2, 3].

In general the drying rate in constant rate period in fluidized bed dryingare modeled using i) simple empirical correlation relating drying rate to the influencing parameters, or utilizing heat/mass transfer coefficient between solids and gas in fluidized bed [4-8] and ii) mass/heat transfer models, assuming the bed to be made of bubble phase, emulsion phase and a dense phase with the exchange of mass and energy between these phases [9-15]. Similarly drying kinetics in falling rate period is modeled with complex models, which serve to improve the understanding of drying process. However, these models may not serve forpractical applications in a straightforward manner, due to their complexity [16].

Simple models that can be used to design drying systems are much sought after to provide an optimum solution to different aspects of drying operation, with a minimum number of parameter. A series of simple models based on exponential time decay were developed in the past and are being continuously revised/improved, which are popularly known as Newton model, Page model, Henderson and Pabis model, Two term exponential model and Approximate diffusion model. These simple models are recently utilized for drying application by Mujumdar [17], Diamante and Munro [18], Zhang and Litchfield [19], Henderson [20], Yaldeiz and Ertekin [21], Sharaf-Eldeen, et al. [22] to represent the drying kinetics. On the other hand, complex models are based on the popular Fick's diffusion equation, to estimate the drying kinetics. The predicated drying kinetics using the model, varied depending on the applied boundary conditions, often requiring numerical computations [8, 15, 23-26].

The objective of the present study is to experimentally investigate the drying kinetics of mustard in fluidized bed, with respect to the operating parameters such as the temperature, flow rate of the drying medium and the solids holdup. Although the effect of operating parameters on the drying rates are well known and one expects the influencing parameters to respond in similar fashion qualitatively, the drying kinetics can vary quantitatively depending on the nature of the material and the drying conditions. The present study further attempted to verify the compatibility of experimental drying kinetics with various simple models reported in literature [Table 2], and with complex models such as Fick's diffusion equation. The model parameters are estimated by minimizing the root mean sum of square of error (RMSE) between the experimental drying rate andthe model prediction [Table 3, 4].

II. MATERIALS AND EXPERIMENTAL METHODS

Drying experiments were conducted using a fluidization column of 0.245m in internal diameter with a height of 0.6m. The gas distributor was 2mm thick with 2mm perforations having 13% free area. A fine wire mesh was spot welded over the distributor plate to arrest the flow of solids from the fluidized bed in to the air chamber. Air blower with volumetric discharge capacity of 200m^3/h was used. Air from the blower was metered using a calibrated orifice meter, before being heated and fed to the fluidization column, through the air chamber. The electrical heater consisted of a multiple heating element each of 2KW rating. A temperature controller, provided to the air chamber, facilitated control of airtemperature within ±2°C of the set temperature.

Figure:1 Fluidized bed dryer set-up

Table 1 shows the physical characters of mustard as well as the experimental conditions covered in the present study. A good fluidization behavior in terms of perfect mixing of the bed material was observed visibly. This was substantiated with low fluctuation in the bed pressure drop, which is an indication for smooth fluidization without formation slugs.

Table 1: Characteristics of the Mustard and ExperimentalParameters in the Present Study

Name of Material	**Mustard (Brassica juncea)**
Shape of Material	Spherical
Particle density, kg/m^3	1100
Temperature of fluidizing air, ^{0}C	40, 50, 60
Fluidizing air velocity, m/s	1.32, 1.05
Solid holdup, kg	0.1,0.125

Air at desired temperature and flow rate was allowed to flow through the fluidization column. A known quantity of mustard with known initial moisture content was introduced into the column after ensuring the steady temperature and the air flow rate. Samples of mustard, weighing approximately around 1gram, were scooped out of the bed at regular intervals of time for estimation of moisture. The moisture content of mustard was determined by drying the samplestill constant weight in an air oven at 105°C. The moisture contents were expressed on dry basis as kilograms of moisture per kilogram of dry solid. The experimental data was checked for reproducibility, and were found to deviate within ±2%. The equilibrium moisture content was estimated by keeping the samples in a humidity controlled air chamber at the desired temperatures until no further weight change.

III. RESULTS AND DISCUSSION

Experimental data generated in the present study were depicted as plots of relativemoisture content X/X_i Vs time in the figures 1 to 4. The figures address the effect of operating variables such as the temperature, flow rate of the heating medium and the solids holdup. In general it can be observed from the figures, that the duration of constant drying rate period is insignificant compared to the total drying time. The rate of drying is higher at the early stage of drying while the moisture content was high and reduces as the moisture content decreases.

Figure 2 and 3 shows the effect of temperature of the heating medium at two different solids holdup. An increase in temperature of the heating medium increases the drying rate and it can be attributed to the higher bed temperature of particles in the bed, which increases the intra particle moisture diffusion leading to a higher drying rate. The increased transport properties of the fluids with increase in temperatures is well known and the experimental data are in concurrence with the basic concepts of mass transfer.

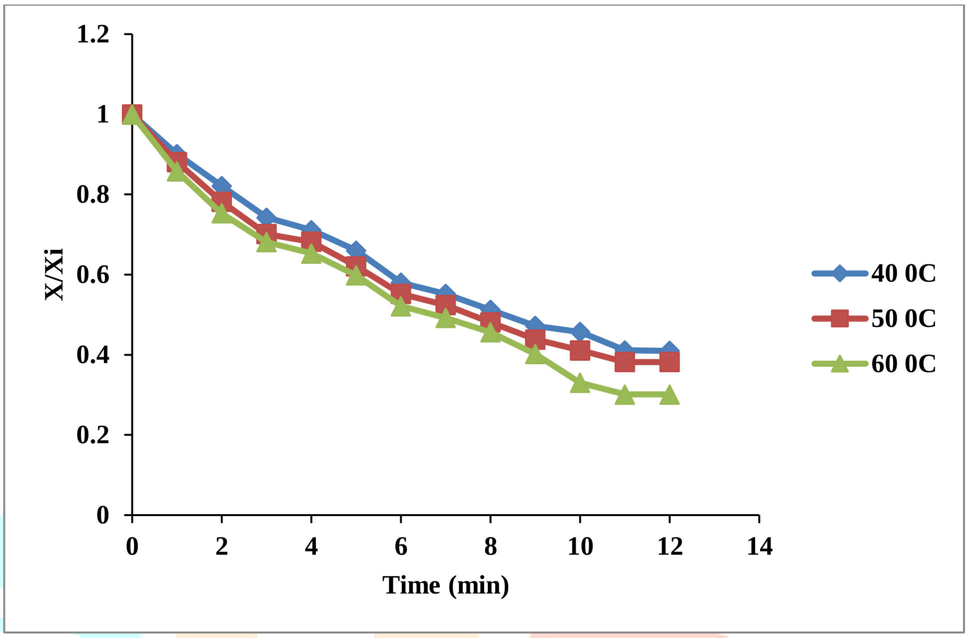

Figure 2: Effect of Temperature of the heating medium [U: 1.32m/s; W: 0.100 kg]

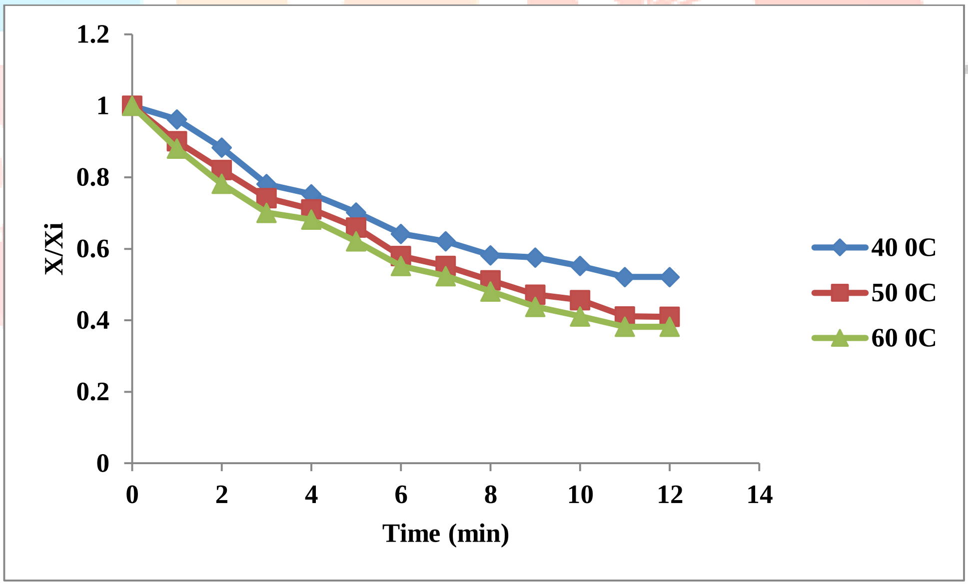

Figure 3: Effect of Temperature of the heating medium [U: 1.32m/s; W:0.125 kg]

Figure 4 shows an increase in drying rate with air flow rate. In general an increase in flow rate of the heating medium reduces the external resistance for mass transfer, aiding a higher transfer rate. Since the intra particle moisture diffusion controls drying rate during falling rate period, the decrease in external resistance for mass transfer is not expected to increase the rate of drying significantly. However an increase in the flow rate of the heating medium results in dispersion of diffused moisture from solids in a larger volume of inlet air,which results in higher bed temperature. The higher bed temperature increases the moisture diffusion rate, resulting in an increased drying rate. This was evidenced by continuous recording of the bed temperature which indicated ahigher bed temperature at higher air flow rates.

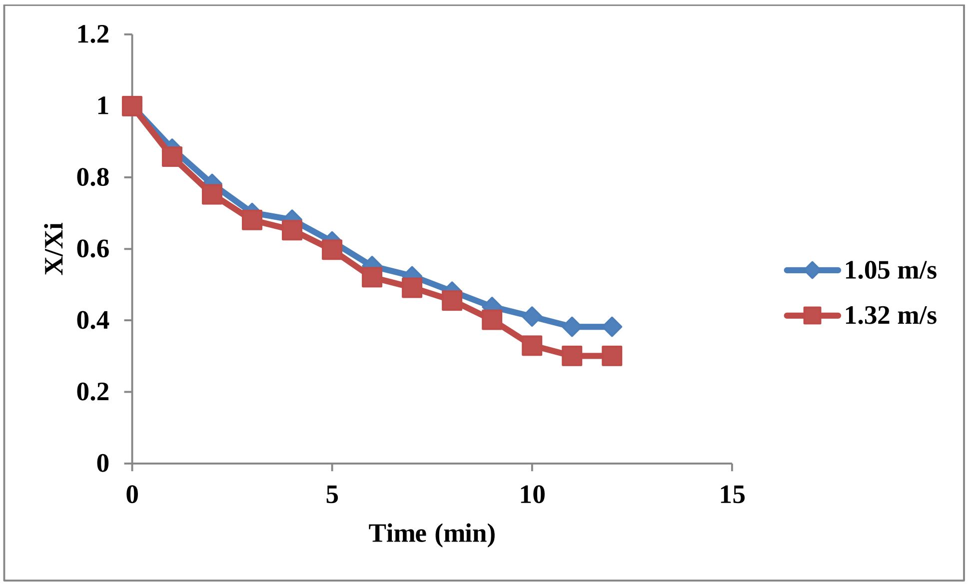

Figure 4: Effect of flow rate of the heating medium [T: 60 ^{0}C; W: 0.100 kg]

An increase in the solids holdup is found to decrease the drying rate (Figure 5) and it can again be attributed to the lower effective bed temperature at higher solids holdup. An increase in solids holdup increases the surface area of contact between the heating medium and solids proportionally. With the external conditions remaining unaltered, the higher area of contact between the phases enhances the transfer rate and hence a larger quantity of moisture diffuses fromthe solids resulting in a lower bed temperature. The larger contact area increases the drying rate while the lower bed temperature reduces the drying rate. Both the competing process results in lowering of drying rate with increase in solids holdup. The lower bed temperature is evidenced from the continuous recordingof the bed temperature during drying. All the three observations are in qualitative agreement with most of the earlier observations reported in literature [8, 15, 27, 28]. However Topuz et al.[15] reported a reduction in drying rate with increase in flow rate of the heating medium, which was attributed to poor contact between thesolid and gas phase, due to spouting of the bed at higher flow rate.

The experimental drying data's were converted to dimensionless moistureratio, $MR = \frac{X-X_e}{X_i-X_e}$ for the sake of comparison with the various models.List of simple exponential time decay models, popularly known as Newton model, Page model, Henderson and Pabis model, Approximation of diffusion model, Two term exponential model as listed in Table 2, were compared with the experimental data.

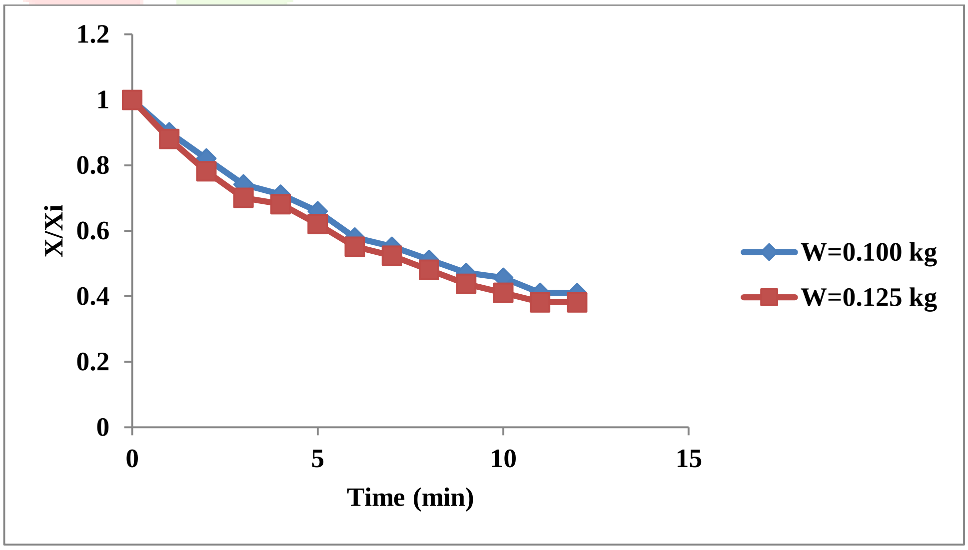

Figure 5: Effect of solids holdup [T: 50 ^{0}C, U: 1.32 m/s]

Table 2: List of Various Simple Models Tested with the Drying Data of thePresent Study

Name of Model	Model Equation
Newton Model (**NM**)	MR = exp (-kt)
Page Model (**PM**)	MR = exp (-kt^n)
Henderson and Pabis (**HPB**)	MR = a exp (-kt)

The experimental drying rates were fitted with various model equations,by minimizing the Root Mean Square Error (RMSE) between the experimental drying rate and the model equation. The RMSE is defined as,

$$RMSE = \left[\frac{1}{N} \sum_{i=1}^{n} (MR_{pre,i} - MR_{\exp,i})^2 \right]^{0.5} X100 \qquad (1)$$

The RMSE values were found to be more than 6% for all the models except the Page model and the approximate diffusion model. Table 3 compares the model parameters along with the RMSE values for Page model and approximate diffusion model. It can be seen from Table 3 that the Page model is found to match experimental data very closely, with the RMSE error less than 2.0%. The standard deviation between the experimental data and the model prediction using the page model is less than 5%. Although the approximate diffusion model is a three parameter model, the RMSE values are higher than the Page model, confirming the suitability of the Page model with the experimental data. Figure 5 shows the proximity of the Page model with the experimental data. The model parameters can be utilized to estimate the drying time as well as for designing andscale up of the drying process.

Table 3: Evaluated Model Parameters at Various Operating Conditions

T, C	W,kg	U,m/s	PM			ADM			
			k X10^3	n X10	RMSE	a X10	k X10	b X10^3	RMSE
40	0.1	1.32	15.79	5.72	1.16	2.32	4.38	1.09	3.31
40	0.125	1.32	3.74	7.43	1.39	1.03	4.46	1.07	3.27
50	0.1	1.32	19.14	5.59	1.36	2.60	3.41	1.53	3.28
50	0.125	1.32	8.91	6.44	1.34	1.75	4.60	1.09	3.77
60	0.1	1.32	15.0	5.97	2.00	2.26	3.57	1.65	4.83
60	0.125	1.05	6.22	6.98	1.30	1.46	2.71	2.05	3.27

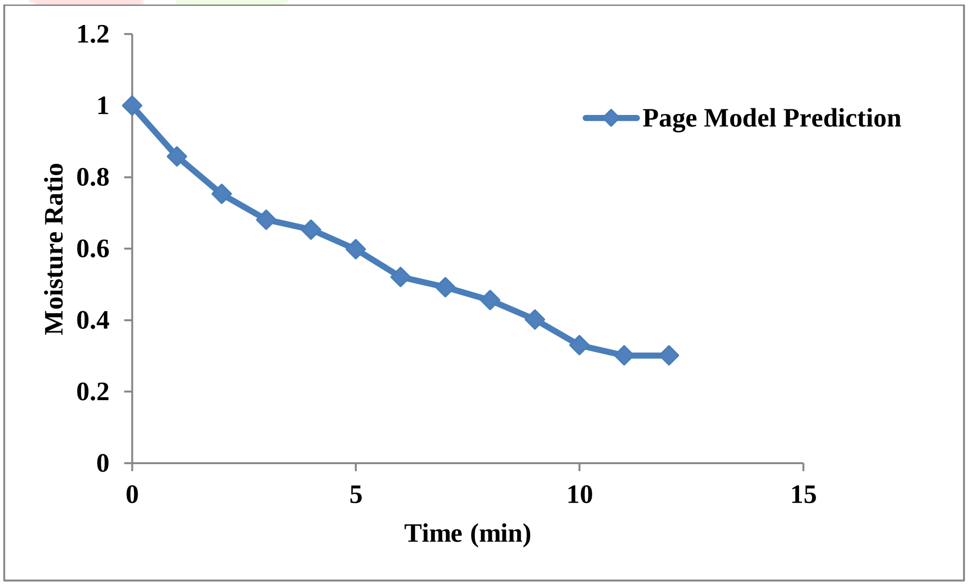

Figure 6: Comparison of Page model prediction with the experimental data [T: 40 ^{0}C, U: 1.32 m/s, W: 0.100 kg]

It is also attempted to model the experimental data with a fundamental diffusion equation for moisture distribution within the solid particle, with appropriate boundary conditions. The model assumes that the moisture diffuses from inside the particle to the surface of the particle and evaporates at the surface and that all the particles are uniform in size and spherical in shape. The fluidized beds are perfectly mixed beds, and the solids at any point in the bed are exposedto same drying conditions. The general form of the diffusion equation known as Fick's diffusion equation is,

$$\frac{\delta C}{\delta t} = D_{eff}\left[\frac{\delta^2 C}{\delta r^2} + \frac{2}{r}\frac{\delta C}{\delta r}\right] \qquad (2)$$

the boundary conditions are,

at	$t = 0$;	$0<r<R_s$;	$C = C_i$
at	$t > 0$;	$r = 0$;	$\delta C/\delta r = 0$
at	$t > 0$;	$r = R_s$;	$-D_{eff}(\delta C/\delta r) = K(C_{sj} - C_{be})$

where C_{sj} is the moisture concentration just within the sphere and C_{be} is the concentration required to maintain equilibrium with the surrounding atmosphere. Analytical solution to equation (2) for the above boundary conditions was provided by Crank [29] as given below,

$$\frac{C - C_e}{C_i - C_e} = \sum_{n=1}^{\infty} \frac{6 Bi_m^2 \exp\left(-\beta_n^2 D_{eff} t / R_s^2\right)}{\beta_n^2\left(\beta_n^2 + Bi_m(Bi_m - 1)\right)} \qquad (3)$$

where β_n are the roots of the equation,

$$\beta_n \cot \beta_n + Bi_m - 1 = 0 \qquad (4)$$

The mass Biot number (Bi_m) is defined as $K R_s/D_{eff}$ and the mass transfer coefficient (K) is calculated based on the equation due to Richardson and Szekely [30] as given below,

$$Sh = \frac{Kd_p}{D} = \begin{matrix} 0.374\,\mathrm{Re}^{1.16} & for\ 0.1 < \mathrm{Re} < 15 \\ 2.01\,\mathrm{Re}^{0.5} & for\ 15 < \mathrm{Re} < 250 \end{matrix} \qquad (5)$$

Sherwood number is the ratio of external convective mass transfer to the molecular diffusion, while Biot number is the ratio of external mass transfer resistance to the overall mass transfer resistance.

The evaluated effective diffusivities are reported in Table 4 along with the RMSE values. The effective diffusivity is found to increase with increase in temperature of the heating medium, and with increase in the flow rate of the heating medium. The increase or decrease in effective diffusivity coefficient is in accordance with the magnitude of drying kinetics. An increase in the solids holdup is found to decrease the effective diffusivity coefficient. The increase in drying rate with flow rate and decreased solids holdup has been attributed to the increase in effective bed temperature. The effective diffusivity is found to vary within $1.69*10^{-11}$ to $3.26*10^{-11}$ m^2/s with RMSE less than 4%. Although the errors are higher, these kinetic parameters are very essential in the design andscale up of drying process with certain degree of confidence. Uckan and Ulku

[25] have reported an effective diffusivity of 2.1 to $3.9*10^{-11}$ m^2/s for drying of corn in fluidized bed. Senadeera et al (31) have reported effective diffusivities to be within 10^{-8} to 10^{-12} for drying of potatoes and peas in fluidized bed, in accordance with the diffusivity coefficients reported by Zogzas et al (32). The estimated effective diffusion coefficient in the present study is in the same orderof magnitude as reported in literature.

Table 4: Effective Diffusivity Coefficient at Various Operating Conditions

T, ^{0}C	W, kg	U, m/s	C_e, kg/kg	Deff X10^{11} m^2/s	RMSE
40	0.1	1.32	0.044	2.67	1.88
40	0.125	1.32	0.044	2.05	4.00
50	0.1	1.32	0.035	3.26	1.81
50	0.125	1.32	0.035	2.54	2.18
60	0.1	1.32	0.025	3.38	2.80
60	0.125	1.32	0.025	2.67	2.95
60	0.1	1.05	0.025	1.69	3.10

Although the simple models could closely match with the experimental data, much better than the complex model, they are more empirical in nature and lacks scientific background, restricting their applicability only within the experimental range covered in the present study. However the complex models although have higher error; they can be extended even beyond the experimental range, with certain degree of confidence.

IV CONCLUSION

The drying characteristics of mustard, one of the popular oil seeds in India have been assessed in a fluidized bed dryer with respect to the various operatingvariables. The drying rate was found to increase significantly with increase in temperature and flow rate of the heating medium, while decrease with increase in solids holdup. The duration of constant rate period was found to be insignificant, considering the total duration of drying. The kinetics of drying was tested with various simple exponential decay models and the Page model was found to match the experimental drying rate closely with the RMSE value less than 2.0%. The experimental data were also modeled using fundamental Fick's diffusion equationand the effective diffusivity coefficient was estimated to be within $1.69*10^{-11}$ to $3.26*10^{-11} m^2/s$ for the entire range of experimental data with RMSE less than 4%. The estimated effective diffusion coefficient is compared with the literature reported effective diffusion coefficient for other grains and found to be within the same order of magnitude.

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