# Effect of Distributor-Orifice on Drying Kinetics in a Fluidized Bed Drier (EDODKFBD)

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***Abstract*—The drying characteristics of tea-particles have been analyzed in a fluidized bed drier with different orifice sizes while keeping the number of orifices constant. The experiment has been conducted in a fluidized bed drier having 5 cm inner diameter and 80 cm height with four number of distributors of different orifice sizes. In each experimental run the tea-particles have initial moisture content of 13.38 % on dry weight basis. The initial static bed height and air flow rate are kept constant. Hot air is used as the heating medium, air temperatures are varied for 40, 50 and 60 degree centigrade in the present study. The weight of the tea-particles is measured in 10 minute time interval for total drying time of 130 minutes. The amounts of moisture removed are calculated each time. The drying constants are calculated for different distributors for different temperatures. The experimental results show that the drying rate increases with the orifice size up to a specific limit, after which it starts to fall thereby indicating the effect of other pre-dominant parameters.**

***Index Terms*—Fluidized bed drier, drying kinetics, drying constant and distributor orifice.**

## I. Introduction

Batch drying of granular particles using fluidized beds is an established fluid solid contacting technique. Fluidized beds provide more efficient air solid contact and hence, faster drying than any other methods because of homogeneous mixing and uniformity can be achieved by fluidization [1]. The product chamber of a fluidized bed dryer can be of any convenient shape viz. cylindrical and non cylindrical [2]. The fluidization operation can be achieved by increasing the upward velocity of air [3]. The solid material is suspended at superficial gas velocities well in excess of the minimum fluidization velocity so that gas – solid contact is maintained in the unit.

The fluidization technology is commonly used in drying agro-food materials & other materials. It is also used in freezing systems. Fluid bed drying has been recognized as a gentle, uniform drying method capable of drying down to very low moisture content with a high degree of efficiency [4]. This process is characterized by high moisture and heat transfer rates and excellent thermal control capacity compared to conventional drying process [5]. It is also a very convenient method for drying heat sensitive food materials as it prevents them from overheating due to mixing [6]. The mass and heat transfer effects are determined by the bubble-characteristics of the fluidizing medium, which mainly depends on design of distributor [7].

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## II. Literature

The principle behind the drying process is to blow a hot gas upward through a bed of particulate material to be dried. The gas is distributed evenly through small orifices supporting the material at a rate sufficiently high to cause incipient fluidization of particles but not so high as to give the appearance of a vigorously bubbling bed. The fluidized bed drier can be operated continuous or batch wise. The drying kinetics is important for the estimation of the performance of the drier. A complex transport phenomenon takes place during drying process including unsteady state heat and mass transfer which occur simultaneously [8]. Heat necessary for evaporation is supplied to the particles and moisture vapor is removed from the material into the drying medium. Heat is transported by convection from the surrounding to the particle surfaces and then by conduction from there into the inside of particles. Moisture is transported in the opposite direction as a liquid or vapor on the surface evaporates and passes on to the surroundings by convection [9]. The heat and moisture transfer rates are related to drying air temperature and Reynolds number as a function of velocity of circulating air. Thus mass and energy balance mechanisms are involved in drying operations.

The developments of the regime of fluidization and subsequent design modifications have made fluidized bed drying a desirable choice among other driers. However the efficiency of the conventional drying system is usually low. It is therefore desirable to improve the efficiency of the drying process. Drying process has usually three typical drying rate periods [10] namely pre-warming period, constant rate and falling rate period. The different theories have been used to describe moisture evaporation in the different periods. The diffusion controls the drying process especially in the falling rate period [11].

Sender et al. [12] reviews common air distributors used in fluidized bed drying of food particulates and also reviews special methods of fluidizing larger irregular food particulates. Depending on the size and shape of the materials, methods employed to achieve effective fluidization during fluid bed drying varies from use of simple hole distributors for small, light weight materials to special techniques for larger and/or moist materials [12].

Tanfara et al. [13] studied the effect of particle size distribution on local voidage in a bench-scale conical fluidized bed dryer containing dried placebo pharmaceutical granule. They found the local voidage to be sensitive to small changes in static bed heights. They also observed that as the static bed height increases the gas spreads more uniformly over the bed cross-section for a wide particle size distribution at constant velocity (up to 0.5m/s). The gas flow (at velocity above 0.5m/s) was observed to be more centralized with increasing bed heights. Inci and Dursun, [14] modeled the drying kinetics of single apricot by varying the drying rate with time and moisture by different methods. They also compared different models where Logarithmic model as mentioned below was found to be the best to explain the drying behavior of apricots.

$$t = \frac{4L^2}{\pi^2 D} \ln \frac{8(M_o - M_e)}{\pi^2 (M - M_e)} \tag{1}$$

Senadeera and Desbiolles, [15] undertook some experiments to study the relationship between time and shrinkage constant of food products during fluidized bed drying. Time based volumetric shrinkage coefficient was modeled and found to be varying with drying temperature and sample proportion.

The objective of the present study is to study the effect of the distributor plates with respect to the orifice size on the drying kinetics of tea-particles using fluidized bed technique.

### A. Theoretical Analysis

There are two ways of specifying the moisture content [16]. Moisture content in any sample is calculated as the percentage difference between the wet weight of the sample and the dry weight of the sample. If the percentage difference is calculated relative to the wet weight of the sample, it is called the wet basis moisture content and if it is calculated relative to the dry weight of the sample, it is called the dry basis moisture content. Thus wet-weight basis ($m_w$) and dry-weight basis ($m_d$) are calculated as the following.

$$m_w = \frac{w - d}{w} \tag{2}$$

And

$$m_d = \frac{w - d}{d} \tag{3}$$

where, $w$ = total weight of product, kg & $d$ = weight of dry product, kg

Moisture content on dry basis has been used for all calculation purpose in this paper.

### B. Drying Kinetics

Drying occurs in three different periods, or phases, which can be clearly defined. The first phase, or initial period, is where sensible heat is transferred to the product and the contained moisture. This is the heating up of the product from the inlet condition to the process condition, which enables the subsequent processes to take place. The rate of evaporation increases dramatically during this period with mostly free moisture being removed.

During the second phase, or constant rate period, free moisture persists on the surfaces and the rate of evaporation alters very little as the moisture content reduces. During this period, drying rates are high and higher inlet air temperatures than in subsequent drying stages can be used without detrimental effect to the product. There is a gradual and relatively small increase in the product temperature during this period. Interestingly, a common occurrence is that the time scale of the constant rate period may determine and affect the rate of drying in the next phase.

The third phase, or falling rate period, is the phase during which migration of moisture from the inner interstices of each particle to the outer surface becomes the limiting factor that reduces the drying rate.

Drying occurred only in the falling rate period for all materials during this investigation. The variation of average moisture content of the sample as non-dimensional moisture ratio with time of drying [17] operation is expressed as

$$MR = \frac{M - M_e}{M_o - M_e} = \exp(-kt) \tag{4}$$

Higher the drying constant means lower the amount of moisture is removed from the sample which indicates poor drying operation. The Wang and Singh model as shown in equation (5) is a second order polynomial model which have earlier been used to characterize the drying kinetics of mushrooms [18] and Rough rice [19]

$$MR = \frac{M - M_e}{M_o - M_e} = 1 + at + bt^2 \tag{5}$$

where a and b are coefficients of the model.

## III. Experimentations

A schematic diagram of the experimental set up is shown in the *Figure. - 1*. The overall experimental set up consists of (1) Compressor; (2), (3), (4), (5) Flow controlling Valves; (6) Orifice Meter; (7) Rotameter; (8) Air Heater; (9) Temperature controller with digital temperature meter; (10) Fluidized bed Drier; (11) Distributor plate; (12) U- tube Manometer; (13) Calming Section containing spherical balls & Silica gel; (14) Thermometer for outlet air temperature; and (15) Fluidizing Material.

### A. Fluidized Bed Dryer

The experimental set-up for fluidized bed drying is shown in Fig. -1 consists of a cylindrical column of 5 cm diameter and 80 cm high, made up of Perspex material. The bottom part is of conical shape, named as calming section, contains spherical steel balls of 3mm size and silica gel for uniform distribution of heating medium through the bed. Air distributor plate is fitted at the bottom.

### B. Distributor Plates

Four numbers of circular perforated plates have been used. 66 nos. of holes (orifices) arranged in equilateral triangular pitch are incorporated in each distributor plate. Diameter of orifices is varied as 1.5mm, 2mm, 3mm and 3.5 mm respectively. The plates are made of copper metal. The thickness of plate is 2mm. The open areas for the plates with respect to the total area are 5.94%, 10.56%, 23.76% and 32.34% respectively. A mild steel wire mesh is placed over the distributor to prevent the materials falling into the calming section. Table -1 shows the numbers and size of orifices for different distributors.

TABLE 1 CHARACTERISTICS OF DISTRIBUTORS.

| Distributor | Number of orifices | Diameter of orifices ($d_o$), mm |
|---|---|---|
| D1 | 66 | 1.5 |
| D2 | 66 | 2 |
| D3 | 66 | 3 |
| D4 | 66 | 3.5 |

### C. *Fluidizing Material and Experimental Methods*

The experiment uses tea-particles as bed material for drying with initial moisture content of 13.38% on dry basis**.** The tea material is loaded with initial static bed height of 5cm. A K.G. type stationary water- cooled air compressor, driven by 3-phase induction motor was used. Air from the compressor was metered using a calibrated orifice meter and rotameter, before being heated and fed to the fluidizing column. Air flow rate was maintained at 75 liters/min ($U_0$=38.2m/s) from the ambient temperature i.e. at free air conditions. The scope of the experiment is shown in Table - 2. An electrical heater consisting of multiple heating elements each of 500W rating was used. A temperature controller was used in the air chamber to control air temperature within ± 2°C of the set temperature.

TABLE 2 SCOPE OF THE EXPERIMENT

| Name of the Material | Tea |
|---|---|
| Shape of the material | Irregular |
| Size , $d_p$ $\times 10^{-6}$ m | 1282.5 |
| Particle density, gm/cc | 0.9091 |
| Temperature of fluidizing air ,°C | 40,50,60 |
| Fluidizing air flow rate , Liter/Min. | 75 |
| Static bed height , cm | 5 |
| Initial Moisture content ,% | 13.38 |
| color | Brown |
| $U_{mf}$, m/s | 19.782 |
| Uo, m/s | 38.1972 |

Air at desired temperature and at a constant flow rate was allowed to pass through the fluidizing chamber. A known quantity of tea-particle with known initial moisture content was fed to the column after ensuring steady state temperature and flow rate. Tea samples, weighing around 1.0 gram were scooped out of the bed at regular intervals of time (i.e. at 10 min. interval) for estimation of moisture. The moisture content of tea was determined by drying the samples till a 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.

## IV. RESULTS AND DISCUSSION

The dry bulb temperature was measured as 85$^0$F and the wet bulb depression was found out to be 8$^0$F. Thus the equivalent moisture content of the sample and the relative humidity were found out to be 12.5% and 70% [20] respectively. Experimental data collected are plotted as relative moisture content $C/C_i$ vs. time in Fig. - 2. The drying kinetics is obtained in polynomial forms from the plots of moisture losses against drying time for which the coefficients are tabulated in Table – 3. The obtained second order polynomial model is in the following form.

$$ML = M_o - M = at^2 + bt + c \tag{6}$$

where a, b and c are the coefficients of the model. It is observed that the coefficient *'a'* is almost same for all the systems. This implies that the first coefficient of the above model is independent of the size of orifices in the distributor as well as of temperature of drying. The second coefficient, *'b'* increases slightly with the increase in size of orifices of the distributor up to certain limit and then decreases slightly. This also increases with the temperature slightly for first distributor whereas for the other distributors the coefficient, *'b'* increases with temperature up to 50$^0$C after which it does not change further. These slight changes in the values of the coefficient, *'b'* with temperature and/or orifice sizes can be neglected and may be considered as independent of these parameters. It is also observed that the third coefficient, *'c'* changes with both of these parameters. It increases with the increase in temperature and with the distributor orifices up to certain numbers. It is seen that the coefficient, *'c'* increases up to the D3 and decreases for the fourth distributor, D4 as tabulated on Table-3.

TABLE 3 COEFFICIENTS OF THE DRYING MODEL (EQ.-6) OBTAINED FROM MOISTURE LOSS AGAINST THE DRYING TIME FOR DIFFERENT DISTRIBUTORS AT DIFFERENT TEMPERATURES.

| | Coefficient | 40$^0$C | 50$^0$C | 60$^0$C |
|---|---|---|---|---|
| Distributor-1 | a | -1E-05 | -1E-05 | -1E-05 |
| | b | 0.0019 | 0.0021 | 0.0023 |
| | c | 0.007 | 0.0112 | 0.0148 |
| | $R^2$ | 0.9431 | 0.9264 | 0.9271 |
| Distributor-2 | a | -9E-06 | -1E-05 | -1E-05 |
| | b | 0.0018 | 0.0021 | 0.0021 |
| | c | 0.0113 | 0.0159 | 0.0241 |
| | $R^2$ | 0.949 | 0.9134 | 0.8845 |
| Distributor-3 | a | -1E-05 | -1E-05 | -1E-05 |
| | b | 0.002 | 0.0022 | 0.0022 |
| | c | 0.0142 | 0.0253 | 0.0324 |
| | $R^2$ | 0.9351 | 0.8721 | 0.8149 |
| Distributor-4 | a | -9E-06 | -1E-05 | -1E-05 |
| | b | 0.0018 | 0.0021 | 0.0021 |
| | c | 0.0118 | 0.0165 | 0.0251 |
| | $R^2$ | 0.9407 | 0.9253 | 0.8825 |

TABLE 4 DRYING KINETICS IN THE FORM OF SECOND ORDER POLYNOMIALS FOR DIFFERENT DISTRIBUTORS AT DIFFERENT TEMPERATURES

| TEMP, °C | DR1 | DR2 | DR3 | DR4 |
|---|---|---|---|---|
| 40 | $MR = 0.0012x^2 - 0.2212x + 0.2058$<br>$R^2 = 0.9431$ | $MR = 0.0011t^2 - 0.2091t - 0.2804$<br>$R^2 = 0.949$ | $MR= 0.0012t^2 - 0.2249t - 0.6153$<br>$R^2 = 0.9351$ | $MR=0.0011t^2 - 0.2061t - 0.3395$<br>$R^2 = 0.9407$ |
| 50 | $MR = 0.0013x^2 - 0.2353x - 0.2774$<br>$R^2 = 0.9264$ | $MR = 0.0013x^2 - 0.2346x - 0.8053$<br>$R^2 = 0.9134$ | $MR = 0.0014x^2 - 0.2496x - 1.8782$<br>$R^2 = 0.8721$ | $MR = 0.0013t^2 - 0.2384t - 0.8792$<br>$R^2 = 0.9253$ |
| 60 | $MR = 0.0014t^2 - 0.2581t - 0.6873$<br>$R^2 = 0.9271$ | $MR = 0.0013t^2 - 0.2397t - 1.7371$<br>$R^2 = 0.8845$ | $MR = 0.0014t^2 - 0.2489t - 2.6786$<br>$R^2 = 0.8149$ | $MR = 0.0013t^2 - 0.244t - 1.8525$<br>$R^2 = 0.8825$ |

The drying kinetics of tea has also been studied by plotting the moisture content of the sample as non-dimensional moisture ratio (MR) against the time of drying. The model is also a second order polynomial of the form as discussed in eq-(6). The polynomials for different systems are listed in Table-4. It is seen that the coefficients *'b'* and *'c'* are negative in all cases except c being positive for the distributor-1 at 40$^0$C. On the other hand the coefficient '*a'* is positive for all cases. But almost constant value of this coefficient as seen in Table-4 indicates that other two coefficients (i.e. b and c) affect the drying to some extent. Again these polynomials imply that higher the values of these coefficients lower the drying rate. Thus drying kinetics of tea can be analyzed from these polynomials.

The drying kinetics obtained in exponential forms (Eq-4) in terms of moisture loss and drying time are tabulated in Table–5. The drying constants thus obtained for different distributors openings at different temperatures are tabulated in Table-6.

TABLE 5 DATA ON DRYING KINETICS OBSERVED FROM EXPERIMENTS

| Temp, $^0$C | D1 | D2 | D3 | D4 |
|---|---|---|---|---|
| 40 | $ML=0.0418*e^{0.0079t}$ | $ML=0.046*e^{0.0076t}$ | $ML=0.0534*e^{0.0064t}$ | $ML=0.0464*e^{0.0071t}$ |
| 50 | $ML=0.0496*e^{0.0072t}$ | $ML=0.057*e^{0.006t}$ | $ML=0.074*e^{0.004t}$ | $ML=0.059*e^{0.006t}$ |
| 60 | $ML=0.0591*e^{0.006t}$ | $ML=0.0709*e^{0.0047t}$ | $ML=0.0855*e^{0.0033t}$ | $ML=0.0733*e^{0.0045t}$ |

TABLE 6 DRYING CONSTANTS OBTAINED FOR DIFFERENT DISTRIBUTOR ORIFICES AND TEMPERATURES

| Temp, $^0$C | D1 | D2 | D3 | D4 |
|---|---|---|---|---|
| 40 | 0.0079 | 0.0076 | 0.0064 | 0.0071 |
| 50 | 0.0072 | 0.006 | 0.0043 | 0.006 |
| 60 | 0.0063 | 0.0047 | 0.0033 | 0.0045 |

The Figures – 2 and 3 as well as Tables 3,4, 5 and 6 show that the drying rate is increasing from D1 to D3 at different temperatures and decreasing for the distributor, D4. The drying rate for the fourth distributor (D4) is falling in between D3 and D2. Again it is also observed that the drying rate increases with the temperatures irrespective of the size of the orifices of the distributors. Thus it is clear from these observations that the size of the orifices of the distributor plays an important role in the fluidized bed drying operation and moisture removal rate increases with the orifice size up to a certain extent.

The above phenomenon of decrease in drying rate with increase in orifice sizes after certain limit may be attributed to the followings.

1) With all the other operating conditions remaining same the total pressure drop for the fluidization process is constant, ($\Delta p_t = \Delta p_b + \Delta p_d$).

For perforated plate distributor $\Delta p_d$ must be large when operating close to $u_{mf}$, but can be lower when the bed operates at high $u_o$ (last paragraph of page -103, Kunii and Levenspiel (1977). Thus as the distributor opening increases the distributor pressure drop decreases at high $u_o$ thereby increasing the bed pressure drop. Increased bed pressure drop decreases the drying kinetics. Up to D3 the decrease in distributor pressure drops might not be prominent as a result the drying kinetics increase for increased amount of heating medium only.

2) For stable operation of fluidized bed $\frac{\Delta P_d}{\Delta P_b} = 0.15$ if

$$\frac{U_o}{U_{mf}} = 1 - 2 \text{ (p-106 of [3]).}$$

In the present case it is calculated that the minimum fluidization velocity, $U_{mf}$ is 19.782 m/s and the superficial velocity is 38.2 m/s. Thus Uo is almost twice the minimum fluidization velocity. Therefore the above relation holds good and the distributor pressure drop will about 15% of the bed pressure drop.

With the increase of orifice size in the distributor the distributor pressure drop decreases. Therefore the bed pressure drop must have increased to maintain the constant total pressure drop for the system. Thus the value of the ratio $\frac{\Delta P_d}{\Delta P_b}$ decreases further. But the $\frac{U_o}{U_{mf}}$ ratio remains same which may cause some instability in the bed behavior thereby affecting the drying operation. Thus non-uniformity in drying might have resulted and the drying kinetics might have affected. This non-uniformity in drying effect may be negligible up to certain orifice sizes. In the present study it may be up to D3.

## V. CONCLUSION

The drying characteristics of materials (tea granules) have been assessed in a fluidized bed dryer with respect to the various operating variables and distributor plates. The drying rate is found to increase significantly with increase in temperature and flow rate of the heating medium .The drying rate is found to increase with the orifice sizes in the distributor. But it is observed that it increases up to a certain extent and afterwards it decreases. The more drying rate is due to the fact that the larger the hole- diameter, the better is the contact of heating medium and the solids in the bed. But when the hole- diameter is more than the specific size, the drying rate reduces because of the decreased pressure drop at the distributor and thereby the decrease in the ratio of distributor pressure drop to bed pressure drops which in turn results in non-uniform drying operation. Although there is good contact of heating medium and the solids in the fluidized bed dryer, the flow rate and distributor pressure drop factors affect the proper contact of material and the heating medium predominantly.

Change in drying constant values for the bed materials (tea granules) with the temperatures and different orifice sizes of the distributor plates is in a very narrow margin with drying kinetics. Thus the drying efficiency and the drying rates are affected. As the relationship between bed pressure drop and distributor pressure drop is a well known critical design parameter, further work can be carried out to find out the optimum ratio of this parameter at which the drying rate is peaked which would be very useful to practical designers. Simple mathematical models can also be derived for different samples of bed materials showing the relationship with drying constant and other system variables. These models could be further improved by incorporating structural changes during drying.

## NOMENCLATURE

$m_w$ : Mass in wet-weight basis, kg
$m_d$ : Mass in dry-weight basis , kg
$w$ : Total weight , kg
$d$ : Dry weight, kg
D: Apparent Diffusion coefficient, $m^2/s$
k:The drying constant
L : Half slice thickness, mm
M : Moisture Content, % wet basis
$M_o$: Initial Moisture Content, % wet basis
$M_e$: Equilibrium Moisture Content, % wet basis
ML: Moisture loss from sample (g water/g dry solids)
MR: Moisture ratio, non-dimensional
T : drying time, min
$D_1$, D2, D3, D4 : distributor plates, 1st, 2nd, 3rd and 4th.
$D_P$ : PARTICLE SIZE, IN μM.
C : moisture content (dry basis), at time t, minute.
Ci: initial moisture content (dry basis)
$\Delta p_b$: Bed pressure drop, Pa
$\Delta p_d$: Distributor pressure drop, Pa
$u_{mf}$ : Minimum fluidization velocity, m/s
$u_o$ : Superficial velocity of fluid, m/s

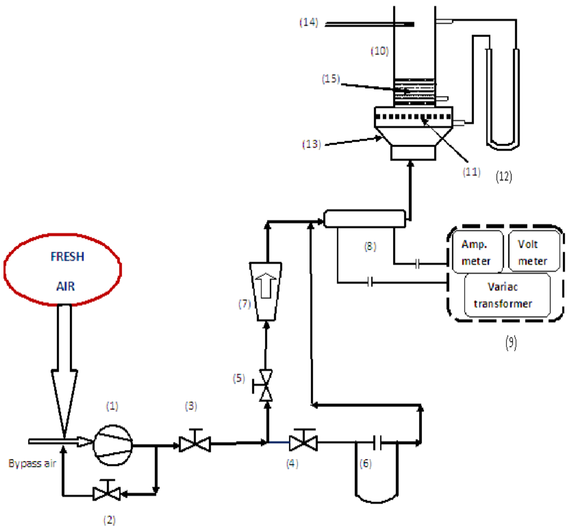


Fig.1. Schematic Diagram of the experimental set-up

Figure.2 [C/Ci Ratios Of Different Distributors]

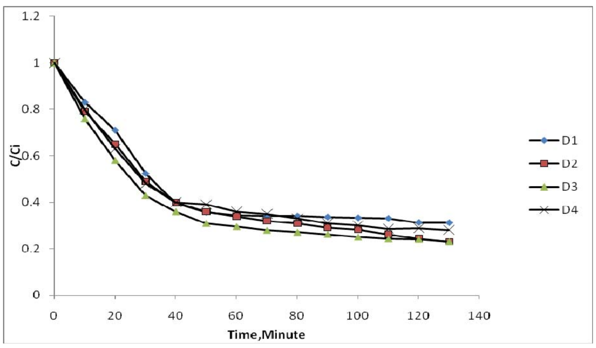


Fig.2 (i) Comparison of drying kinetics at 40°C

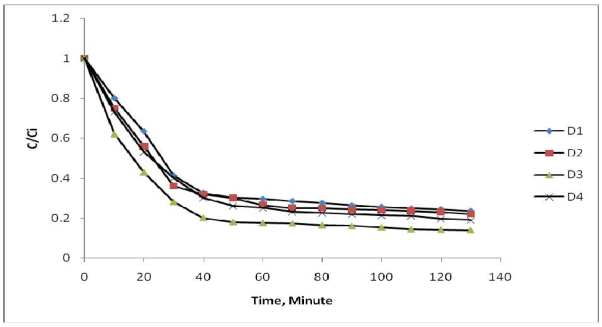


Fig. 2 (ii) Comparison of drying kinetics at 50°C

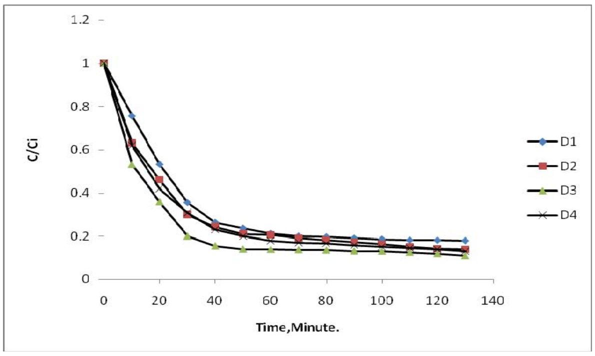


Fig. 2 (iii) Comparison of drying kinetics at 60°C

Figure-3 [comparison of mr values for drying of tea at different temperatures].

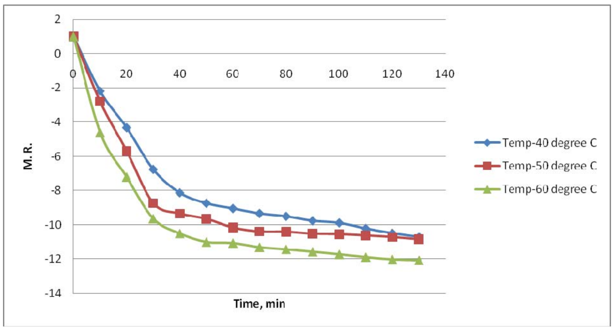


Fig.3 (i) Drying Kinetics for Distributor-1

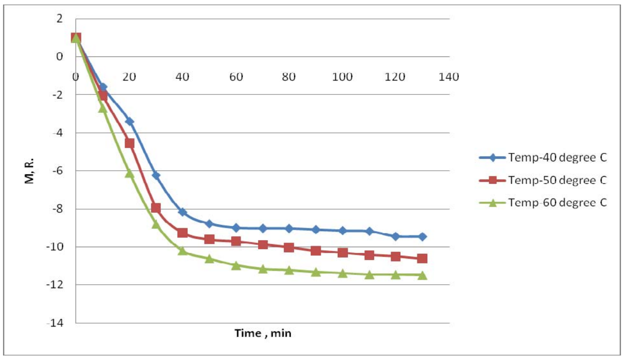


Fig.3 (ii) For Distributor-2

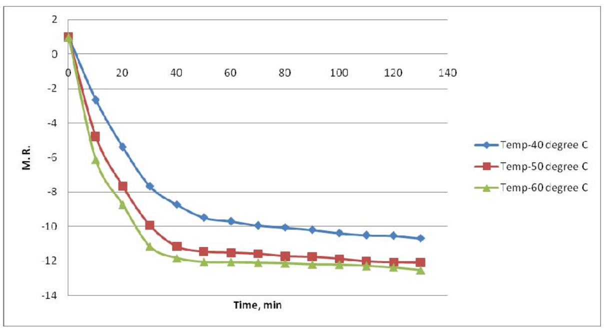


Fig.3 (iii) For Distributor-3

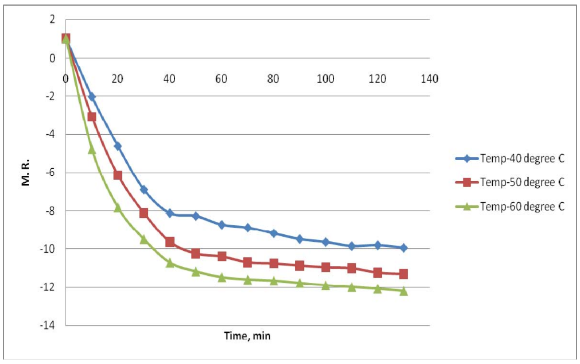

Fig.3 (IV) For Distributor-4

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