CHAPTER 6

NUMERICAL STUDIES

6.1 INTRODUCTION

The characterization of the drying behavior using a strict experimental approach constitutes a formidable challenge due to the excessively large number of variables involved. It is therefore necessary to develop a procedure to simulate the product drying behavior, which allows one to extend the results of the experimental drying investigations. Mathematical modeling has now become a practice in analyzing the drying phenomena (Kaya et al., 2006) due to the cost and time involved in experimental studies. In this way, the impact of many variables on drying behavior can be examined and interpreted without having to resort to an extensive procedure of experimental testing (Can 2007).

The models are established with the governing differential equations coupled with the initial and boundary conditions. As the governing equations are generally nonlinear in structure, their analytical solution is mostly complicated; hence, numerical solutions are sought (Chua et al., 2002; Dietl et al., 1998). In the majority of the modeling works, the heat and mass transfer coefficient values at the interface were assumed to be constant throughout the surface of the object (Karim and Hawlader 2005; Balaban and Pigott 1988). But in reality, these coefficients vary along the surface of the product. Hence, during drying, heat and mass transport have to be studied along with the flow field as a conjugate problem (Olivera and Haghighi...
In this work, a two-dimensional numerical model for bagasse drying considering air and steam as the drying medium are analyzed separately, considering variable heat and mass transfer coefficients. The model description and solution procedure for air and steam drying studies are separately presented.

6.2 AIR DRYING STUDIES

A schematic diagram of the bagasse drying process showing both heat and mass transport in all the phases is shown in Figure 6.1. Here, the product is considered to be placed in a drying chamber and hot air is passed over the product. Here the main transport mechanism has been identified and numerically expressed in the form of equations. The two-dimensional model is developed taking into account the transport mechanisms such as, conductive heat transfer from the heating plate to the product’s bottom surface, conductive heat transfer within the product mass, convective heat transfer between air and the product’s surface, moisture diffusion in the solid towards its external surface and convective mass transfer of the vapor into the air stream.

![Figure 6.1 Schematic of the physical model](image-url)
The flow fields are numerically modeled using a Fluent CFD package to analyze the heat and mass transfer coefficients at different time intervals. For the solid zone a numerical procedure is developed using MATLAB codes to analyze the heat and mass transfer through diffusion inside the product; thereby temperature and moisture profile at different time intervals are studied.

The assumptions made to analyze the transport mechanism through diffusion inside the product are as given below:

a. The drying product is compact and homogeneous with uniform initial moisture content and uniform initial temperature.

b. The thermo-physical properties of the material are temperature and moisture content- dependency.

c. Negligible shrinkage and deformation of the material during drying and the mode of moisture transfer is only by liquid diffusion.

d. Negligible heat generation inside the moist object and negligible radiation effects.

### 6.2.1 Governing Equations for the Flow Field

The equations governing the drying stream in two-dimensional geometry are the mass, momentum, energy and species transport. They are as follows:
Mass conservation equation:

\[
\frac{\partial \rho}{\partial t} + \rho \frac{\partial u}{\partial x} + \rho v \frac{\partial v}{\partial y} = 0 \tag{6.1}
\]

Momentum equations:

\[
\rho \left( \frac{\partial u}{\partial t} + u \frac{\partial u}{\partial x} + v \frac{\partial u}{\partial y} \right) = -\frac{\partial p}{\partial x} + \mu \left( \frac{\partial^2 u}{\partial x^2} + \frac{\partial^2 u}{\partial y^2} \right) \tag{6.2}
\]

\[
\rho \left( \frac{\partial v}{\partial t} + u \frac{\partial v}{\partial x} + v \frac{\partial v}{\partial y} \right) = -\frac{\partial p}{\partial y} + \mu \left( \frac{\partial^2 v}{\partial x^2} + \frac{\partial^2 v}{\partial y^2} \right) \tag{6.3}
\]

Energy equation:

\[
\frac{\partial T}{\partial t} + u \frac{\partial T}{\partial x} + v \frac{\partial T}{\partial y} = \alpha \left( \frac{\partial^2 T}{\partial x^2} + \frac{\partial^2 T}{\partial y^2} \right) \tag{6.4}
\]

Species transport equation:

\[
\frac{\partial}{\partial t} (\rho Y_{H_2,O}) + u \frac{\partial}{\partial x} (\rho Y_{H_2,O}) + v \frac{\partial}{\partial y} (\rho Y_{H_2,O}) = \rho \left( \frac{\partial^2 Y_{H_2,O}}{\partial x^2} + \frac{\partial^2 Y_{H_2,O}}{\partial y^2} \right) \tag{6.5}
\]

The Fluent CFD package based on the finite volume method is used to transform and solve these equations for the required time steps. The discretization scheme used is a second order upwind scheme for the convective terms in the momentum and energy equations, and the SIMPLE algorithm for pressure-velocity coupling. The mesh is generated in the gambit preprocessor to produce a grid-independent solution.

From the flow field for every time step, the local convective heat transfer coefficient can be determined using the temperature field obtained.
from the flow regime. After the convective heat transfer coefficient \( h_c \) is determined, using the analogy between the thermal and concentration boundary layers (Kaya et al., 2006), the convective mass transfer coefficient can be calculated using,

\[
h_m = h_c \left( \frac{D L e^n}{k} \right)
\]  

(6.6)

Where;

\( h_c \) is the convective heat transfer coefficient determined using the temperature obtained from the flow regime,

\( D \), diffusivity value is obtained using the equation (4.18).

\( L e \), Lewis number is obtained as ratio of thermal to moisture diffusivity (\( \frac{\alpha}{D} \)).

The constant \( n \) is assumed as (1/3) (Kaya et al., 2006),

\( k \) is the thermal conductivity of bagasse (0.45W/mK).

### 6.2.2 Governing Equations for Solids

In this section, a numerical procedure is developed to analyze the heat and mass transfer through diffusion inside the object being dried with some assumptions as mentioned above. The governing equations and the boundary conditions for the solid zone are given as;

\[
\frac{1}{\alpha} \frac{\partial T}{\partial t} = \frac{\partial^2 T}{\partial x^2} + \frac{\partial^2 T}{\partial y^2},
\]  

(6.7)

\[
\frac{1}{D_c} \frac{\partial M}{\partial t} = \frac{\partial^2 M}{\partial x^2} + \frac{\partial^2 M}{\partial y^2},
\]  

(6.8)

with the following initial and boundary conditions:

\[ T(x, y, 0) = T_0, \quad M(x, y, 0) = M_0 \text{ at } t = 0 \]

at time \( t > 0 \),
\begin{align*}
y = W \quad \text{and} \quad 0 \leq x \leq L \quad & -k \frac{\partial T}{\partial y} = h_c (T - T_{air}), \\
x = 0 \quad \text{and} \quad 0 \leq y \leq W \quad & \frac{\partial M}{\partial x} = 0, \\
x = L \quad \text{and} \quad 0 \leq y \leq W \quad & \frac{\partial M}{\partial x} = 0, \\
y = 0 \quad \text{and} \quad 0 \leq x \leq L \quad & \frac{\partial M}{\partial y} = 0, \\
y = 0 \quad \text{and} \quad 0 \leq x \leq L \quad & -k \frac{\partial T}{\partial y} = q'', \\
y = W \quad \text{and} \quad 0 \leq x \leq L \quad & -D_e \frac{\partial M}{\partial y} = h_m (M - M_{air})
\end{align*}

where \(D_e\) is the moisture diffusivity and is obtained from the Arrhenius equation for bagasse drying.

\[D_e = D_0 \exp \left( \frac{-E_u}{RT} \right) \quad (6.9)\]

### 6.2.3 Numerical Solution Procedure

The governing equations of the flow and energy are solved for two-dimensional geometry using the commercial CFD package Fluent V.6.1.18, which uses a segregated finite volume based approach. A second order upwind scheme is used for discretization, with SIMPLE algorithm for pressure velocity coupling. The global residues for each time step are below 1E-04 and care was taken to obtain a grid independent solution.

For each time step, the heat transfer coefficient \(h_c\), is obtained from Fluent, which is used to calculate the mass transfer coefficient \(h_m\), using heat and mass transfer analogy as given by Equation (6.6). This is then fed to the code which solves the transient heat and moisture transfer equations as explained by Equations (6.7 and 6.8). These equations are solved using the
Figure 6.2 Flow Chart Illustrating the numerical solution Procedure
finite volume method with a tri-diagonal matrix algorithm (Patankar 1980) to solve the resulting algebraic equations. In the solution a conservative first order implicit backward Euler integration approach is used for a transient term. The complete code is written and compiled in a commercial package MATLAB 7. The resultant average temperature and moisture at the surfaces of the solid is updated in Fluent for the next time step and the procedure is continued until the required drying conditions are attained. The flow chart illustrating the solution procedure is shown in Figure 6.2.

The purpose of grid independence test is to determine the minimum grid resolution required to generate a solution that is independent of the grid used. Here for the part of modeling where the commercial software (FLUENT) used; it is checked for a grid independent results. The grid independence test was performed on the fluid flow geometry and for the heat transfer coefficient which is the parameter of interest. Starting with a coarse grid the number of cells was increased in the region of interest until the solution from each grid was unchanged for successive grid refinements. For the three grid refinement performed it is observed that the parameter of interest, the heat transfer coefficient was varying with in 5 percent.

For the solid region where the elliptic equations were solved using Matlab, the code was tested for a 2-dimensional heat conduction analysis for which analytical solution is available (Incropera 2001).

6.2.4 Simulation Results

The drying process is simulated under real operating conditions based on experimental thin layer drying kinetics. On the basis of a detailed experimental investigation on drying of bagasse, the validation of the mathematical model developed (for the applied parameter range) was carried
out. During the drying process, it is important to know how the temperature and moisture distribution change with time. This information can be used to adjust the drying parameters more effectively. The temperature and moisture profile of the product depends on the drying condition and concentration of the surrounding fluid field. In this analysis the moisture diffusivity is considered as temperature dependent (Equation 6.9). The variable moisture diffusivity values are incorporated into the model to study how the moisture transfer coefficient changes with the surface coordinate for different drying conditions.

In order to compare the model predictions with the experimental results, the average moisture content of the sample was estimated from numerical results for the predicted moisture distribution. The predicted average moisture content of the sample is compared with the experimental values. A detailed comparison of the experimental results with the average predicted moisture content during drying of the product under varied operating conditions is shown in Figures 6.3 and 6.4. The comparison of the predicted and experimental moisture content for varying air temperature from 80 to 120°C with air velocity maintained at 1m/s for a bed thickness of 40 mm is shown in Figure 6.3. It is observed that the rate of moisture reduction increases with drying temperature, since the moisture diffusivity depends on temperature. Figure 6.4 shows the comparison of the moisture content data for varied drying bed thickness (20, 40 and 60 mm) with drying air conditions at 100°C and 1 m/s.
Figure 6.3  Simulation and experimental results of average moisture content during drying of product at bed thickness of 40 mm and air velocity at 1 m/s while the air temperature varied.

Figure 6.4  Simulation and experimental results of average moisture content during drying with conditions of air at 100°C and 1 m/s while the bed thickness is varied.
As seen from this Figures, it is evident that the moisture content decreases as drying progress; this is due to the decrease in the concentration gradient between the drying air and the product. The rate of decrease in the moisture content depends on the product thickness, air temperature and velocity. For both numerical and experimental results the product thickness has a higher influence than that of air temperature. From Figures 6.3 and 6.4, it can be observed that generally there is a close agreement between predicted and experimental moisture values. The capacity of this model to represent the experimental data was realized in terms of the mean relative percentage deviation modulus $E$ defined by:

$$E = \frac{100}{Z} \sum_{i=1}^{Z} \frac{|e_i - p_i|}{e_i}$$  (6.10)

where, $e_i$ and $p_i$ are the experimental and predicted data being compared; and $Z$ is the total number of samples compared. It is seen from the figure, that generally there is a close agreement between predicted moisture values and experimental data. The largest discrepancy between the model and experimental data for the specified drying conditions was 7.64%. The discrepancies between predictions and experimental results could be attributed to the sum of errors inherent to the experiment investigations and to the exothermic reactions possibly taking place within the product. However, the simulations predict well the general trend of the experimental moisture content curves, which validates the mathematical model describing the drying process.

For this conduction coupled convection drying, the measured drying rate was higher than the calculated one in the initial stage of drying. This is because, the model assumes zero conductive heat input before drying commences, whereas in reality the heating plate was initially brought into the required temperature and maintained constantly before the start of drying;
hence in reality, the conduction heat supplied will be higher than the model predictions during the initial stage.

Figures 6.5 and 6.6 shows the temperature and moisture profile predicted by the model at various drying times (30, 300 and 600s) for a bed thickness of 40 mm and drying air conditions at 100°C and 1 m/s. From the temperature profiles obtained for different times it is noticed that the temperature reaches almost a steady state within a few minutes when the drying commenced, which is due to the effect of the combined heat addition to the product by conduction from the bottom and by convection from the top surface as shown in Figure 6.5.

**Figure 6.5** Temperature profile at various drying time for bed thickness of 40 mm
From the moisture profile contours in Figure 6.6, the propagation of the drying front within the product with time can be seen; this is due to the existence of the concentration gradient between the product surface and the drying medium.
Figures 6.7 and 6.8 shows the variation in temperature and moisture contours for varied bed thickness at the same drying conditions for time $t = 600$ s. The variation in propagation of the drying front for different bed thickness is shown; as expected, the drying front propagation is faster for lower bed thickness.

**Temperature profile for product thickness of 20 mm**

**Temperature profile for product thickness of 40 mm**

**Temperature profile for product thickness of 60 mm**

*Figure 6.7 Temperature profile for varied product thickness at time $t = 600s$*
Moisture profile for product thickness of 20 mm

Moisture profile for product thickness of 40 mm

Moisture profile for product thickness of 60 mm

Figure 6.8  Moisture profile for varied product thickness at time $t = 600s$

Figures 6.9 and 6.10 show the temperature and moisture profile predicted by the model for varied air temperature conditions of 80, 100, and 120°C respectively at time $t=600s$ for a bed thickness of 40 mm. From the
temperature and moisture profiles it is evident that the operating temperatures have a considerable influence on the drying rate of the product.

Figure 6.9  Temperature profile for varied air temperature with air velocity at 1 m/s
Figure 6.10 Moisture profile for varied air temperature with air velocity at 1m/s

Figures 6.11 and 6.12 shows the temperature and moisture profile predicted by the model for varied air velocity conditions of 0.5, 1.0 and 1.5 m/s, at time t = 600s for bed thickness of 40 mm and air temperature at 100°C. From the temperature and moisture profiles it is evident that the air
velocity has less impact on the drying rate of the product compared to that of air temperatures.

Figure 6.11 Temperature profile for varied air velocity with drying air temperature at 100°C
Figure 6.12 Moisture profile for varied air velocity with drying air temperature at 100°C

The model also predicts the temperature and moisture distribution at any section of the chamber during drying for a given time. The moisture distributions along the Y-axis (x = L/2, 0 ≤ y ≤ W) for varied air temperature
and air velocity for a bed thickness of 40 mm at time $t=600\text{s}$ is shown in Figures 6.13 and 6.14 respectively.

**Figure 6.13** Moisture profile along Y-axis at mid section for varied temperature at product thickness of 40 mm and air velocity at 1 m/s

**Figure 6.14** Moisture profile along Y-axis at mid section for varied air velocity at product thickness of 40 mm and air temperature at 100°C
The moisture in this section was lowest at the top surface where the hot gas and the product are put in contact, which results in the evaporation of liquid moisture from the product surface by absorbing the heat of vaporization. The moisture migrates toward the external surfaces of the product under the influence of the moisture content gradient. Hence the moisture content at the bottom surface of the product will be higher and their corresponding distributions are as shown in Figure 6.14.

6.3 STEAM DRYING STUDIES

The superheated steam drying (SSD) technique has been known for a long time, but its application to industrial drying has received little attention. In the superheated steam drying (SSD) process, superheated steam acts both as the heat source and as the drying medium to take away the evaporated water from the drying product (Pronyk et al., 2004). As this SSD necessarily produces steam equal in amount to the water evaporated from the product, it would be possible to recover all the latent heat supplied during the drying process by condensing the exhaust steam or by mechanical or thermal compression for re-use in the dryer or elsewhere for process applications. Energy recovery in steam drying leads to lower net energy consumption of the dryer compared to conventional air drying (Tatemoto et al., 2001).

The sugar industry is one such industry which utilizes low pressure and temperature steam in large quantities for process applications. Bagasse which is the main fuel for the sugar industry needs to be dried in order to use it efficiently. The sugar industry is a continuous process industry which requires a continuous supply of low pressure steam for process applications (Hugot 1986). The steam drying in the industry could contribute to a substantial reduction in steam generation, which otherwise has to be supplied by a steam generator.
Flue gas drying has been used for the drying of bagasse by many researchers (Kinoshita 1991). However, flue gas drying does not solve the principal problem of utilizing the evaporated moisture in the wet bagasse, and hence, the latent heat of the generated vapor is lost by sending it through the chimney. Steam drying is a measure that allows the recovery of the energy which is spent for evaporating water from the bagasse by making use of the vapor generated during drying for sugar factory process applications. Besides the conventional measures to reduce the steam demand of the sugar process, drying of bagasse can contribute to an increase in the electrical power yield. With the steam drying of bagasse, power production is predicted to increase by about 17% and boiler efficiency is predicted to increase to 81% if bagasse is dried to 10% moisture (Boris Morgenroth and Druce Batstone 2004). They also indicated that by applying steam drying in combination with an optimization of energy in a cane sugar factory, an export of up to 145 kWh/t cane is possible.

Normally, due to a low initial temperature of a material to be dried, during the initial stage of superheated steam drying, condensation temporarily occurs on the material surface, and then the condensed water evaporates into superheated steam, which is a phenomenon peculiar to superheated steam drying referred to as the “reverse process” (Iyota et al., 2001). In this work an attempt is made to mathematically model the Steam-driven Bagasse Dryer, in which wet bagasse is directly contacted with superheated steam. The heat and mass transfer model for the thin layer superheated steam drier was developed by considering the phenomena which occur during the initial stage of drying; i.e., condensation of superheated steam on material surfaces, and the subsequent shift from condensation to evaporation leading to the beginning of the actual drying (Reverse Process Model). The drying equations for the process are formulated, based on the thin layer drying kinetics model to calculate the moisture distributions in the material. The simulation model can
be used to analyze the influence of the drying variables on the product moisture profile and the drying rate in the bed. In addition, the steam and air drying simulation results are compared for the change in the mass of the material with time and the characteristic drying curve.

6.3.1 Model Description

The steam drying models differ from the air drying models by exhibiting three distinct zones during the drying process. When the material temperature is less than the boiling temperature of water \( (T_m < T_{\text{boil}}) \) and when exposed to superheated steam, the superheated steam contacting the material gets cooled and condenses onto the material surface. In the second zone when the temperature of the material rises and equals to \( T_{\text{boil}} \) \( (T_m = T_{\text{boil}}) \), the condensed water on the material surface begins to evaporate which is indicated by the restoration process in Figure 6.15. Thereafter, the actual drying process continues. During this process, if the surface temperature of the material exceeds \( T_{\text{boil}} \), it indicates that the drying is completed in that zone.

Figure 6.15 represents the phenomenon of steam drying showing the different stages involved in steam drying. In the Figure 6.15, \( t_{\text{rev}} \) is the ‘reverse time’ which represents the completion of the initial stage of condensation and shifts to evaporation; and at this time, the mass of the material reaches its maximum. Once past the reverse time \( t_{\text{rev}} \), the evaporation process starts and the condensed water begins to evaporate. This period of the drying is termed as ‘restoration time’ \( t_{\text{res}} \) and this stage extends until the original mass has returned. Moreover, the process from reverse time \( t_{\text{rev}} \) through restoration time \( t_{\text{res}} \) is termed as the ‘restoration process’ and the stage after that restoration process is the actual drying process.
Figure 6.15 Superheated steam drying characteristics curve with reverse process (Iyota et al., 2001)

The heat balance for the model at different stages such as condensation, evaporation on the surface, and evaporation inside of the material is shown in Figure 6.16. Here, $y$ indicates the absolute space coordinate, $Y_s$ is the surface position and $Y_p$ is the position of the phase change interface, where the phase change occurs. The convective heat flux added to the material surface from the drying medium is defined as $q_H$ (W/m$^2$). The water evaporation rate from the material is defined as $J$ (kg/m$^2$s). As condensation progresses on the surface of the material, heat generation in the material is accompanied, this is given by $q_p$. This heat flux given to the material due to condensation (Phase change) is larger than that of $q_H$; hence, during this process the product temperature rises rapidly. The conductive heat flux of the material is given by $q_{cd}$. 
Defining $\gamma$ as the latent heat of evaporation / condensation of water, the evaporation rate $J$, the phase change heat flux $q_p$ can be related as

$$q_p = J\gamma \quad (6.11)$$

In the condensation stage of the model Fig. 6.16(a) the conductive heat flux $q_{cd}$ from the material surface into the interior may be considered as the sum of the convective heat flux $q_H$ and the heat flux due to condensation $q_p$. As the material temperature rises and equal to $T_{\text{boil}}$, no more condensation heat flux is added to the material; hence $q_H = q_{cd}$, which is represented in Figure 6.15 by “$t_{\text{rev}}$” where the material mass reaches its maximum.

Thereafter, as shown in Figure 6.16(c) moisture evaporation from the surface begins making the heat flux $q_p$ as negative. As drying progresses, the surface moisture content reaches its critical moisture content; hence, the phase change interface begins to recede inside the material as shown in Figure 6.16(c). The heat balance at this interface ($Y = Y_p$) can be represented as

$$q_H = (q_{cd} - q_p)_{Y = Y_p} \quad (6.12)$$
6.3.2 Basic Equations

The theoretical analysis in this study is based on the following assumptions. The heat and mass transfer are two dimensional and the shrinkage of the sample is negligible. The calculation is performed from the surface \( y=Y \) to the bottom \( y=0 \) in the sample.

The mass transfer within the material is expressed in the same form as the diffusion equation.

\[
\frac{1}{D_e} \frac{\partial M}{\partial t} = \frac{\partial^2 M}{\partial x^2} + \frac{\partial^2 M}{\partial y^2}
\]  (6.13)

The two-dimensional unsteady energy balance equation in the humid zone is expressed as

\[
\frac{1}{\alpha} \frac{\partial T}{\partial t} = \left( \frac{\partial^2 T}{\partial x^2} + \frac{\partial^2 T}{\partial y^2} \right) \pm J_y
\]  (6.14)

The convective heat transfer from the drying medium to the product surface can be expressed as

\[
q_H = h(T_s - T_w)
\]  (6.15)

Based on the reverse model, while, the temperature in the sample is lower than the boiling point of water condensation proceeds on the interface, resulting in an additional rate of heat generation in the material. At the boiling point, all the heat conducted into a local region is used for the evaporation. When the temperature in the sample is higher than the boiling point the drying
The boundary conditions of the model for various conditions of material surface temperature is given below:

\[ T_0 < T_m < T_{boil}; \quad h(T_g - T_{sf}) = -\lambda \left( \frac{\partial T}{\partial y} \right) + J_y \]  
\[ \text{(6.16)} \]

\[ T_m = T_{boil}; \quad h(T_g - T_{sf}) = -\lambda \left( \frac{\partial T}{\partial y} \right) - J_y \]  
\[ \text{(6.17)} \]

\[ T_m > T_{boil}; \quad J = 0 \quad \& \quad h(T_g - T_{sf}) = -\lambda \left( \frac{\partial T}{\partial y} \right) \]  
\[ \text{(6.18)} \]

At the start of drying, the temperature and moisture content are uniform in the sample.

\[ \text{At } t = 0 \text{ and } 0 < y < Y; \quad T = T_0, \quad M = M_0 \]

At the sample bottom, there is neither heat nor mass transfer

\[ y = 0; \quad \frac{\partial T}{\partial y} = 0; \quad \frac{\partial M}{\partial y} = 0 \]

The moisture diffusivity \( D_e \) which is an important parameter for actual drying is unknown for the steam drying study; hence the diffusivity relation which was used for the air drying of bagasse is used.

\[ D_e = D_o \exp \left( -\frac{E_u}{RT} \right) \]  
\[ \text{(6.19)} \]

### 6.3.3 Numerical Solution Procedure

The numerical solution procedure for the steam drying process also follows the same procedure as explained for the case of air drying. In the
model, the flow field is analysed by the CFD Fluent package, whereas the product bed section is analysed using MATLAB codes, considering steam as the drying medium. The heat and moisture transfer equations as explained by Equations (6.13) and (6.14), are solved using the finite volume method with the tri-diagonal matrix algorithm to solve the resulting algebraic equations. In the solution, a conservative first order implicit backward Euler integration approach is used for transient terms.

In this model the initial condensation of steam on the product is considered (Reverse Model). The code is looped to take care of the conditions of the product surface temperature \( T_{sf} \) as described by the reverse model:

\[
T_{sf} < T_{boil}, \\
T_{sf} = T_{boil} \text{ and} \\
T_{sf} > T_{boil}.
\]

In the energy balance equation if \( T_{sf} < T_{boil} \), additional heat in the form of latent heat is imparted to the product because of condensation. At the boiling point when \( T_{sf} = T_{boil} \), all the heat conducted to the material is used for evaporation. For \( T_{sf} > T_{boil} \), where the temperature on the surface greater than boiling temperature, evaporation ends. Using the obtained temperature \( T_m \), the moisture diffusivity which is a temperature dependent function is estimated. Using the moisture diffusivity obtained, the moisture equation is solved and the corresponding profile can be obtained by processing the results. Again, for the next time step, the average temperature and moisture at the surface is given as input to the Fluent for the bottom wall conditions and the iteration is continued until the desired time. The solution procedure is illustrated in the flow chart as shown in Figure 6.17.
Figure 6.17 Flow chart illustrating the solution procedure
6.3.4 Simulation Results

The steam drying process for bagasse is simulated under real operating conditions based on thin-layer drying kinetics. In this work an attempt is made to simulate the drying characteristics of the bagasse layer, using steam as the drying medium. The steam considered for this simulation work is at atmospheric pressures with temperatures at 130°C. In the model, the initial condensation effect due to steam drying is considered. In this analysis, the moisture diffusivity is considered as temperature dependent, as represented by Equation (6.19). The variable moisture diffusivity values are incorporated into the model to study the moisture change with the surface coordinate for different drying conditions. The steam drying model predictions are compared with those of the models for air drying; these are validated using the experimental results.

From the model predictions for steam drying with steam temperature at 130°C, product thickness 40 mm, the initial temperature and moisture content of the material at 50°C and 0.5 kg/kg of product respectively, the moisture content distributions for the bagasse layer are as shown in Figure 6.18. In the Figure, the moisture distributions are shown for different periods during the steam drying process.

As mentioned, the reverse model exhibits three distinct zones such as condensation, restoration, and the actual drying process. These zones are analyzed for the model considering a bed thickness of 40 mm. As the processing begins, this condensation process continued until the surface temperature reaches 100°C. This phenomenon continued for 45s of the initial period of drying for the bed thickness of 40 mm. During this time, due to the effect of the condensation of steam over the material surface, the latent heat is imparted to the material; hence the material temperature rises at a significant
rate during this stage of the drying process. The moisture profile during this initial condensation effect is represented in Figure 6.18(a); eventually, due to the condensation phenomenon, the moisture content of the product increases and the surface moisture content are almost doubled during this period of drying. It can be observed, that the condensed water also permeates into the interior during this stage. The end of this condensation phenomenon is called as the reverse point, which occurs when the surface temperature reaches 100°C. The occurrence of the reverse point during steam drying depends mainly on the heat transfer coefficient and the product bed thickness. For the bed thickness of 40 mm the reverse point occurred at 45s.

After the reverse point, the actual evaporation process takes place. The initial period of the evaporation process is the restoration process, which extends until the initial moisture content of the product is reached. The restoration period for steam drying depends mainly on the heat transfer coefficient and bed thickness employed during drying. However, for the simulated conditions at bed thickness of 40 mm, it is observed that the restoration process continued up to 195s. Figure 6.18(b) represents the moisture profile of the product during the restoration process at time t = 100s, where it can be observed that the surface moisture content starts decreasing due to evaporation. After restoration time, there exists no region with moisture content higher than the initial moisture content.

After the restoration period, the actual drying process proceeds, and during this process it can be observed that there is an improvement in the drying rate, this is attributed to the sudden rise in the material temperature due to the condensation effect at the initial stage of drying. The improvement in the drying rate is due to the fact that, the moisture transfer during drying depends on the moisture diffusivity, which in turn, varies with temperature. Figure 6.18(c and d) represents the moisture profile during the actual drying stage of the process at time t = 300 and 600s respectively.
Figure 6.18 Moisture profile at different stages of low pressure steam drying

a) Moisture Profile @ reverse point (t = 45s)

b) Moisture Profile during restoration process @ t = 100s

c) Moisture Profile during actual drying process @ t = 300s

d) Moisture Profile at the end of desired drying time @ t = 600s
The effect of the bed thickness on the reverse model is also comparatively studied for varied bed conditions of 20, 40 and 60 mm. Figure 6.19 shows the moisture profile at these bed conditions during the initial condensation phenomenon.

Figure 6.19  Moisture Profiles for varied bed thickness at corresponding reverse point time

It can be noted that as the bed thickness increases the reverse point extends and hence the moisture added rises with the bed thickness. Figure 6.20 shows the mean moisture content with time, for the varied bed conditions. It can be seen from the Figure that the reverse point time
correspondingly varied from 37, 45 and 57s for the bed thickness of 20, 40 and 60 mm. Also, it can be noted, that the corresponding restoration period for the drying conditions increased with the bed height.

![Graph showing moisture content over drying time for different bed thicknesses.]

**Figure 6.20 Comparison of mean moisture content for varied bed thickness**

From the profiles recorded using the model, the changes in mean moisture content for the bed thickness of 40 mm is represented for drying, with conditions of steam at 130°C and 180°C as shown in Figure 6.21. The Figure represents the initial stages of drying indicating the reverse process for the two steam temperatures used. From the analysis it can be noted that when the steam temperature rises from 130°C to 180°C, the restoration time decreases by 0.42 times for the conditions under which the simulation was carried out. And the maximum amount of condensed water decreases to about 0.52 times due to the increase of the heat transfer coefficients.
Figure 6.21 Comparison of drying results at different steam temperatures

This is further compared with those results from air drying simulation and the variation in the drying rate is recorded as shown in Figure 6.22. Due to the initial condensation effect the reverse phenomenon is recorded with the steam drying conditions; hence, the drying rate seems to be better for air drying conditions during the initial period. However, after the restoration stage the drying rate of the steam drying conditions is faster than that of the air drying conditions.

Figure 6.23 shows the characteristic drying rate curves derived from the results of Figure 6.22. From Figure 6.23 it is observed that, in the initial stages of superheated steam drying (pre-heating period) there is a region where the drying speed becomes negative due to condensation. Also, in this Figure, the characteristic drying rate curve for conventional air drying is shown to indicate the difference in the drying rate curve for steam and air drying.
Figure 6.22 Comparison of steam drying with air drying at different temperatures

Figure 6.23 Comparison of drying rate for steam and air drying simulations
6.4 CONCLUSION

A two-dimensional numerical modeling study was performed in this work using both air and steam as the drying medium. The model equations and the solution procedures for the analysis were discussed. The developed air drying model was capable of predicting the moisture removal of bagasse at varied drying conditions with a maximum deviation of 7.64%.

The steam drying model was developed considering the reverse process with initial condensation of steam. From the model results it is concluded, that due to the initial condensation phenomena the latent heat is added to the material in addition to the convective heat from the drying medium. This phenomenon improves the rate of drying for the steam drying process. Compared with the air drying process the drying rate of steam drying is distinctly ahead; hence, the duration of drying was noted to be considerably reduced. The effect of bed thickness on the reverse and restoration phenomena of steam drying is also discussed.