DRYING MODEL FOR SKIM MILK IN A VERTICAL COCURRENT SPRAY DRYER

A DISSERTATION

Submitted in partial fulfillment of the requirements for the award of the degree
of
MASTER OF TECHNOLOGY
in
CHEMICAL ENGINEERING
(With Specialization in Computer Aided Process Plant Design)

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CANDIDATE'S DECLARATION

I hereby declare that the work which is being presented in this dissertation entitled **"MODELLING OF SPRAY DRYER"**, in partial fulfillment of the requirements for the award of the degree of Master of Technology in Chemical Engineering with specialization **COMPUTER AIDED PROCESS PLANT DESIGN** of Indian Institute of Technology Roorkee, Roorkee is an authentic record of my own work carried out during the period from July 2006 to June 2007, under the guidance of **Dr. (Mrs.) SHASHI,** Assistant Professor, Department of Chemical Engineering, Indian Institute of Technology Roorkee.

I have not submitted the matter presented in this dissertation for the award of any other degree.

Date: 08-06-2007

Place: IIT Roorkee.

Nagini

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CERTIFICATE

This is to certify that the above statement made by the candidate is correct to the best of my knowledge.

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ABSTRACT

In this thesis, a mathematical model has been developed for analysis, design and simulation of a vertical co-current spray dryer. The model considers several important features of drying as described below.

- Drop/particle size ranges from 80 to 100 μm.
- Particle shrinks as a shrinking balloon in the constant rate zone, and after attaining critical moisture content the diameter of particle remains constant.
- Dependence of physical properties and other constitutive parameters on temperature.
- Porosity of particles.

In the model, drying medium is assumed to be saturated on the surface of the particle, intraparticle thermal gradients being small are neglected, and nozzle zone and any thermal degradation are not considered.

The model comprises a set of non-linear coupled ordinary differential equations. These are solved by mathematical equation solver MATLAB 7.0. Model has been validated with the data available in the literature.

Spray drying of skim milk is taken as the system for the purpose of analysis and simulation. An empirical correlation has been developed for the decrease in drying rate after the attainment of critical moisture content. This correlation is based on the data available in the literature, and it facilitates numerical simulation process. Detailed results of numerical simulation has been presented.

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Nagini

(Nagini Sabbineni)

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NOMENCLATURE

CHAPTERS-I & II

A	Net average area created per minute	m^2/min
a	Empirical constant in distribution equation	(-)
B	Constant	(-)
Bi(H)	Biot Number (Heat)	(-)
Bi(M)	Biot Number (Mass)	(-)
C	Molar density	mol/m^3
C_w	Molar density of water	mol/m^3
D_{AB}	Diffusivity	m^2/s
D_{eff}	Effective moisture diffusivity	m/s^2
D_0	Specific surface average particle size	m
d_P	Particle diameter	m
d_{vs}	Sauter mean drop diameter	m
C_P	Specific heat at constant pressure	J/Kg-K
G	Mass velocity	Kg/m-s
g	Acceleration due to gravity	m/s^2
h	Heat transfer coefficient	W/m^2-K
h_g	Air side heat transfer coefficient	W/m^2-^{0}C
j	Empirical constant	(-)
k_{drop}	Thermal conductivity of the drop	W/m-^{0}C
k_f	Thermal conductivity of gas film	W/m-K
k_g	Mass transfer coefficient	kmol/m^2-h
k_t	Turbulence kinetic energy per unit mass	m^2/s^2
L_w	Wetted periphery of centrifugal disk atomizer	m
m	Mass	kg
N	Rate of rotation	r.p.m
N_θ	The number of radians a liquid particle passes through before reaching the edge of the disk	rad
n	Number of particles	(-)
P	Power required to create new surface	W

p	Power	(-)
Q_a	Volumetric flow rate of air	m^3/s
Q_L	Volumetric flow rate of liquid	m^3/s
R	Gas constant	J/mol-K
R_d	Radius of centrifugal disk atomizer	m
r	Radial distance	m
r_o	Radial distance from the point of dispersion	m
T	Absolute temperature	K
t	Time	s
Δt_m	Mean temperature	K
v	Relative velocity between gas and liquid	m/s
V_r	Radial component of velocity	m/s
V_z	Instantaneous velocity in axial direction	m/s
X_w	Mole fraction of water	(-)
x	Particle size in distribution equation	m
z	Axial distance	m
θ	Angle	rad
θ_e	Evaporation time of liquid drops	s
μ_a	Viscosity of air	$N\text{-}s/m^2$
μ_L	Viscosity of liquid	$N\text{-}s/m^2$
ρ_a	Density of air	kg/m^3
ρ_f	Density of feed	kg/m^3
ρ_G	Density of gas	kg/m^3
Γ_Φ	Transport coefficient	(-)
ρ_L	Density of liquid	kg/m^3
λ	Latent heat of evaporation	J/kg
σ	Surface tension	N/m

CHAPTER-III & IV

Uppercase Letters

C	Drag coefficient	(-)
G	Mass of single particle	Kg
H	Humidity	Kg/Kg
L	Drying medium flow rate	Kg/s

N	Rate of Drying	Kg/m^2-s
T	Absolute temperature	K
V	Volume of single particle	m^3
X	Mass of the moisture per mass of the dry solid	Kg/Kg

Lowercase Letters

a	Interfacial area	m^2
a	Letter used in equation (4.24)	W/m-°C
b	Letter used in equation (4.24)	W/m-°C^2
c	Specific heat	J/Kg-°C
d	Diameter of particle	m
g	Acceleration due to gravity	m/s^2
h	Height of drying column	m
h_c	Heat transfer coefficient	W/m^2-°C
k	Themal conductivity of air	W/m-°C
p_i	Partial pressure	Kg/m-s^2
q	Heat	J
t	Temperature	°C
ν	Velocity of air	m/s

Greek Symbols

τ	Residence time of the particle in the chamber	s
ψ	Evaporation coefficient	Kg/m^2-s
β	Function, which takes into account the decrease of evaporation rate caused by falling rate period	(-)
λ	Latent heat of vaporization	J/Kg
ξ	Number of particles	(-)
Φ	Function	(-)
ρ	Density	Kg/m^3
ε	Porosity of a particle	(-)
μ	Viscosity of air	N-s/m^2

Subscripts

0	Reference point
D	Drag
G	In vapor phase
L	For drying medium (air)
b	Bulk
c	Critical
i	Interfacial
m	Moisture
p	Particle
P	Constant pressure
s	Solid
sw	Solid and water
w	Water
wb	Wet bulb

Dimensionless Numbers

Le	Lewis Number
Nu	Nusselt Number
Pr	Prandtl Number
Re	Reynolds Number
Sc	Schmidt Number

CHAPTER-I

INTRODUCTION

Drying as a thermal process of separation, plays a leading role in chemical engineering. In drying we are especially interested in methods, which can be applied for drying of dispersions. Several branches of chemical engineering demand these methods. Spray drying is nearly the most important method for drying foodstuff, detergents, pharmaceuticals etc.

The growing importance of spray drying is abundantly evident from the increasing number of industrial applications. In spite of these impressive developments and of the large number of fragmented experimental studies which have appeared in the technical literature, it is still mainly based on the extensive experience and the vast body of operating data which manufacturers have acquired over the years.

Spray drying is used to dry pharmaceuticals, fine chemicals, foods, many kinds of trade wastes, organic and inorganic chemicals, rubber latex, and clay slips. In the soap and detergent industries it has become a major operation, and has been developed into a highly specialized art. Unlike most other drying operations, spray drying requires the designer to consider several equally important operations in addition to drying. These operations usually include mixing and agitation to prepare suitably a liquid feed for atomization, de-aeration of the feed after preparation, pumping of sludge's, slurries, and clear liquids of all ranges of viscosity, atomization of simple and complex liquids, the conveying of solids, either pneumatically or mechanically, dust collection by either mechanical or electrical methods, or both, and in some cases, scrubbing of exhaust gases either to remove fumes or to recover valuable vapors.

1.1 Everyday Applications of Spray Dryer

Spray drying is widely used in almost all major industries. From foodstuffs to home fittings, it has many associations. Each product requires different powder requirements to be met during manufacture. These can be met only by spray drying. Foodstuffs which are produced by spray drying technique are instant coffee, coffee whitener, dried eggs, milk, soups, baby foods, powdered cheese and fruits. These have direct connection with spray dryer. Milk powders are produced in agglomerated form, whereas eggs, soup, coffee whitener have powdery and fruits have granular forms.

Apart from dried foodstuffs that are consumed directly, there are many spray dried products used in cooking. Examples include condiments (garlic, pimento), flavoring compounds and ingredients in biscuits and cakes. Meat, fresh fruits and vegetables are foodstuffs that have indirect connection with spray drying. Coming to household commodities spray drying is used in the manufacture of detergents, soaps, surface active agents and optical brighteners. The spray dried powder form is ideal for rapid assimilation into the body organs. Therefore pharmaceuticals like antibiotics are produced under the most aseptic conditions as finely divided powders, which are often made in to tablets prior to marketing. Spray drying technique is also used in the manufacture of cosmetics like face powder and lipstick.

Applications of spray drying to home fittings and furnishings are also extensive. Wall tiling is formed by pressing colored spray dried clays. Paints contain spray-dried pigments. Electrical insulation material is spray dried prior to pressing into parts for electronics and electric power supplies. Spray dried starch and its derivatives are widely used in ice-cream, confectionery, jellies, preserves, frozen fruit, soft drinks. In non-food manufacture, spray dried starch is used in textiles, papermaking, printing and adhesives.

1.2 Advantages of Spray Drying

The principal advantages are summarized as follows:

- Certain product properties and quality values may be effectively controlled and varied by spray drying.
    a. Product density can be varied within a given range.
    b. A particle shape approximating a sphere, sometimes hollow, sometimes solid, is usually obtained in spray drying. Generally speaking, such a particle shape is unobtainable by other drying methods.
    c. The particle size of the product may frequently be controlled or varied in a given range by control of the operating conditions.

d. Spray drying frequently preserves the quality of a product because drying is so rapid and the material in the hot drying zone is always so wet that it does not become overheated and degraded. Further more, the dry product does not become overheated if the gases cool sufficiently from the evaporation process.

- Spray drying is particularly suited to the atmospheric drying of certain heat-sensitive materials, such as foods, and pharmaceuticals, which otherwise would require high vacuum, low-temperature drying. However, spray drying is not always the answer to drying heat-sensitive materials, especially those which must be dried in the absence of oxygen and/or carbon dioxide. In certain applications spray drying has been performed in inert atmospheres in a closed system.
- Spray drying as it is practiced today may frequently show marked advantages for high tonnage production, since as the output increases, the drying cost per pound of product becomes less than that for other types of dryers. This is generally true when the moisture content of the feed to the spray dryer is not much greater than that of the feed to other types of dryers.
- A material dried in a spray dryer does not contact solid surfaces until it has become dry. This frequently simplifies corrosion problems and the selection of materials of construction.
- Spray drying may frequently simplify or eliminate other operations, such as filtration of the feed, and size reduction of the dry product.
- Since a spray dryer usually operates at temperatures ranging from 205°C to 540°C., its efficiency is comparable with that of other types of direct dryers; e.g., rotary dryers, and tunnel dryers.

1.3 Disadvantages of Spray Drying

Some disadvantages are inherent in the operation, and others are due to a lack of fundamental knowledge of the operation. The principal ones as spray drying is practiced today are:

- Low bulk densities are frequently obtained when a high density product is required. This is often the case for inorganic materials which are shipped in carload lots. Although the bulk density of a product from a spray dryer is subject to variation, it sometimes cannot attain the same value as that obtained from other types of dryers without resorting to further processing such as briquetting.
- In general, spray dryers are relatively inflexible. Thus, a spray dryer designed for fine atomization is generally incapable of producing a coarse product, if such is required.
- For a given capacity, larger evaporative loads are generally required for spray drying a given material than would be required with other types of dryers. This is due to the requirement that the material to be dried must be in a pumpable form for delivery to the atomizer, which usually requires the addition of water. This fact may be one reason for a common misconception that a spray dryer is inefficient. Actually, a spray dryer may be more efficient than other types from a heat utilization standpoint.
- In general, spray dryers as currently designed involve a higher initial investment than other types of continuous dryers except at high capacities; e.g., above 1500lb. of product/hr. For this reason, spray drying has not found wide application in the chemical industries to low capacity, continuous drying operations. However, it does compare favorably at low capacities with high vacuum, low-temperature drying operations.
- Frequently the problems of product recovery and dust collection increase the cost of drying by an appreciable factor, especially when bag filters or scrubbers are required to recover dust exhausted from a cyclone collector system.

1.4 General Description of Spray Drying Process

1.4.1 Spray drying

Spray drying is by definition the transformation of feed from a liquid state into a dried particulate form by spraying the feed into a hot drying medium. It is a one step, continuous particle processing operation involving drying. The feed can either be a solution, suspension or paste. The resulting dried product conforms to powders, granules or agglomerates, the form of which depends upon the type of dried product and, therefore, upon the physical and chemical properties of the feed, the dryer design and operations.

1.4.2 Spray dryer

A spray dryer is a large, usually vertical chamber. The hot gas, which acts as drying medium, is blown from top or bottom according to our requirement. A suitable atomizer from the top of the chamber sprays a feed solution, slurry or pumpable paste. Thus the droplets of feed are dispersed into stream of hot air. The moisture from the droplets is evaporated rapidly and the droplets get dried before they can be carried to the sides of chamber. The bulk of dried powder falls to the conical bottom of chamber to be removed from a stream of air to the dust collector. Fig. 1.1 shows a schematic diagram of spray dryer.

1.4.3 Classification of spray dryers

Based on the mode of operation spray dryers are classified into three groups.

- Co-current spray dryers
- Counter-current spray dryers
- Mixed type spray dryers

1.4.4 Basic principles of spray drying

A spray dryer operates on the basis of the creation of highly dispersed liquid state in a high temperature gas zone. Three equally important processes must occur. They are

1. Atomization
2. Drying of liquid drops
3. Spray gas mixing

1.4.4.1 Atomization

Atomization is the heart of spray drying operation. Its principle effects are to produce a finely divided product with special physical characteristic of particle shape and density. Three fundamentally different methods can be used to atomize liquids in spray dryers. These are

a. Pressure atomization by means of pressure nozzles.
b. Two-fluid or gas stream atomization usually with air or steam as the atomizing fluid.
c. High speed rotating disks, which atomize by bringing a liquid up to a high velocity by centrifugal force and discharging it into a hot gas zone.

Pressure atomization:

Studies of the disintegration of jets from pressure nozzles have led to the following conclusions

- If the liquid jet is turbulent throughout, it will disperse without application of any external force. This is because in a liquid stream in turbulent flow the liquid particles have definite radial velocity components so that when the jet is no longer confined by the orifice walls the particles are restrained only by the surface tension force of the liquid. Disintegration of the jet occurs as the surface tension is overcome. Thus it may be concluded that at high pressures, surface tension is a controlling factor in atomization.
- If the liquid jet is in semi turbulent flow i.e., with a laminar layer surrounding a turbulent core, then disintegration of the jet will occur after leaving the nozzle as the turbulent core forges ahead of the laminar layer.
- If the jet is in laminar flow, air friction or some other external disturbances is essential to its disintegration. This leads to the conclusion that at low nozzle pressure, viscosity will be a controlling factor. It also suggests the possibility that a jet in laminar flow will not atomize in vacuum.
- Regardless of the type of flow, disintegration or atomization is believed to be favored by air friction.
- The higher the viscosity of the liquid, the longer the breakup distance of the jet.
- As the pressure increases, the breakup distance decreases.
- The breakup of a jet is influenced by roughness of the orifice and other factors affecting the general turbulence conditions of the liquid as well as the smoothness of the liquid surface.

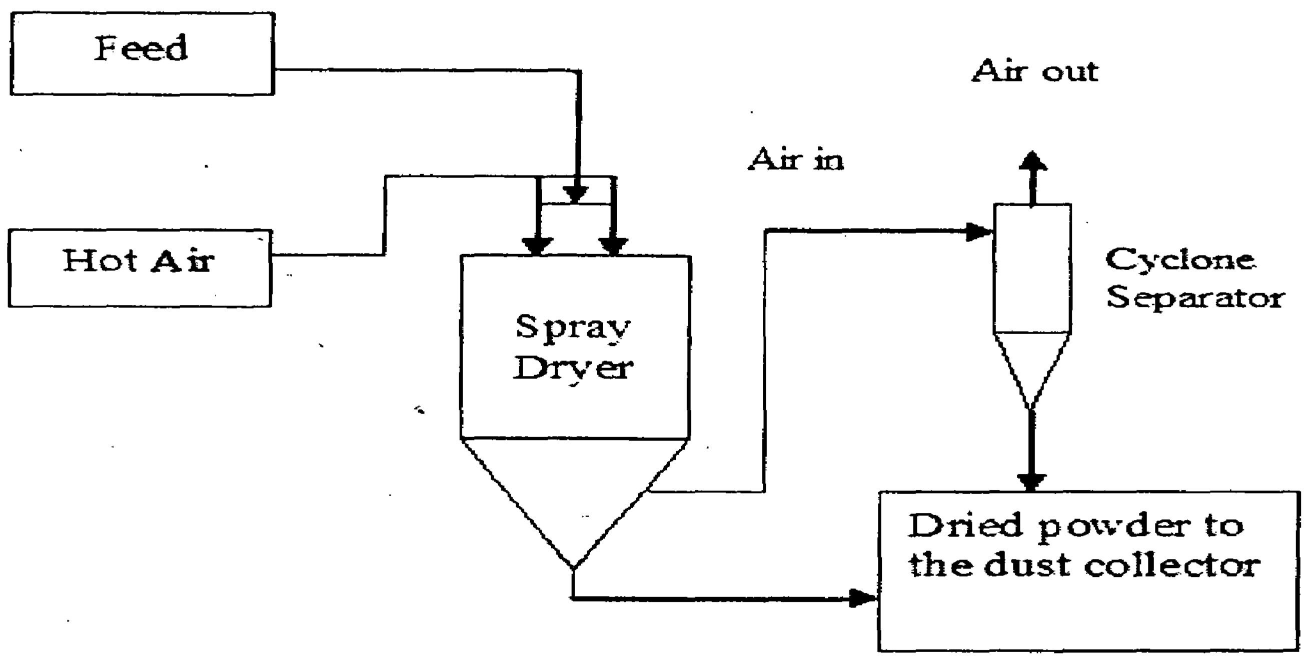


Fig. 1.1 Schematic diagram of Spray Dryer

Two fluid atomization:

To postulate a mechanism for this method of atomization, Castleman[19] assumed that

- It is a necessary step first to form ligaments from the large mass of liquids.
- The rate of collapse for this ligament follows Reyleigh's theory for the collapse of a liquid column.

Spark photograph by Lee showed that such ligaments actually are torn by air friction from the main body of the liquid. Studies by the Sauter, Nukiyama and Tanasawa, Lihaye and Lewis et. al.[19] showed that as the air velocity is increased, finer atomization is obtained resulting from finer ligaments being torn from the liquid at higher air velocities.

Nukiyama and Tansawa[19] presented the following empirical equation for predicting a surface average drop size for atomization by high velocity gas streams;

$$D_0 = \frac{585\sqrt{\sigma}}{v\sqrt{\rho}} + 597\left(\frac{\mu}{\sqrt{\sigma\rho}}\right)^{0.45}\left(\frac{1000Q_L}{Q_a}\right)^{1.5} \tag{1.1}$$

It can be seen that the effect of term involving viscosity, becomes negligible when the ratio of the volumetric rates of liquid and gas becomes less than 10^{-4}.

The distribution of particle size obtained by atomization with high velocity gas streams might be expressed by the following relationship [19]

$$\frac{dn}{dx} = ax^{p}e^{-mxj} \tag{1.2}$$

Where j is a function of nozzle design.

Rotating disk atomization:

This type has found wide spread application in spray drying. Operation of rotating disk atomizers consist of centrifugally accelerating a liquid to a high velocity before discharging it into a gas atmosphere. This method differs from the pressure nozzle. In this the liquid attains its velocity without being subjected to high pressures, frequently with a free liquid surface exposed to a gas phase during passage over the disk. In some designs, the liquid is forced through radial tubes.

Several types of rotating disk may be used to atomize liquids. The simplest of course, is a flat, smooth disk near the centre of which the liquid is deposited. If no slippage occurs between the liquid and disk, and if no frictional force retards the flow, the time required for the liquid to reach the disk periphery can be calculated by the following equations

$$N_\theta = arc\left[\cosh\left(r/_{r_o}\right)\right] \tag{1.3}$$

$$N_\theta = 2\pi Nt \tag{1.4}$$

A very useful correlation is that obtained by Fried Man [7]. They correlated the Sauter mean diameter to the disk radius in terms of dimensionless groups as follows

$$\frac{d_{vs}}{R_d} = 0\cdot 4\left(\frac{G}{\rho_t N R_d^2}\right)^{0\cdot 6}\left(\frac{\mu}{G}\right)^{0\cdot 2}\left(\frac{\sigma\rho_f L_w}{G^2}\right)^{0\cdot 1} \tag{1.5}$$

They found the following relationship between the largest droplet diameter and Sauter mean diameter.

$$d_m = 3d_{vs} \tag{1.6}$$

It is interesting to note that the power required to create new surface, based on thermodynamic considerations, is only a small fraction of that actually expanded a conventional atomizer. Thus to create new surface, the theoretical power required is

$$P = A\sigma \tag{1.7}$$

The particle trajectories of the various atomizer are a function of the type and design. In general, the spray pattern from nozzles tends to be conical in shape, the included angle of the cone seldom exceeding 120°. The spray pattern of the disk, however, resembles more nearly an umbrella, with the particle trajectories usually in the plane of the disk. The various trajectories and spray patterns of atomizers materially influence the design of a spray dryer, and the configuration of the drying chamber.

One controversial question concerning the various types of atomizers is that which type will produce the most uniform drop size distribution for given condition? However this has never been answered by comparative studies of each type. It is believed that with proper design and operation a pressure nozzle and a rotating disk can be made to produces sprays with approximately the same particle size distribution, even though each type may not produce a particle size distribution following the same mathematical laws. A comparative study of each atomizer under different conditions involves the tedious problem of sampling the spray and counting the drops in various size ranges. Several thousand drops must be counted in each sample for satisfactory accuracy. This laborious procedure has represented a tremendous obstacle to atomization research, but with the development of atomic drop counters. It is anticipated that considerable progress will be made on comparative studies of the various types of atomizers. It is well to point out that there may be a considerable difference between the drop size created at the atomizer and the finished product particle size. Puffing of the drop and agglomeration contribute to affect the change in original drop size.

It is proposed that a comparison of atomizers should involve the following fundamental aspects.

1. Nozzle capacity expressed as square meter of surface created per kg of liquid atomized per minute.
2. Efficiency expressed as power consumed per square meter of surface created.
3. Drop size distribution or sharpness index for identical flow rates.
4. Weight flow distribution.

A further comparison of nozzles and disks reveals the latter to be somewhat more flexible from the standpoint of variations in operating conditions. Thus for a given problem a disk might handle a variation of feed rate in a range of $\pm$ 25% of design capacity without greatly affecting the particle size of the product and without the necessity of changing operating conditions, provided excess power is available. In order to handle an increased capacity with nozzles, either the pressure is desired or larger nozzles must be substituted. Nozzles on the other hand are more acceptable to countercurrent flow operation because of the more confined nature of their spray pattern. Adoption of a disk to counter flow operation involves methods for bending the spray trajectories either by airflow or disk design, or both. Rather high air velocities are required to disrupt appreciably the trajectory of the spray of liquid near the spinning disks.

In general, both nozzles and disks will handle fluid of the same consistency. Disk may handle heavy sludge's, which cannot be satisfactorily pumped with piston pumps and atomized by pressure nozzle. On the other hand, pressure nozzle have been found well suited, for atomizing a viscous liquids which tends to form starting on disks, by superheating the liquid under pressure so that on ejection from the nozzle an added disrupting force due to sudden vapor formation aids atomization.

1.4.4.1.1 Selection method for atomization

Selection of the atomizer type depends upon the nature of the feed and desired characteristics of the desired product. In all atomizer types, increased amounts of energy available for liquid atomization result in spray having smaller droplet size. If the available atomization energy is held constant but the feed rate is increased spray having longer droplet sizes will result. The degree of atomization depends also upon the fluid properties of the feed materials, where higher values of viscosity and surface tension result in large droplet sizes for the same amount of available energy for atomization.

The most desirable characteristics of any atomizer are listed as follows

a. Simple construction.
b. Designs conducive to easy maintenance.
c. Available in both small and large sizes.
d. Spray size distributions controllable through adjustment to atomizer operating conditions.
e. Operated by standard pumping equipment.
f. Handle feed without internal wear.

Selection of atomizer based on spray characteristics:

With proper design and operation, nozzles and rotary atomizers can produce sprays having similar droplet size distribution. For a given spray drying application, the selection between rotary and nozzle atomizers involves the following considerations.

1. The feed capacity range of the atomizer where complete atomization is attained.
2. The power requirement of the atomizer to attain complete atomization of the feed (atomizer efficiency).
3. The droplet size distribution at identical feed rates.
4. The maximum and minimum droplet size levels (spray homogeneity).
5. The operational flexibility.

6. The drying chamber design most suitable for atomizer operation.
7. The feed properties most suitable for atomizer operation.

Droplet size distribution from rotary and nozzle atomizers have similar characteristics at low to intermediate feed rates. At high feed rates spray homogeneity is generally greater with rotary atomizers. Atomization by rotary atomizers is generally more flexible than nozzle atomization from the operational standpoint. When selecting atomizers, dryer design plays an important role. From this view point, nozzles are more adaptable. The confined nature of nozzle sprays enables any nozzle positioning in either co-current, counter-current or mixed-flow drying chambers with air dispersers creating rotating or parallel air patterns. Rotary atomizers generally require a rotating air pattern.

1.4.4.2 Drying of drops

The second important phase of spray drying concerns the drying rate of drops. Considerable attention has been given to the evaporation of pure liquid drops. Sherwood and Williams [19] correlated much of the significant published data on mass transfer from liquid drops, and on heat transfer to solids spherical particles. Their recommended equations for mass transfer for three different ranges of Reynolds number may be written as follows

I. For $N_{Re} < 4 \cdot 0$,

$$N_{Nu} = \left(\frac{k_g RT d_p}{D_{AB}} \right) = 2 \cdot 0 \tag{1.8}$$

II. For $4 \cdot 0 < N_{Re} < 400$,

$$N_{Nu} = 1 \cdot 5 (N_{Sc})^{1/3} (N_{Re})^{0 \cdot 35} \tag{1.9}$$

III. For $N_{Re} > 400$,

$$N_{Nu} = 0 \cdot 43 (N_{Sc})^{1/3} (N_{Re})^{0 \cdot 56}$$

(1.10)

Frosseling [1] the following equation for all values of N_{Re}

$$N_{Nu} = \left(\frac{k_g RT d_p}{D_{AB}} \right) = 2\left[1 + 0 \cdot 276 (N_{Sc})^{1/3} (N_{Re})^{0 \cdot 5}\right] \tag{1.11}$$

The corresponding equation for heat transfer is

$$N_{Nu} = \frac{hd_p}{k_f} = 2\left[1 + 0\cdot 276(N_{Pr})^{0.5}(N_{Re})^{0.5}\right] \tag{1.12}$$

It is evident that equation (1.8) and (1.9) are equivalent at low values of Reynolds number.

1.4.4.2.1 Evaporation life time

Since the rate of evaporation of a pure liquid drop is proportional to the rate of heat transfer to its surface, a heat balance gives

$$-\rho_L \frac{d}{d\theta_e}\left(\frac{\pi d_p^3}{6}\right) = \left(\frac{h\pi d_p^2 \Delta t_m}{\lambda}\right) \tag{1.13}$$

This equation can be reduced to

$$\frac{2\Delta t_m}{\rho_L \lambda} d\theta_e = -\frac{dd_p}{h} \tag{1.14}$$

If $\frac{2\Delta t_m}{\rho_L \lambda}$ is effectively constant, a condition that is often met under actual drying operations, integration of (1.13) gives the lifetime of a drop. Thus,

$$\theta_e = \frac{\rho_L \lambda}{2\Delta t_m} \int_0^{D_{p1}} \frac{dd_p}{h} \tag{1.15}$$

This equation can be generalized by substituting h from equation (1.12) as

$$\theta_e = \frac{\rho_L \lambda}{4\Delta t_m k_t} \int_0^{D_{p0}} \frac{d_p dd_p}{\left[1 + 0\cdot 276\left(\frac{\rho\mu_a}{k_f}\right)^{1/3}\left(\frac{d_p v\rho}{\mu_a}\right)^{0.5}\right]} \tag{1.16}$$

This equation is useful only if relationship between d_p and the velocity v, in the Reynolds number is known.

Case 1: Drop falling at zero relative velocity (i.e., v=0)

For zero relative velocity from equation (1.12) we have the value of h as

$$h = \frac{2k_f}{d_p} \tag{1.17}$$

and hence by substituting the value of h into the equation (1.15) we get the time for complete evaporation as

$$\theta_e = \frac{\lambda \rho_L d_{P1}^2}{8 k_f \Delta t_m} \quad (1.18)$$

Thus the lifetime of a drop for zero relative velocity is found to be proportional to the square of its initial diameter and inversely proportional to Δt_m and the gas film thermal conductivity k_f.

Case 2: Drop falling at its terminal velocity

If drop is falling at its terminal velocity in Stoke's law region or Stream line flow region, the relation between terminal velocity and drop diameter can be expressed as

$$v_t = \frac{g(\rho_L - \rho_a) d_P^2}{18 \mu_a} \quad (1.19)$$

and equation (1.16) becomes

$$\theta_e = \frac{\rho_L \lambda d_{P1}^2}{8 \Delta t_m k_f} \left[1 - \frac{2}{d_{P1}^2} \int \frac{B d_P^{2.5} dd_P}{1 + B d_P^{1.5}} \right] \quad (1.20)$$

where,

$$B = 0 \cdot 276 \left(\frac{c_P \mu_a}{k_f} \right)^{1/3} \left[\frac{g(\rho_L - \rho_a)\rho_a}{18 \mu_a^2} \right]^{0.5} \quad (1.21)$$

The value of B for air will be varying from approximately 240 at 400K to 81 at 1000K.

The Reynolds number range, which has greatest significance in spray drying application, is 10^{-1} to 10^{-2}. The Stoke's law region is generally accepted to be in the range of N_{Re}<2.0. Particle of 50μm and less subjected to relative velocities 90 to 150 m/s will not exceed a Reynolds number of 200. However the timing during which such relative velocities exist, is short, since the particle is rapidly speeded up or slowed down to the air velocity. Consequently it would be expected that a sprayed particle in a spray dryer would be, for the most part, in the Stoke's law region of flow.

1.4.4.3 Spray gas mixing

The third important step in the spray drying process is effective mixing of the spray and gas. The various spray dryer designs today reflect in large measure the various attempts that have been made to solve this spray gas mixing problem. The flow of gas and spray may be co-current, countercurrent, or a combination of both. The spray may be directed vertically down or up, horizontally, or at an angle to the vertical. Spray dryers embodying all of these features are in existence. A factor greatly influencing this problem of mixing is the trajectory pattern of the various atomizing devices. Thus, the problem of effective intimate mixing of a spray coming off at all points around a rotating disk, may be more difficult than mixing the conical shaped spray from a pressure nozzle.

1.5 Design of spray dryer

1.5.1 Design criterion:

An important consideration in spray dryer design is the selection of a suitable criterion against which the adequacy of the design results can be tested. A number of such criteria are the attainment of a specified humidity and temperature in the exit gas, attainment of a given moisture content for a specified fraction of the drop size distribution (DSD) in a specified zone of the chamber, maximum thermal efficiency, etc.

1.5.2 Method of design [7]

The design of spray dryer involves the following steps

- From the desired product size distribution and nature of the product , the DSD near the nozzle for an optimum concentration of the feed is determined, and then the largest droplet size is predicted.
- The drying gas flow patterns and the type of the chamber are selected on the basis of the nature of the product and the drying characteristics of the feed solution.
- The droplet trajectories both in the nozzle and entrainment zones are to be calculated.

The calculations are repeated until the dryer dimensions are just adequate to satisfy the drying condition.

1.5.3 Design and operating faults to avoid

- Poor spray gas mixing: The efficiency of spray dryer is severely impaired if the spray and hot gases are not efficiently mixed. Efficient mixing implies uniformity in distribution as well as in speed.
- Product degradation: The designer of a spray dryer should avoid the passing of the product through the fans of other equipment causing particle degradation. This increases the problem of dust recovery as well as degradation of the properties of the product.
- Nature of non uniform performance: A series of operating faults may be encountered in connection with non uniform atomizer performance. Non uniformity of the atomization may be of two types:
  - Non uniformity particle size distribution, i.e., the criterion of a higher percentage of fine or coarse particles than the usual size distribution function for the spray would predict.
  - Non uniform wet distribution in the spray pattern.

The first type of non uniformity will lead either to over heating and degrading of fine particles, or non drying of coarse particles and sticking of wet material on the chamber. This fault can be avoided by selection of proper atomizer within the range of design capacity. The second type of non uniformity is non uniform spray distribution and is generally due to faulty atomizer design. Other non uniformities may concern to impair atomizer and dryer performance. These are, non uniform flow feed and non uniform feed concentration.

These are obviously operational problems and are responsibilities of the product manager. It cannot be overstressed that satisfactory spray dryer performance depends on steady, balanced or symmetrical airflow in to the dryer, uniform or balanced atomizer performance, and a steady uniform feed. Fluctuations in any of these conditions lead to poor dryer performance.

1.6 Description of the problem

Spray dryers are widely used in industry. Although they are mostly found in the end point of the process, they have an important place in whole process, not in the last because of their capital investment, size and operating cost.

Several mathematical models have been developed in the past decades. Most of them are based on crude assumptions of what takes place in the spray drying chamber, especially with respect to spray air mixing. In most of the models, variations in air temperature and residence times are neglected; the flow of air and particle is not considered. The problem associated with this approach becomes apparent when one considers a handful of product. It consists of thousands of particles. A measure of the overall quality of the sample is the average of those thousands of particles, each having a different size, air temperature, residence time and air humidity. Further, many properties are non linear in the composing parameter. For example moisture content is strongly non linear in time. It is also very difficult to describe the exact particle trajectory. Hence complete description of performance of a industrial spray dryer is still a difficult task to process engineers.

1.7 Objectives of the thesis

Based on the reviewed literature following objectives have been planned

- To develop the mathematical model of a spray dryer and to study the effect of operating and design parameters on its performance.
- To solve the model equations by using suitable numerical technique or by using mathematical equation solver like MATLAB 7.0.
- To validate the proposed model with available data.

1.8 System

Before second world war, dried milk was not well known to the public, yet it is an old product. Dry milk is a very essential and useful food for daily life. Therefore, great improvements have been made in its manufacture. Spray drying is the most widely used method to produce and preserve a high quality milk powder. The system proposed to study is spray drying of milk.

1.9 Organization of Thesis

Chapter-I includes the introduction of thesis, applications of spray drying technique, it's advantages and disadvantages, brief process description, basic principles of spray drying, design method of spray dryer, problem description, and objectives of the thesis. Chapter-II gives a brief literature review and the comparative study of the models. Chapter-III includes the model development. Chapter-IV describe different constitutive relationships, related to corresponding constitutive properties, some of them taken from literature and remaining derived using fundamental laws. Chapter-V describes the solution technique and initial conditions. Chapter-VI gives results and discussion obtained from this model. Chapter-VII gives the concluding remarks and recommendations.

LITERATURE REVIEW

2.0 Introduction

Literature review is the heart of every dissertation work. In this chapter we will discuss briefly different literature available on spray drying.

The principle of spray drying is the atomization of the liquid phase into tiny drops and their deposition in hot gas stream, because the atomization of liquid increases the surface area for heat and mass transfer (the interfacial surface area per volume of all drops) and the increased surface area causes a shorter drying period.

When dealing with multiphase flows, one always faces the difficult problems of accounting the momentum, mass and energy couplings between two phases. The exact analysis of these flows in three dimensional motion becomes enormously complex. Due to these complexities, spray dryer design operations are not well understood. Very complicated droplet trajectories and their integrations with the air flow prevent us from making a simple design scheme for various types of spray dryers. Therefore, ***Modelling of Spray Dryer*** is very essential.

Many papers have been published on the development of spray dryer by different authors. These papers have been reviewed and in this chapter a brief description of their work has been presented. Useful information about spray dryer has also been given. We divide this chapter into four parts. First part gives the brief description about those papers, which deal with the process description and general information about the process and the system. Second part gives different models for spray dryer with available software like CFD. Third part describes the motivation for the proposed work. Fourth part gives the concluding remarks.

2.1 Process Description

There are a number of research workers who have worked on the process of spray drying. They studied about various parameters and their effect on the process. In this part we give the brief description of their work.

Marshall *et al.* **(1950)** have described the fundamentals and principles of spray dryer operation. They have also given a brief description of the elements of spray dryer design. They have dealt with the advantages and disadvantages of spray dryer. Further, few parameters such as operating variables, atomization selection procedure, selection of drying temperature, method of product removal, design and operating faults, spray dryer performance and spray dryer economics have also been studied.

Buckham *et al.* **(1955)** have described the factor effecting gas recirculation and particle expansion in spray drying. All possible operating variables are considered and are reduced to those, which directly and indirectly affect the drying operation. These are the size and size distribution of the spray droplets, the gas temperature of the drying zone, and the feed concentration. The principle product property is the ratio of the final to original particle diameter. The average particle expansion ratio for a given run increases linearly with the mean product particle mass. The order of the magnitude of the particle expansion is such as to keep the surface area to mass ratio nearly uniform for all particles. Gas circulation in the drying chamber occurred to the extent of about three fourth of complete mixing as indicated by vertical drying zone temperature and humidity distribution. The vertical gas mixing results in a relatively constant gas temperature in the drying zone, and greatly increases the spread of the residence times of small particles in the drying chamber over that which otherwise would be extracted.

Santosh (1972) have analyzed about the volatile retention during drying of liquid foods. They analyzed through a ternary diffusion model. Flux equations for both water and trace organic components are solved numerically by using finite difference formula for typical drying situations. Concentration profile for both water and dilute volatile components are experimentally measured during nearly isothermal drying of gelled slabs of synthetic sugar solutions and natural fruit juice concentrates. The predicted internal maximum concentration in volatiles is born out by the experiment and a satisfactory agreement of observed with predicted volatiles retention is found, within the limits of experiment.

Gauvin *et al.* (1976) have discussed about the major parameters in the design of spray dryers. Lagrangian approach, combining experimental data with theoretical concepts, is proposed to develop the design procedure. They have also proposed computational methods to calculate droplet trajectories, and also proposed the optimal chamber dimensions and operating conditions for maximum thermal efficiency and minimum operating cost. Applications of these basic principles are illustrated by the design of an industrial size spray drying chamber for a specific feed solution and production rate. It has been shown that a spray dryer can be designed on a theoretical basis, providing a minimum amount of information.

Crosby *et al.*(1977) have reviewed the effects of foaming spray dried feed stocks on product properties, drying characteristics, and equipment performance. They have also discussed the methods of foam generation and the physical nature of foam drops. Illustrative drying histories of single, foamed drops are presented and limiting models for description of drying mechanisms are suggested.

Kieckbusch ***et al.*** **(1980)** have presented the mechanism of volatile loss during atomization in spray drying. They have measured the losses of volatile acetates in the vicinity of pressure nozzle during the spraying of sucrose and malto-dextrine solutions. Large losses occur very near the atomizer, where only a small portion of the water has been evaporated. The effects of atomizer design, nozzle pressure, sucrose concentration, air flow rate, air temperature, and liquid feed temperature have all been measured. Correlation of volatile loss vs. percent water evaporated has also accounted for most of the effects of changes in spray pattern and drop size distribution. Individual contribution for gas and liquid phase mass transfer are determined from the relative retentions of acetates of different molecular weight and from the effect of sucrose concentration. In the expanding film at the nozzle, both gas and liquid phase resistances are important, but the loss becomes entirely liquid phase controlled once drops are formed.

Greenwald *et al.* (1982) have introduced the mechanism of particle expansion in spray drying of foods. These include formation of bubbles of water vapor, entrainment of air during atomization and desorption of air. Morphological changes were investigated as a function of temperature, initial drop size, types of solute and solute concentration. The assumption taken for this work is that the water transport in the drying drop can be described as binary diffusion in a homogeneous liquid.

Model equation:

$$\frac{\partial C_w}{\partial t}+\frac{1}{r^2}\frac{\partial}{\partial r}\left(r^2 C_w V_r\right)=\frac{1}{r^2}\frac{\partial}{\partial r}\left(r^2 C D_{AB}\frac{\partial X}{\partial r}\right) \tag{2.1}$$

Usui *et al.* (1985) have analyzed turbulent flow in a spray drying chamber. The turbulent fluctuation in the spray drying chamber was extremely large than usual turbulent level of free jet or tube flow. They have suggested that the large fluctuations can be attributed to the unsteady nature of the down stream jet, caused by the interaction between the shape of the chamber and the downward jet. They have developed the following equation for velocity profile.

$$\overline{V_r}=-\frac{1}{r}\int_0^r r\left(\frac{\partial V_z}{\partial h}\right)dr \tag{2.2}$$

Tarric *et al.* (1990) have monitored the morphological changes of drops of sucrose and maltodextrin solution, coffee extract and skim milk during drying. The effect of composition, initial concentration, air temperature and presence or absence of dissolved gases is determined. Particles of coffee extract from the falling drop dryer evidences surface blowholes, replicate what is observed in commercial spray dried coffee. This phenomena is rationalized in terms of viscous resistance to sealing flows.

Langrish *et al.* (2001) have introduced the assessment of the characteristics of a drying curve for milk powder. These information's are used in computational fluid dynamics modelling for different particle sizes and drying conditions typically of those in spray dryers. A review of previous literature, with particle diameters greater than 2mm, has shown that a linear falling rate curve is an acceptable approximation for the hindered drying of this material. For the diameters of milk particles in spray dryers, the drying times are predicted using a linear falling rate and are found to be of the order of 1s compared with residence times of 20-80 s in full scale equipment.

Zbicinski *et al.* (2002) developed a method for measuring drying kinetics of different products in a disperse system. In order to carry out the experiments a 9m long spray drying tunnel was designed, built and tested. The Phase Doppler Anemometry (PDA) technique was used to determine initial spray atomization parameters, the structure of spray during drying, particle size distribution, velocity of the particles, mass concentration of the liquid phase, etc. Measurements were made at different distances from the atomizer and in the cross-section of the spray stream. Maltodextrin was used as a raw material in the experiments. Extensive spray drying tests were performed to determine the influence of operating process parameters on spray. Examples of the drying kinetic tests are presented and discussed in the paper.

Fabrizio Scala *et al.* (2004) have carried Spray-dry flue gas desulphurization (FGD) in conjunction with bag house particulate collection. This represents a viable alternative to wet scrubbing in boilers burning low to medium sulfur coal or fuel oil. The advantages of spray-drying over other technologies include: the production of a dry waste byproduct not requesting sludge handling equipment; no scaling and corrosion problems enabling the use of cheaper materials; smaller space needed and possibility of easily retrofitting existing plants; no requirement of flue gas reheating; flexibility in operation with regard to varying boiler load; low energy consumption; reduced installation and operating costs. On the contrary, spray-dryers hardly exceed 70% SO_2 removal efficiency at 1–2 calcium to sulfur ratios (Ca/S), as opposed to values higher than 90% for wet scrubbing, making this technology attractive when SO_2 concentration in the flue gas is relatively low. This process is divided into three steps: a first short phase after atomization in which the droplets decelerate until they reach their terminal velocity; a second 'constant rate drying' phase, accounting for most of the sulfur removal, until the solid shell starts to form at the surface of the drops; a third 'falling rate drying' phase until the particles are dried. The dried solid product leaves the spray-dry chamber holding typically only few percents of free moisture and is separated in a downstream collection device, usually a bag house.

Kentish *et al.* (2005) considered the effect of variable milk composition, specifically the quantity of fat, lactose and protein, upon mass transfer resistance during drying using a combination of experimental measurement and mathematical modelling. Drying by convective heat transfer is the usual mechanism studied since it is an essential element of any spray drying process. Typically the temperature in the drying milk sample is taken to be uniform and to change with time according to the opposing effects of convective heating from the external gas stream and heat loss due to vaporization. In his work, he used microwave heating which delivers heat volumetrically, rather than to the surface of the sample. This approach has the potential to deliver heat more evenly to the interior of the sample, as long as the microwave penetration depth is not exceeded. In this work, he used a thin uniform milk layer (3mm depth) as a sample rather than a spherical drop. He assumed that the temperature within this sample is constant as the layer depth is much smaller than the microwave penetration depth and the Biot number is below 0.1 under the experimental conditions.

Adhikari *et al.* (2005) showed that a maltodextrin drop exhibited peak stickiness at average moisture (u) $\approx$ 1.0 and the drop surface became completely non-sticky at u = 0.69 while drying in a 63°C and 2.5% relative humidity air stream. They further showed that the T_g of the drop based on average moistures, u = 1.0 and u = 0.69 was −95.9 and −81.1°C, respectively, and that the drop surface was completely non-sticky even when the drop temperature was about 138°C above this glass transition temperature. These findings cannot be explained based on the sticky point temperature concept. This concept fails to explain the stickiness (adhesion) of foods on the equipment surfaces for three reasons. Firstly, the results obtained from sticky point tests are drawn from experiments carried out from powders with moistures (u) below 0.1 and it is not reported that these results can be extrapolated to higher moistures. Secondly, in the sticky point tests, the cohesive force between the particle surfaces is dominant rather than the adhesive force at the food-equipment interface. Thirdly, the samples in sticky point tests are equilibrated to have uniform moisture, while sharp moisture gradients exist in a drop subjected to convective drying. Hence, it is unlikely that the conclusions drawn from sticky point tests are automatically applicable to explain the surface stickiness of drops subjected to dynamic drying conditions.

2.2 Mathematical Modelling Studies

Mathematical modelling of spray dryer has been a subject of quiet a number of research works. This is connected with the much practical importance of this drying method and with some difficulties in a complete mathematical description of the method. Recently Zbicinski *et al.* proposed a classification of the mathematical models based on the type of equations, which describe the process of spray drying in mathematical model, presented in the literature. They classified the entire mathematical model into three groups. The first group consists of those models whose equations can be analytically solved. The second group encompasses the models in which the type of equations is similar to that in the first group but due to expansions of few simplifying assumptions, which causes that numerical solution is essential. The third group of models comprises the balances in form of differential equations. In this chapter we will describe the third group of models.

Parti *et al.* (1974) have developed a mathematical model for dimensioning of spray dryer. To develop the model they have taken following assumptions:

- The particle velocity is parallel to the axis of the equipment.
- The diameters of the particles are of same size.
- The evaporation takes place only from the outer surface of the particle known as evaporative surface does not recede into the particle.
- The sizes of the drops are generally so small, that the thermal gradient inside the drops can be neglected. So it can be inferred that the interfacial surface temperature is equal to average temperature of the particle.
- Operation is at steady state.
- Drying medium is saturated on the surface of the particle.

Wijihuizen *et al.* (1979) have presented a theoretical study of the inactivation of phosphates during spray drying of skim milk. From the unsteady state diffusion equation the water concentration distribution is calculated as a function of time. Simultaneously the thermal degradation of the enzyme alkaline phosphate is calculated an integrated over the particle. Effect accounted for temperature and concentration dependence of water, diffusion coefficient and degradation rate constant, and the shrinkage due to water loss. Droplet geometries considered for both solid sphere and hollow sphere model (with air bubble in the centre). They have also investigated the effect on the final phosphates retention of the following process variables: dissolved solid content of the feed, air temperature, particle size, and the gas bubble size. To perform this work they have taken following assumptions

1. The droplets are assumed to be perfectly sphere and to consist either totally of liquid (solid sphere model) or a central air bubble (hollow sphere model) or a central air bubble (hollow sphere model).
2. Air in the dryer is ideally mixed with moisture.
3. Initial radius is so small that during the whole drying period heat and mass transfer coefficient for the air phase can be described accurately by Sh = Nu =2.0
4. Water transport inside the droplet is spherically symmetric and inside the droplet can effectively be described by binary diffusion equation for the water and dissolved solids.
5. Mechanical equilibrium exists between the pressure inside the gas bubble and outside the droplet for hollow sphere model.

•**Sano *et al.* (1982)** have derived equations to describe the evaporation from a single, isolated, spherical droplet containing material to yield a solid particle, which may be hollow. To derive the equation the following assumptions have been considered.

- Once the equilibrium vapor pressure of the moisture inside the particle exceeds the pressure of the ambient air, the particles inflates and rupture forming a hollow sphere.
- Only a single inflation rupture takes place.

- The void radius remains same, while the outer radius R shrinks from R_{max}.
- The maximum radius does not change but void radius increases due to the moisture loss.

Zbiciniski ***et al.*** **(1988)** have introduced a mathematical model for spray drying. A novel mathematical model have been proposed and solved. In the model experimental data distributions of particle diameters during spraying are used. In their calculation physicochemical parameters of the drying agent along the cross section of the dryer are assumed to be constant. Trajectories of changes in temperature, humidity and temperature of the drying agent, moisture content and diameters of particular fractions of particles along the dryer height are obtained. A comparison of theoretical and experimental data revealed that they were in good agreement. To develop the model they have taken the following simplified assumptions:

- In the dryer cross sectional area, air temperature and humidity are constant.
- There is no temperature gradient inside the particle.
- Flow of the drying agent is co-current. Air is supplied to the dryer in the axial direction.

Clement ***et al.*** **(1991)** have constructed a dynamic model for spray drying process. The model describes the gas phase as a single well mixed mass, and the solid phase as a number of separate, well mixed masses each having different residence time. A shrinkage core model is considered to describe the evaporation from the droplets. The model contains only one adjustable parameter that is the diffusion coefficient of water in the droplets. To develop the model they have taken following assumptions:

- The particles are assumed to be well mixed in the chamber, all experiencing the same boundary condition.
- There is no temperature gradient inside the particle.
- The moisture content in the particle is assumed to be evenly distributed during the first period of drying.

Langrish *et al.* (1994) have minimized the deposition rate in a spray dryer. They have solved the equation of continuity and Navier-Stoke's equation inside the dryer using k-ϵ method of turbulence. The numerical simulation has been used to explore methods for decreasing the wall deposition rate, simple modifications to the air inlet geometry (to eliminate swirl in the inlet air), and a reduction in spray cone angle from 60° to 45° with in the constraints imposed by the experimental equipment. Their work has suggested that the maximum spray cone angle (60°) and maximum degree of swirl in the inlet air (62°) tend to minimize the wall deposition rate. To perform this work they have taken the following assumptions.

- All particles have been assumed to stick to the wall, and any re-suspension of particles due to bouncing or entrainment of wall deposits has been ignored.
- First order drying kinetics is applied, giving a linear falling rate curve.
- Radial component of velocity of air is neglected.
- The length scale of turbulence is assumed to be equal with the diameter of the opening to the drying chamber.

A general form of governing equations for axially symmetric flow in cylindrical coordinates, including the equation of continuity is given below:

$$\frac{\partial(\rho_G u\Phi)}{\partial x}+\frac{1}{r}\frac{\partial(r\rho_G v\Phi)}{\partial r}=\frac{\partial}{\partial x}\left(\Gamma_\Phi\frac{\partial\Phi}{\partial x}\right)+\frac{1}{r}\frac{\partial}{\partial r}\left(r\Gamma_\Phi\frac{\partial\Phi}{\partial r}\right)+S_\Phi \tag{2.3}$$

Perez-Correa *et al.* (1995) have introduced a simulation study on dynamic behavior of spray dryer. This paper presents a new model for predicting the dynamic behavior of milk spray dryer, and also develops and evaluates a simple and robust control structure by means of simulations. The new model includes a rigorous energy balance in the gaseous phase, and considers various droplet size and density. The predicted response of this model differs appreciably from the response of the models found in the literature. The control structure manipulates the inlet liquid flow rate and measures the humidity of exiting gas. With the aim of improving the control system's robustness, a cascade control is proposed, based on periodic off-line measurement of solid water content at the outlet. This system is capable of controlling solid humidity as it leaves the dryer, despite any kind of disturbances.

Cakaloz ***et al.*** **(1997)** have introduced model for horizontal spray dryer. The moisture content changes under optimum drying conditions of a droplet of $\alpha-$amylase solution leaving the nozzle in a specially designed horizontal spray dryer of $6m \times 3m \times 3m$ size, the gas entered at 155°C co-currently with the $\alpha-$amylase solution being sprayed from two pressure nozzles. The particles were collected in a channel through which they were transported to the cyclone pneumatically. The air rate was adjusted to keep the temperature below 60°C during the powder travel in the channel. During the operation, the activity loss of the powder was determined as 9%. The fully dry, equilibrium and critical moisture contents of the powder were determined by offline experiments. These values were fed to the computer program to predict the final moisture contents of the particles. A good agreement between the predicted and experimental values was found.

Straatsma ***et al.*** **(1999)** have stated that the thermal load of food products during drying is an important factor for the final powder quality. The exposure to heat can result in the formation of insoluble material, which is undesirable especially for instant powders. In this paper it is demonstrated how to carry out experiments in order to derive a kinetic model which describes the rate of formation of insoluble material during the drying of milk. It was proven possible to achieve a good fit between the calculated values of the insolubility index obtained from an idealized model and the values obtained experimentally. This method is also suitable for other food products. The kinetic model can be used as a sub model for drying simulation models that describe the behavior of the individual particles during spray drying. The combination of these models is an effective tool in giving indications of how to adapt industrial dryers to reduce the insolubility of the powder and get a better product quality.

Hill ***et al.*** **(2000)** presented a paper on flue gas desulphurization by spray dry absorption. In this paper he tells that the absorption efficiency of sulphurdioxide in spray dry absorption depends on the superposition of the absorption process with the drying process. Experiments in this work are performed with an artificial flue gas which contains only sulphur dioxide as an acid component in order to determine the influence of operating parameters on absorption efficiency systematically. A model has been established considering heat and mass transfer processes for a single droplet and the two phase flow inside the spray dryer. Model predictions are compared with experimental data showing the influence of drying conditions and stoichiometric ratio on the absorption efficiency.

Fabrizio Scala *et al.* (2004) presented a detailed single-drop model for gas absorption followed by an instantaneous irreversible chemical reaction for a rigid droplet containing sparingly soluble fine reactant particles. The model takes into account external
and internal mass transfer resistances together with slurry particles dissolution. Under suitable assumptions, the model was solved analytically giving a simple and easy to handle expression for the instantaneous gas absorption rate to the droplet. In this paper the expression derived is applied to the SO_2-lime slurry system and combined to a steady state one-dimensional spray-dryer model. Combination of the two models allows to easily carry out the material balance on sulfur dioxide along the column in order to calculate the desulphurization efficiency profile. The fate of the droplets is followed from atomization until formation of the crust (i.e. during the 'decelerating' and 'constant rate drying' phases). The model has been used to predict the influence of the main operating variables on the spray-dryer desulphurization performance. In particular the following variables have been investigated: inlet gas temperature, approach to adiabatic temperature, stoichiometric calcium to sulphur molar feed ratio, average droplet initial size and velocity, average suspended calcium hydroxide particle size, sulfur dioxide inlet concentration. Model results have been analyzed in the light of output variables profiles along the spray-dry column and of overall desulphurization performance. Comparison of model predictions with experimental data available in the literature has been used for model validation.

Adhikari *et al.* (2005) proposed a model to predict the surface stickiness of droplets of sugar-rich foods in spray drying conditions based on the drying kinetics and T_g of the surface layer. A prediction was made on how strongly the drop is likely to adhere to the dryer surface upon impact. Furthermore, they proposed a dimensionless time, ψ, as an indicator of the degree of spray dryability of a sticky material. The results were verified using an in situ stickiness testing device, which operated under convective drying conditions. However, they did not demonstrate how their findings would be realized in spray drying operations. This paper aims at relating the surface stickiness of droplets, subjected to a spray drying environment, to their surface layer T_g and powder recovery in spray dryers. He discussed key physical parameters, namely effective moisture diffusivity, water activity and mixture T_g, which are required for the implementation of the model. He also discussed the results obtained from modeling and compared these with the recovery of powders obtained from a pilot scale spray dryer.

2.3 Comparative study of the models

Here comparative study of the models have been done and presented in the following tables

Table-2.1 Comparative Study (T1)

Author	Aim	System	Assumptions	Conclusions
Parti, *et al.* (1974)	To develop a mathematical model for dimensioning of spray dryer.	-	1. The particle velocity is parallel to the axis of the equipment. 2. Particles are of the same size. 3. Evaporation takes place on the surface of the particles. 4. Thermal gradient inside the particle is negligible. 5. Drying medium is saturated on the surface of the particles.	The model is applicable for study of the spray drying process as well as for dimensioning of spray dryers.
Gauvin, *et al.* (1976)	To develop a design method from theoretical concepts.	30% solution of sodium nitrate.	-	This study shows that a spray dryer can be designed on a theoretical basis.
Wijilhuizen, *et al.* (1979)	Theoretical study of inactivation of phosphates during spray drying of skim milk.	Skim milk.	1. The droplets are perfectly spherical. 2. Air in the dryer is ideally mixed with moisture content. 3. Water transport inside the droplet is spherically symmetrical.	The model can be used to study spray drying process accompanying with first order chemical reaction and in principle it can be extended to higher order of reaction.
Sano, *et al.* (1982)	To study the evaporation from a single isolated, spherical droplet containing colloidal material.	Skim milk.	1. Once Equilibrium vapor pressure exceeds the pressure of the ambient air, the particle inflates and ruptures. 2. Only a single inflation rupture takes place. 3. The maximum radius does not change but void radius increases due to moisture loss. 4. The void radius remains the same while the outer radius shrinks from maximum radius.	Model is useful for single inflation rupture.

Table-2.2 Comparative Study (T2)

Author	Aim	System	Assumptions	Conclusions
Zbicinski, *et al.* (1988)	To develop a mathematical model for spray drying.	Latex suspension.	1. Air temperature and humidity are constant in the dryer cross sectional area. 2. There is no temperature gradient inside the particle. 3. Flow of air is co current and it is supplied to the dryer in the axial direction.	Mathematical model with non-uniform spraying was presented. Theoretical results revealed 10-15% error as related to experimental result.
Clement, *et al.* (1991)	To minimize the wall deposition rate in spray drying.	Skim milk.	1. Particles are assumed to be well mixed in the chamber, all experiencing same boundary condition. 2. There is no temperature gradient inside the particle. 3. The moisture content in the particle is assumed to be evenly distributed during the first period of spray drying.	The model can be used for controller design.
Langrish, *et al.* (1994)	To minimize the wall deposition rate in spray dryer.	20% Sodium chloride solution.	1. All particles stick to the wall, and any re-suspension of particles due to the bouncing or entrainment of wall deposits have been ignored. 2. First order drying kinetics is applied, giving a linear falling rate curve. 3. Negligible radial component of velocity of air. 4. The conversion of swirl momentum through the plenum chamber. 5. The length scale of turbulence was assumed to be diameter of the opening of the drying chamber.	The measured wall deposition has been predicted 16%. High swirl in the inlet air and large spray cone angle give the lowest wall deposition.

Table-2.3 Comparative Study (T3)

Author	Aim	System	Assumptions	Conclusions
Cakaloz, *et al.* (1997)	To develop a drying model for horizontal spray dryer.	α-amylase solution.	Deceleration zone from nozzle is assumed to be initially in the horizontal.	
Adhikari, *et al.* (2005)	Proposed a model to predict the surface stickiness of droplets based on drying kinetics and T_g of the surface layer.	Gelled solution	1. The drop is non-hollow sphere. 2. It shrinks uniformly with loss of water. 3. There are no temperature gradients within the drop. 4. Moisture transfer within the drop is by molecular diffusion and species convection. 5. Heat transfer to the drop is solely by convection. 6. There is no segregation of solids within the particle. 7. The drop is pseudo-binary in composition.	A safe drying regime is defined as the T_g of the surface layer was more than 10^0C above the drop temperature

Table-2.4 Comparative Study (T4)

Author	Particle velocity	Diameter of particle	Temperature profile of particle	Flow pattern
Parti, *et al.* (1974)	Parallel to the axis of chamber	Diameters of the particles are same.	Negligible	Co-current and counter current.
Gauvin, *et al.* (1976)	Three dimensional velocity	Droplet size distribution	Negligible	Co-current and counter current.
Wijihuizen, *et al.* (1979)	-	Initial sizes of the particles are same.	Negligible	-
Sano, *et al.* (1982)	-	-	Negligible	-
Zbicinski, *et al.* (1988)	Three dimensional velocity	Drop size distribution	Negligible	Co-current
Clement, *et al.* (1991)	-	-	Negligible	-
Langrish, *et al.* (1994)	Three dimensional velocity	-	Negligible	-
Perez-correa, *et al.* (1995)	-	-	Negligible	Co-current
Cakaloz, *et al.* (1997)	-	Drop size distribution	Negligible	Horizontal co-current
Adhikari, *et al.* (2005)	-	-	Negligible	-

Table-2.5 Comparative Study (T5)

Author	Application	Mode of operation	Experimental equipment size	Co-relation for mass transfer coefficient
Parti, *et al.* (1974)	The model applies for change of transport coefficient. Also applicable to design and analysis purpose.	Steady state operation	-	Filonenko's relation.
Gauvin, *et al.* (1976)	To design from theoretical concepts.	Steady state operation	1.22m ID and 0.61m upper jacket cylindrical section and 1.22m high lower conical section.	$Sh = 2.0 + \mathrm{Re}^{0.5}\, Sc^{0.33}$
Wijilhuizen, *et al* (1979)	To study first order chemical reaction during drying.	Steady state operation.	-	Sh=2.0
Sano, *et al.* (1982)	Calculation for drying of a particle with inflation and rupture.	Steady state operation.	-	$Sh(p_{Bm}/p)^{-0.2} = 0.65\,\mathrm{Re}^{0.5}\, Sc^{0.33}$
Zbicinski, *et al.* (1988)	To study the drying kinetics.	Steady state operation	-	-
Clement, *et al.* (1991)	Applicable to controller design	Steady state and dynamic operation	Diameter = 2m	Sh = 2.0
Langrish, *et al.* (1994)	This model is applicable to reduce the wall deposition.	Steady state operation.	Diameter = 0.93m Height = 1.69m	Chilton-Colburn analogy.
Perez-Correa, *et al.* (1995)	To design a robust controller	Dynamic	-	-
Cakaloz, *et al.* (1997)	To study drying kinetics in horizontal dryer.	Steady state operation.	-	-

Table-2.6 Comparative Study (T6)

Author	Application	Mode of operation	Experimental equipment size	Correlation for mass transfer coefficient
Adhikari, *et al.* (2005)	To relate the surface stickiness of droplets to their surface layer T_g and for powder recovery in spray dryers.	Steady state operation	-	$Bi(M)=\frac{k_g \times r}{3D_{eff}}$

Table-2.7 Comparative Study (T7)

Author	Correlation for heat transfer coefficient	Correlation for drag coefficient	Correlation for saturation vapor pressure
Parti, *et al.* (1974)	$Nu = 2.83 + 0.6\,\mathrm{Re}^{0.5}\,\mathrm{Pr}^{0.33}$	$C_D = \frac{24}{\mathrm{Re}}; \mathrm{Re} \le 0.1$ $C_D = \frac{21}{\mathrm{Re}} + \frac{6}{\left(\mathrm{Re}^{0.5}\,\mathrm{Pr}^{0.33}\right)} + 0.28;$ $0.1 \le \mathrm{Re} \le 4000$	$\log(p_i) = 0.622 \frac{7.5 t_{wb}}{238 + t_{wb}}$
Gauvin, *et al.* (1976)	$Nu = 2.0 + 0.6\,\mathrm{Re}^{0.5}\,\mathrm{Pr}^{0.33}$	C_D was determined by means of the equation presented by Beard and Purppacher.	-
Wijilhuizen, *et al.* (1979)	Nu = 2.0	-	-
Sano, *et al.* (1982)	$Nu = 2.0 + 0.65\,\mathrm{Re}^{0.5}\,\mathrm{Pr}^{0.33}$	-	$p_{AS} = 10^{\left[8.10765 - \{1750.286/(235 + (T - 273))\}\right]}$
Zbicinski, *et al.* (1988)	-	-	-
Clement, *et al.* (1991)	Nu = 2.0	-	-
Langrish, *et al.* (1994)	$Nu = 2.0 + 0.6\,\mathrm{Re}^{0.5}\,\mathrm{Pr}^{0.33}$	C_D has been estimated from the correlation recommended by Clif, *et al.*	-
Perez-Correa, *et al.*(1995)	-	-	-
Cakaloz, *et al.* (1997)	-	-	-
Adhikari, *et al.* (2005)	$Bi(H) = \frac{h_g \times r}{3 k_{drop}}$	-	-

2.4 CFD based models

Recent advancement in Computational Fluid Dynamics (CFD) now gives a considerable insight into the detailed design of spray drying chamber. Industrial equipment is often a large scale and processing products where there are tight constraints of quality. Recently different authors have published a number of papers. A brief description of these papers is given below.

Bahu *et al.* (1993) have presented a paper on *Computational Modelling of Spray Dryer.* They have used AEA technology body fitted FLOW 3D code, to which they have added a discrete droplet and drying model, the gas flow pattern, spray gas mixing, particle temperature and moisture histories can be determined. They have performed considerable work to assess appreciate turbulence model to test the validity of the results against experimental data.

Kieviet *et al.* (1997) have modeled the air flow pattern in a co-current plant spray dryer (diameter = 2.2m). They have modeled a spray chamber without any spray nozzle. The swirl angle was zero. The work was performed with a CFD package (FLOW 3D). The boundary conditions for the CFD models (velocity and turbulence quantities at the inlet) were derived from measurements with a hot wire prob. To validate the CFD model, air velocity magnitudes were measured at numerous locations in the spray drying chamber. To interpret the data, a novel approach was developed based on the interpretation of the velocity distribution rather than the time averaging the signals. This was necessary because of flow reversal and large fluctuations in air velocity. The measurement was compared with the CFD model results and the agreement between model and measurements was reasonable. He has introduced a device for measuring temperature and humidity pattern in a spray drying chamber. They have discussed that the device can be used to separate small particles (approximately 10 μm) from a gas stream. CFD was used in this design process. The device has been proved to be very useful in measuring temperature and humidity in a spray drying chamber.

Southwell *et al.* (1999) introduced a good paper on process intensification in spray dryers by turbulence enhancement. A pair of experiment has been performed for each set of conditions given by Zbiciniski *et al.* by varying the inlet turbulence intensity in the second while the other conditions remain same. The results of three of such pairs of experiments have been studied using the CFD code CFX4. The effect of the varying several model parameters and conditions are considered in order to investigate the sensitivity of the model for this problem. Although the simulation consistently predicted levels of cumulative evaporation was always predicted when inlet turbulence was enhanced. The simulation predicted improvements between 5% and 11% which are significantly less than 16.6% for these cases.

Straatsma *et al.* (1999) developed a drying model, named NIZO-DrySim to simulate aspects of drying processes in the food industry. DrySim simulates the gas flow in a spray dryer in two dimensions and calculates the trajectories and the course of the drying process of the atomized particles. The model makes use of computational fluid dynamics (CFD) techniques. It utilizes the k-ε turbulence model to calculate the gas flow field. The drying of droplets is influenced by both external and internal transport phenomena (from particle surface to surrounding air in the first case and diffusion within particles in the second). The differential equation that describes the diffusion process of spherical particles is solved numerically, simultaneously with the equations for external heat and mass transfer. It is shown that the drying model is an effective tool in giving indications of how to adapt industrial dryers, for example to get a better product quality or reduce fouling problems.

Langrish *et al.* (2001) have introduced a good paper explaining the spray drying characteristics of food ingredients by applying CFD package. Applications include tighter design of spray dryers and reducing operational problem, such as wall deposition. There is still considerable scope for the application of this approach, and this paper reviewed the possible future directions for application. Particular issues in the use of spray drying, food ingredients are identified and discussed, namely thermal degradation, aroma loss and particle stickiness.

Harvie *et al.* (2001) introduced a paper on numerical simulation of gas flow patterns within a Tall-Form spray dryer. The simulations were performed using CFX4.3, a fine volume based CFD package. This study represents the first application of very large eddy simulation (VLES) approach to the simulation of spray dryers. They have performed this job in order to gain a more detailed understanding of the flow patterns and their stability in their design of dryer, which is commonly used in counter current drying applications, such as the drying of detergents. Limited validation of the simulations is achieved through comparison against qualitative experimental flow pattern information. It is found that by altering the angle of the inlet air streams into the dryer, the nature of air flow within the dryer could be significantly altered. In the majority of the simulated cases, large transients developed in the flow, the nature of these transients is critically dependent on the inlet conditions. The existence of such transients would be detrimental to actual spray dryer performance. However introducing a large amount of swirl into chamber can stabilize the flow patterns.

Huang *et al.* (2006) investigated the effects of different chamber geometries, i.e., cylinder-on-cone, lantern, hour-glass and pure cone, on the drying performance and particle residence time using FLUENT code. They showed that it was possible for the designer to select other chamber geometries, and not only the traditional cylinder on-cone geometry. He also carried out a parametric study for a co-current spray dryer fitted with pressure nozzle. An ultrasonic nozzle spray dryer was studied numerically by him, as well. Few studies on spray dryers with rotary disc atomizers can be found in the literature; most of them are experimental in nature. He showed that the RNG k–ε different turbulence model is suitable for simulating the complex, swirling two-phase flow in the spray drying chamber fitted with a rotating disc when compared with three other turbulence models, i.e., standard k–ε, realizable k–ε and Reynolds stress models using FLUENT code. He discussed a set of numerical results obtained using the CFD code Fluent 6.1 for a co-current flow spray dryer fitted with either a rotary disk or a pressure nozzle. A fully three-dimensional configuration, i.e., cylinder-on-cone geometry, is considered. The RNG k–ε different turbulence model was selected in his study based on the previous work. Comparison with limited experimental data is included as well.

2.5 Motivation for the present work

It is evident from the earlier study that spray dryer is an important unit in chemical industries. Spray dryers are widely used in dairy industry, food preserving industries, dyestuff industries, pharmaceutical industries, chemical industries etc. It would have a marked effect on total economy of the industry. Therefore it is essential to understand the complex processes occurring in the spray drying unit and also to quantify them. Mathematical modeling appears to be an attractive tool for the analysis and simulation of spray dryer. Due to complex hydrodynamics, phase transfer, the spray drying also compels a chemical engineer for its modeling and simulation though it is a simple dryer.

It is obvious from the literature review that a large number of mathematical models have been developed for the spray dryer during the past century. These models include complex particle trajectories, drying kinetics which may be used to obtain its behavior under steady state or unsteady state operation. They are limited in their applications and differ from each other in terms of their complexity and the objectives of the studies carried out.

According to Denn (1986), a mathematical model may be defined as follows:

A mathematical model of a process is a system of equations whose solution, for a given input data, is representative of the response of the process to a corresponding set of inputs.

Besides the model equations, which are based upon conservation laws, constitutive properties of the modelled system play a crucial role. It is essential to determine them with a model, is the boundary conditions with which it is solved to obtain its behavior. Generally the boundary conditions are either given or easy to determine. In view of the above studies on mathematical modelling of spray dryer have been under taken to achieve the objectives, mentioned in chapter-I.

2.6 Concluding Remarks

In this chapter models of spray dryer proposed by various research workers, have been critically reviewed. Some models include complex particle trajectories and some models did not consider it. Models also vary in their complexity depending upon the objectives of the modeling. Lastly the motivation for the proposed study has been explained.

MODEL DEVELOPMENT

3.0 Introduction

This chapter describes the development of the mathematical model. We consider a vertical cylindrical co-current spray dryer in which liquid (Skim milk) and drying agent (Air) both are introduced from the top of the chamber. The major transfer processes occur in between the liquid particles (known as solid particles after attaining critical moisture content) and drying air. The basis of model development is the fundamental conservation law of mass, energy and momentum. Besides model requires constitutive relationships pertaining to the phases present in the spray drying chamber so that the solution of model equations may be obtained.

3.1 Premises of Spray Dryer

Generally spray dryers are vertical cylinders in which liquid or slurry or paste is sprayed from top and hot air is blown from top or bottom. As we have considered co-current operation, the air is also introduced from top of the dryer. Figure 3.1 schematically represents the control volume of spray dryer.

3.2 Assumptions

In order to develop mathematical model following simplifying assumptions are taken.

1. The evaporation takes place only on the surface of the particle. That is the evaporative surface does not recede into the particle.
2. The drying medium is saturated on the surface of the particle.
3. The sizes of the drops to be dried in spray dryer are generally so small, that the thermal gradient inside the drops can be neglected. So it can be inferred that the interfacial surface temperature is equal to the average temperature of the particle.
4. The size of the particle changes similar to shrinkage of balloon until the moisture content becomes equal to the critical moisture content i.e., the

volume reduction of the particle is equal to the moisture leaving the particle.

5. At critical moisture content the surface of the particle solidifies and diameter remains constant.
6. The velocity of the air is parallel to the vertical axis of the dryer.
7. No chemical degradation occurs due to heating.

3.3 Choice of Control Volume

For developing the model for overall mass transfer and overall heat transfer, we divided the dryer chamber into small elemental heights. Each of these consists of small volume of bed between two horizontal planes perpendicular to the vertical axis of the drying chamber. The temperature, humidity, moisture content of all particles and other physical properties for all phases assumed to be uniform inside the elemental volume. Thus mass and heat transfer rate at a given time and height, are the same throughout the elemental volume. This elemental volume is control volume. This is mentioned in figure. All the incoming and outgoing streams around control volume with properties are clearly mentioned in the figures.

3.4 Mass Balance Equations

Input + Disappearance due to reaction = Output + Accumulation

Here disappearance = 0 (No chemical reaction)

Input = 0 (No moisture input)

Therefore,

Output = -Accumulation

$$-G_s \frac{dX}{d\tau} = Na \tag{3.1}$$

$$\frac{dX}{d\tau} = -\frac{Na}{G_s}$$

$$\frac{dX}{d\tau} = -\frac{Na\beta}{G_s} \tag{3.2}$$

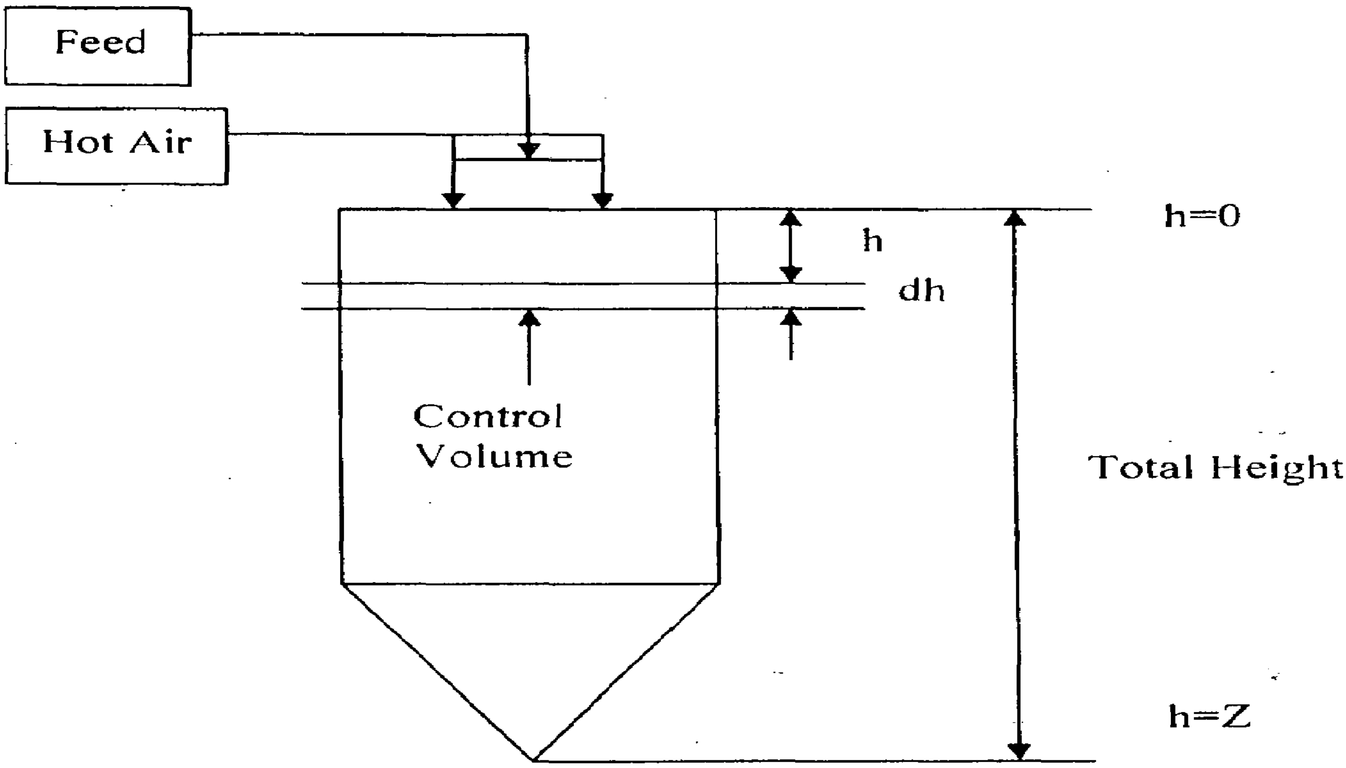


Fig. 3.1 Schematic Diagram of Spray Dryer Modelling

3.4.1 Around a single particle:

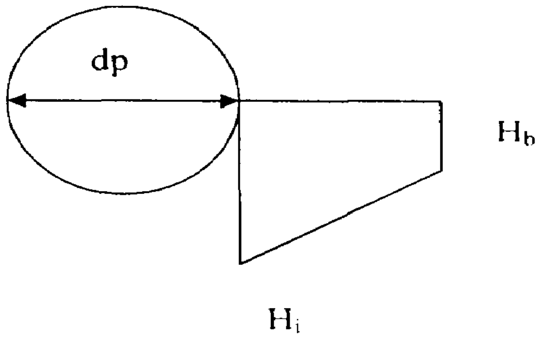


Fig. 3.2 Schematic Diagram of a Single Particle for Mass Transfer.

Where ,β is a derived function, which takes into account the falling rate of drying.

$$\beta = \phi(X)$$

$$\beta = 1 \text{ for } X \geq X_c$$

$$\beta = -0.183X^2 + 1.4052X + 0.0016 \text{ for } X < X_c$$

Now on considering the definition of the density of the wet particle,

$$\rho_{sw} = \frac{G_s + G_m}{V_{sw}} \tag{3.3}$$

$$X = \frac{G_m}{G_s} \tag{3.4}$$

$$G_s = \frac{V_{sw}\rho_{sw}}{1 + X} \tag{3.5}$$

$$V_{sw} = \frac{\pi d_p^3}{6} \tag{3.6}$$

$$a = \pi d_p^2 \tag{3.7}$$

$$\frac{dX}{d\tau} = -\frac{6N\beta(1 + X)}{\rho_{sw} d_p} \tag{3.8}$$

3.4.2 Around the control volume:

$$\xi G_s X + LH_b = \xi G_s (X + dX) + L(H_b + dH_b) \tag{3.9}$$

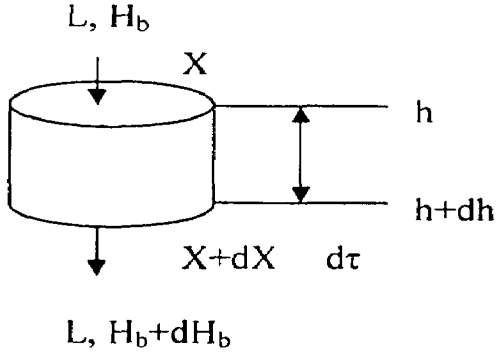


Fig. 3.3 Schematic Diagram of Control Volume for Mass Transfer

$$\xi G_s dX + L dH_b = 0$$

$$\xi G_s \frac{dX}{d\tau} = -L\frac{dH_b}{d\tau}$$

$$\xi G_s\left(-\frac{6N\beta(1+X)}{\rho_{sw}d_p}\right) = -L\frac{dH_b}{d\tau} \tag{3.10}$$

After further simplifications we get,

$$\frac{dH_b}{d\tau} = \frac{\xi\pi d_p^2 N\beta}{L} \tag{3.11}$$

3.5 Energy Balance Equations

3.5.1 Around a single particle:

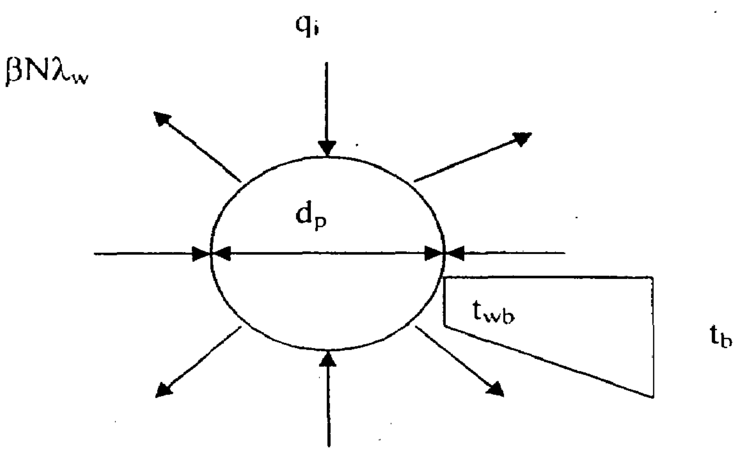


Fig. 3.4 Schematic Diagram of a Single Particle for Heat Transfer

$$C_{sw}V_{sw}\rho_{sw}t_{wb} = \left[C_{sw}V_{sw}\rho_{sw} + d(C_{sw}V_{sw}\rho_{sw})\right](t_{wb} + dt_{wb}) + (N\beta\lambda_{wb} - q_i)a d\tau$$

After simplification we get,

$$C_{sw}V_{sw}\rho_{sw}dt_{wb} = (q_i - \beta N\lambda_{wb})ad\tau - t_{wb}d(C_{sw}V_{sw}\rho_{sw}) \tag{3.12}$$

Now,

$$d(C_{sw}V_{sw}\rho_{sw}) = C_{sw}d(V_{sw}\rho_{sw}) + V_{sw}\rho_{sw}dC_{sw}$$
$$d(V_{sw}\rho_{sw}) = d(G_m + G_s) = dG_m = G_s dX$$
$$dC_{sw} = d(C_s + XC_m) = C_m dX$$
$$V_{sw}\rho_{sw}dC_{sw} = G_s(1+X)C_m dX$$

On substituting the above values into equation (3.12), we get

$$C_{sw}V_{sw}\rho_{sw}dt_{wb} = (q_i - \beta N\lambda_{wb})ad\tau - t_{wb}\left[C_{sw}G_s dX + G_s(1+X)C_m dX\right] \tag{3.13}$$

After further simplifications we get,

$$\frac{dt_{wb}}{d\tau} = \frac{(q_i - \beta N\lambda_{wb})a}{C_{sw}\rho_{sw}V_{sw}} + \frac{6t_{wb}\beta N}{\rho_{sw}d_p}\left[1 + (1+X)\frac{C_m}{C_{sw}}\right] \tag{3.14}$$

We can get the heat flux on surface from the heat balance written for the boundary layer surrounding the particle as shown in the figure.

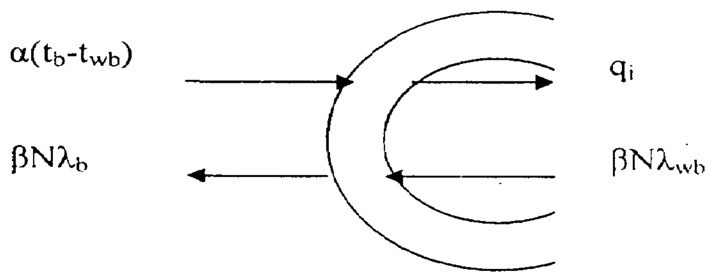


Fig. 3.5 Schematic Diagram for Heat Transfer through film

$$h_c(t_b - t_{wb}) + \beta\lambda_{wb}N = q_i + \beta N\lambda_b$$
$$\lambda_b = \lambda_0 + c_{PwG}t_b \tag{3.15}$$
$$\lambda_{wb} = \lambda_0 + c_{PwG}t_{wb}$$
$$q_i - \beta N\lambda_{wb} = h_c(t_b - t_{wb}) - \beta N\lambda_b$$
$$\frac{dt_{wb}}{d\tau} = \frac{6\left[h_c(t_b - t_{wb}) - \beta N(\lambda_0 + c_{PwG}t_b)\right]}{C_{sw}\rho_{sw}d_p} + \frac{6t_{wb}\beta N}{\rho_{sw}d_p}\left[1 + (1+X)\frac{C_m}{C_{sw}}\right]$$

3.5.2 Around the control volume:

$$L\lambda_b + \beta N\lambda_b \xi a d\tau - L(\lambda_b + d\lambda_b) - h_c(t_b - t_{wb})\xi a d\tau = 0 \tag{3.16}$$

Enthalpy of moist drying medium is given by,

$$\lambda_L = c_{L,w} t_L + \lambda_0 H_L$$

$$\beta N\lambda_b \xi a d\tau = L d\lambda_b + h_c(t_b - t_{wb})\xi a d\tau$$

$$d\lambda_b = c_{L,w} dt_b + (\lambda_0 + c_{PwG} t_b) dH_b$$

$$\beta N(\lambda_0 + c_{PwG} t_b)\xi a d\tau = L c_{L,w} dt_b + L(\lambda_0 c_{PwG} t_b) dH_b + h_c(t_b - t_{wb})\xi a d\tau$$

After further simplifications we get,

$$\frac{dt_b}{d\tau} = -\frac{h_c(t_b - t_{wb})\xi a}{L c_{L,w}} \tag{3.17}$$

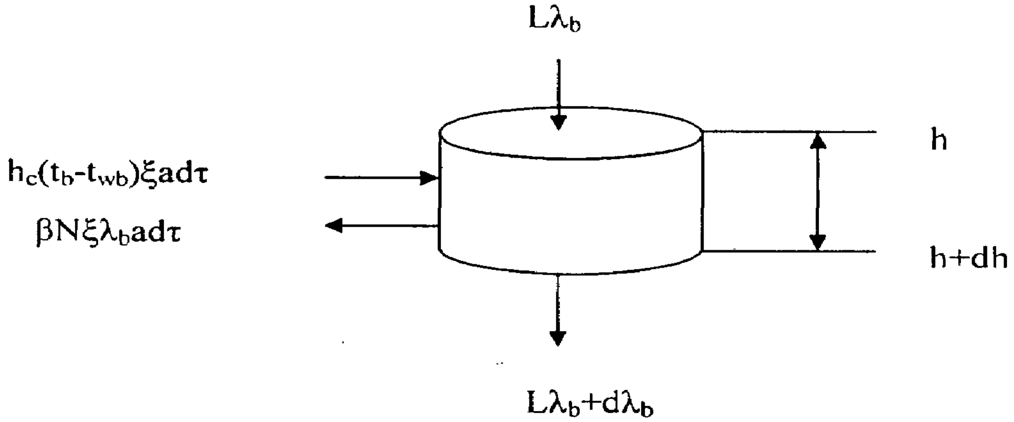


Fig 3.6 Schematic Diagram of Control Volume for Heat Transfer

3.5 Momentum Balance Equations

Momentum balance is taken for particle to formulate the complex particle trajectories. Here we have assumed that the air velocity is parallel to the vertical axis of the spray drying chamber. Hence here we do not take balance for air velocity profile. We have considered the velocity of the particle only in the axial direction. So velocity profile is considered only in the axial direction. Here we have applied the Newton's second law of motion on a single particle.

Forces acting on the particle

1. Drag force in the direction opposite to particle velocity.
2. Buoyancy force in the upward direction.
3. Gravitational force in the downward direction.

According to Newton's second law of motion:

Rate of change = Gravitational – Buoyancy force – Drag force

of momentum force

$$\frac{\pi d_p^3}{6}\rho_{Sw}\frac{dv_p}{dh}=\frac{\pi d_p^3}{6}\rho_{Sw}g-\frac{\pi d_p^3}{6}\rho_{Lw}g-C_D\frac{\pi d_p^2}{4}\left(v_L-v_p\right)^2\rho_{Lw} \tag{3.18}$$

After simplification we get,

$$\frac{dv_p}{dh}=g\frac{\rho_{sw}-\rho_{Lw}}{\rho_{sw}v_p}-\frac{3}{2}\frac{C_D\rho_{Lw}}{d_p\rho_{sw}v_p}\left(v_L-v_p\right)^2 \tag{3.19}$$

Converting this equation in terms of differential time we get

$$\frac{dv_p}{d\tau}=g\frac{\rho_{sw}-\rho_{Lw}}{\rho_{sw}}-\frac{3}{2}\frac{C_D\rho_{Lw}}{d_p\rho_{sw}}\left(v_L-v_p\right)^2 \tag{3.20}$$

3.6 Concluding Remarks

In this detailed derivation, model equations for co-current spray dryer are presented. To derive this model we take some simplifying assumptions. These assumptions are also mentioned. Model equations are derived from the fundamental laws of mass, momentum and energy. We have derived all equations in terms of residence times. These equations form a set of non-linear coupled ordinary differential equations, which form an initial value problem (IVP). These can be solved by MATLAB solver ode45.

CHAPTER-IV

CONSTITUTIVE RELATIONSHIPS

4.0 Introduction

The mathematical model for spray dryer has already been developed on the basis of conservation laws of mass, energy and momentum. In order to predict the performance of the spray dryer, this model should be solved in conjunction with the constitutive relationships pertaining to the phases and processes occurring in the spray dryer. Constitutive relationships by definition are the relationships for those properties, which constitutes the system. These properties are called constitutive properties and are, therefore, specific to the system. Constitutive properties include physical, kinetic and transport properties. The correlations used for constitutive properties of the system are described in brief with their limitations in this chapter. The correlations which are available in the literature, are taken from there, other correlations are derived.

4.1 Correlation for mass flux

Mass flux can be calculated from the following equation:

$$N = \psi(H_i - H_b) \tag{4.1}$$

4.1.1 Correlation for Evaporative Coefficient

Evaporative coefficient includes the mass transfer resistance. Mass transfer resistance is function of Reynold's number. Evaporative coefficient can be calculated from Filonenko's [22] correlation, which is represented as follows:

$$\psi = \phi(h_c, c_{Lw}, Le) \tag{4.2}$$

$$\psi = \frac{h_c}{c_{Lw} Le} \tag{4.3}$$

4.2 Correlation for Density

Density of particles varies as drying starts. The nature of variation of density changes when critical moisture has been achieved. The relationship for density of particle can be derived as follows [22]

Change in moisture content is equal to the change in mass of the particle. So we can write

$$dG = d(V_{sw}\rho_{sw}); \text{ for } X \geq X_c \tag{4.4}$$

After further simplification we get the following equation:

$$\frac{d\rho_{sw}}{d\tau} = -\frac{6\beta N\rho_{sw}\left(\frac{1}{\rho_{sw}} - \frac{1}{\rho_m}\right)}{d_p} \tag{4.5}$$

The analytical solution of the above equation results

$$\rho_{sw} = \frac{\rho_s(1+X)}{1+X\left(\frac{\rho_s}{\rho_m}\right)} \tag{4.6}$$

For solidified surface the equation becomes

$$\frac{d\rho_{sw}}{d\tau} = -\frac{6\beta N}{d_p}; \text{ for } X < X_c \tag{4.7}$$

4.3 Correlation for Diameter

Diameter of the particle also reduces as drying starts. Reduction of diameter will cease when the particle surface becomes solid. Since we have assumed that after attaining critical moisture content particle surface become solid. Therefore after attaining critical moisture content the diameter should remain constant.

Change in volume of the particle will be equal to the volume of moisture transferred from the particle to the drying agent as we assume that particle shrinks as balloon [22]. Therefore we can write,

$$dG_{sw} = \rho_m dV_{sw} = d(V_{sw}\rho_{sw}); \quad \text{for } X \geq X_c \tag{4.8}$$

After further simplification we have the following equation

$$\frac{d(d_p)}{d\tau} = -\frac{2\beta N}{\rho_m} \tag{4.9}$$

Analytical solution of the above equation gives,

$$d_p = d_{p0}\left(\frac{\rho_{sw0} - \rho_m}{\rho_{sw} - \rho_m}\right)^{1/3} \tag{4.10}$$

For solidified surface

$$\frac{d(d_p)}{d\tau}=0 \quad \text{for } X < X_c \tag{4.11}$$

4.4 Correlation for Porosity

After attaining critical moisture content, the particle starts to become porous with further drying. So after attaining critical moisture content the porosity of the particles is proportional to the amount of moisture transferred from the particle. Before that the porosity of the particle should be zero [22]. The equation for porosity can be derived as follows:

For $X \geq X_c$

$$\varepsilon = 0 \text{ and } \frac{d\varepsilon}{d\tau}=0 \tag{4.12}$$

For $X<X_c$

Porosity is proportional to the loss of moisture from the particle

$$\varepsilon_2=\varepsilon_1+\frac{\beta Na}{\rho_m V_{sw}}\Delta\tau \tag{4.13}$$

$$\frac{d\varepsilon}{d\tau}=\frac{6\beta N}{\rho_m d_p} \tag{4.14}$$

4.5 Correlation for reduction in Mass Transfer rate in the falling rate zone

Total drying of a substance is divided into two zones, constant rate zone and falling rate zone. At the constant rate zone we can use equation (4.1) for mass flux calculation. But problem arises when we go for calculating mass transfer rate at falling rate zone. We could not get such correlation. Taking experimental data from literature [ ] we have derived a falling rate function, which is polynomial in nature. The function calculates the decrease in mass transfer rate at the falling rate zone for spray drying of milk. This function can be represented as follows:

$$\beta=\phi(X) \tag{4.15}$$

$$\beta = 1 \text{ for } X \geq X_c \tag{4.16}$$

$$\beta = -0.183X^2+1.4052X+0.0016 \quad \text{for } X < X_c \tag{4.17}$$

We have developed this function from following data [17]

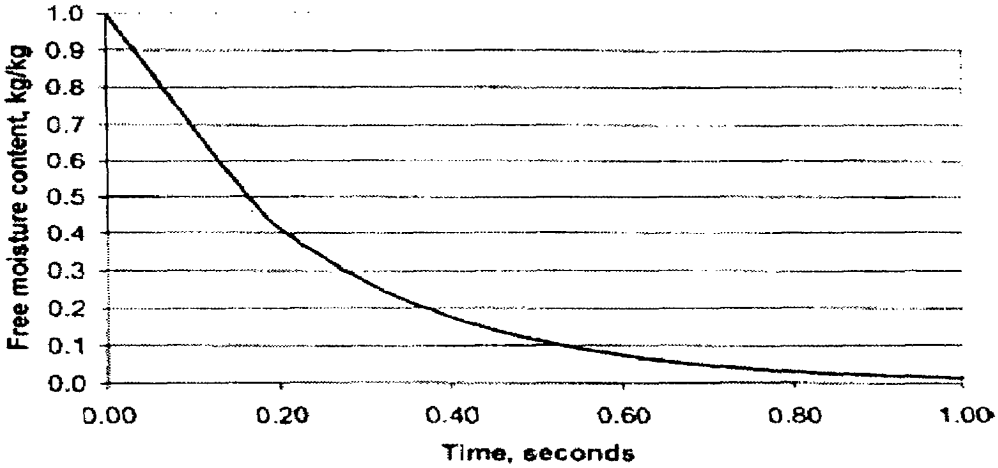


Fig. 4.1 Drying Curve for Milk

4.6 Correlation for Heat Transfer Coefficient

Resistance to heat transfer can be measured in terms of heat transfer coefficient. The heat transfer coefficient [22] is calculated by the following correlation:

$$Nu = 2.83 + 0.6\,\mathrm{Re}^{0.5}\,\mathrm{Pr}^{0.33} \qquad (4.18)$$

4.7 Correlation for Drag Coefficient

Drag coefficient is a function of Reynolds number. To calculate drag coefficient following correlation [21] is used:

$$C_D = 0.44, \text{ when } \mathrm{Re} > 4000 \qquad (4.19)$$

$$C_D = \frac{18}{\mathrm{Re}^{0.6}}, \text{ when } 0.2 < \mathrm{Re} < 4000 \qquad (4.20)$$

$$C_D = \frac{24}{\mathrm{Re}}, \text{ when } \mathrm{Re} < 0.2 \qquad (4.21)$$

4.8 Correlation for Viscosity of Air

Viscosity is a function of temperature and pressure. For our system pressure difference is negligible inside the dryer but temperature is varied from region to region. In this study we take the viscosity as a function of temperature [21] as follows

$$\frac{\mu_T}{\mu_0} = \left(\frac{T}{273}\right)^{0.65} \qquad (4.22)$$

For air $\mu_0 = 1.71*10^{-5}$ Ns/m^2

And so viscosity becomes

$$\mu_T = 1.71\times10^{-5}\left(\frac{t_b + 273}{273}\right)^{0.65} \text{ Ns/m}^2 \qquad (4.23)$$

4.9 Correlation for Thermal Conductivity of Air

Thermal conductivity is a function of temperature. From literature we have found the following correlation [21] for thermal conductivity as a function of temperature:

$$k = a + bT \qquad (4.24)$$

For air,

$$a = 2.2965\times10^{-3}\frac{W}{m^oC} \quad \text{and} \quad b = 7.95\times10^{-5}\frac{W}{m^oC^2} \qquad (4.25)$$

4.10 Correlation for Saturation Vapor Pressure

It has been assumed that the surface of the particle is saturated with drying medium, so water vapor exerts saturated vapor pressure on the surface of the particles. Saturation vapor pressure is the function of temperature. So the particle surface temperature should be equal to the adiabatic saturation temperature of the drying medium. For air water system the adiabatic saturation temperature is equal to the wet bulb temperature of the drying medium. The correlation for saturation vapor pressure is available in literature. We have used the following correlation [22] for saturation vapor pressure calculation.

$$\log p_i = 0.622 + \frac{7.5t_{wb}}{238 + t_{wb}} \qquad (4.26)$$

4.11 Correlation for Humidity

Humidity by definition is the Kg of moisture present in the air per kg of dry air. So, humidity of air is a function of partial pressure of the moisture present in the air. Correlation for this is also available in the open literature. We have used the following correlation [ ] for humidity calculation:

$$H_i = 0.622\frac{p_i}{760 - p_i} \qquad (4.27)$$

4.12 Concluding Remarks

In this chapter we have discussed about the various constitutive relationships and we also discussed their relevancy for our system. Some of the relationships are derived from fundamental laws and some are collected from literature. These relationships are used for solution of the model equations to predict the performance of the dryer.

SOLUTION TECHNIQUE

5.0 Introduction

To predict the performance of the model, solution of model equations is very essential. We found that the model is a set of nonlinear, ordinary coupled differential equations. These equations constitute the initial value problem. So these are solved in MATLAB 7.0 by using ordinary differential equations tool box. In this chapter we discussed the various initial conditions.

5.1 Initial Conditions

Initial values of different operating parameters are given in the following table:

Table-5.1 Initial condition

Parameters	*Initial Values*
X (Kg/Kg)	1.0
H_b (Kg/Kg)	0.012
t_{wb}(^{0}C)	20
t_b (^{0}C)	80
u_{ax} (m/s)	1.5
u (m/s)	1.0
d_p (m)	0.8×10^{-4}
ρ_{sw} (Kg/m^3)	1580
ε	0
L (Kg/s)	0.16

Table-5.2 Feed specification

Type of feed	*Initial moisture content (Kg/Kg)*	*Critical moisture content (Kg/Kg)*	*Final moisture content (Kg/Kg)*
Skim milk	1.0	0.40	0.06

Table-5.3 Other Physical Properties

Parameter	*Notation*	*Units*	*Value*
Specific heat of moisture	c_m	J/(Kg-^{0}C)	4185.8
Specific heat of milk solid	c_s	J/(Kg-^{0}C)	2670
Specific heat of air	c_{Lw}	J/(Kg-^{0}C)	1005
Specific heat of Skim milk	c_{sw}	J/(Kg-^{0}C)	1256
Prandtl Number	Pr		0.77
Lewis Number	Le		1.0
Acceleration due to gravity	g	m/s^2	9.81
Density of water	ρ_m	Kg/m^3	1000
Density of air	ρ_{Lw}	Kg/m^3	1
Latent heat of vaporization	λ_o	J/Kg	2498923

5.2 Concluding Remarks

In this chapter initial conditions and other physical properties are given in tabulated form. The model equations are solved in MATLAB 7.0. And the graphs obtained are listed at the end.

CHAPTER-VI

RESULTS AND DISCUSSION

6.0 Introduction

In this chapter results of the proposed model have been presented and discussed. Our proposed model predicts the values of different parameters as a function of residence time and also the effect of different inlet conditions on these parameters. The numerically calculated values of all the studied parameters are represented graphically.

6.1 Validation of Model

The model developed by Parti *et al.* (1974) has been validated by considering the experimental inlet conditions mentioned in Langrish *et al.* (2001). He considered only falling rate period, but in this both constant and falling rates has been considered. The results are in good agreement with the experimental results and these are shown graphically. The residence time value which has been obtained here is with in the experimental range.

6.2 Moisture Profiles

6.2.1 Moisture Profile of Particles

Figs. (6.1 to 6.5) represent the drying kinetics of a single particle of the proposed system. Here we have developed a falling rate function, which is a polynomial function of particle moisture content. This function accounts the decrease in drying rate after attaining critical moisture content. Fig. 6.1 shows a clear decrease (decrease in slope) in drying rate after attaining critical moisture content of 0.40 [17]. Feed enters the dryer at a temperature equal to the wet bulb temperature of the drying air. Therefore, there is no reverse transport zone. Fig. 6.2 shows the particle moisture profile for different drop sizes. We have taken 80 to 100μm variation in diameter of drops. It is evident from the figure that the drying rate decreases with increase in drop size. This study tells about the importance of drop size distribution. Fig. 6.3 shows the effect of initial humidity of drying air on particle moisture profile. It shows that as humidity increases the drying rate decreases. This is due to decrease in mass transfer driving force. Fig. 6.4 shows particle moisture profile for different inlet temperatures of drying air. High initial temperature fastens the drying rate of drops. High inlet temperature of the dry air also reduces the reverse transport zone. Fig. 6.5 represents the variation moisture profile of a single particle for different liquid feed temperatures. From the curves it is evident that with increasing temperature the drying rate increases.

6.2.2 Moisture profile of drying air

Humidity of air increases with residence time. Figs. (6.6 to 6.10) represents how the humidity of drying medium varies with residence for different inlet parameters. Fig. 6.6 shows that the rate of increase of humidity in the constant rate period is more than the increase of humidity in the falling rate period. Fig. 6.7 shows variation of humidity for different inlet drying air temperatures. From this it is evident that as air inlet temperature increases, humidity increases. Fig. 6.8 shows the variation of humidity with feed temperatures. It shows that the humidity increases with increase in feed temperature. Fig. 6.9 shows the variation of humidity for different inlet humidity's of drying air. With increase in humidity the drying rate decreases. Fig. 6.10 shows the variation of humidity for different drop sizes. As the atomizer does not produce uniform sized particles, the actual humidity should be the average of the above curves.

6.3 Temperature Profiles

6.3.1 Temperature Profile of Particle

Figs. 6.11 to 6.16 show the particle temperature profile for different inlet conditions. Fig. 6.11 and 6.12 shows that initially particle temperature increases rapidly to wet bulb temperature. As we assume that drying medium is saturated on the surface of the particle, temperature of particle becomes constant at wet bulb temperature. After attaining critical moisture content there is a sudden increase of temperature. Attainment of critical moisture implies the appearance of dry spot on the surface of the particle and case hardening has occurred. We assume that after attaining critical moisture content there is no change in diameter. After that particles start to become porous. Fig. 6.13 shows temperature profile for different drop sizes. This shows that small sized particles reside for less time during drying. Fig. 6.14 shows temperature profile for different drying medium humidity's. Increase in inlet humidity of the drying air leads to increase in adiabatic saturation temperature and slows down the drying process as driving force for mass transfer decreases. Fig. 6.15 shows the temperature profile for different air inlet temperatures. Higher the inlet air temperature lower will be the drying time, which is due to increase in drying rate. Fig. 6.16 shows the particle temperature for different feed temperatures. As we assume that the drying medium is saturated on the surface of the particle, so the particle temperature should be equal to the adiabatic saturation temperature. Hence there is a decrease in temperature of particle from feed temperature to adiabatic saturation temperature when feed temperature is greater than the adiabatic saturation temperature and vice versa. Thus when the feed temperature is greater than the adiabatic saturation temperature, the drying will be very fast.

6.3.2 Temperature Profile of the Drying Air

Figs. (6.17 to 6.21) show the temperature history of drying medium as a function of drying time. The temperature of the drying air decreases as drying time increases. The rate of decrease of air temperature reduces as air travels from constant rate to falling rate period. Fig. 6.18 shows the temperature profile for different sized particles. The actual temperature profile of the drying air should be the average of the profiles of fig. 6.18 because particles of different sizes move in the dryer. Fig. 6.19 shows the temperature profile for different drying air initial humidity's. From the figure it is evident that the effect of humidity on temperature profile is not appreciable. Fig. 6.20 shows the temperature profile for different initial hot air temperatures. From this figure it is evident that the rate of decrease of temperature increases with increasing initial temperature. Fig. 6.21 shows the temperature profile for different feed temperatures. The change in temperature of air does not vary much for different feed temperatures.

6.4 Velocity Profile of the Particle

Figs. 6.22 to 6.23 represent the velocity profile of the particle. Here we have considered only one dimensional velocity. Fig. 6.22 shows the axial velocity profile of the particle. The velocity changes rapidly in the constant rate period and remains constant in the falling rate period. Fig. 6.23 shows the variation of velocity for different drop sizes. With increase in drop size the velocity increases.

6.5 Density, Porosity and Diameter Profiles

6.5.1 Density Profile

Figs. 6.24 to 6.28 show the density profiles. Initially the density of particle starts to increase as we have assumed that change of volume is proportional to the change of moisture content (shrinking balloon). After attaining critical moisture content the density of particle starts to decrease as volume does not change but moisture being transported from particle to drying air.

6.5.2 Diameter Profile

The diameter pattern of the particles is represented in the figs. 6.28 to 6.33. Initially the diameter of particle reduces proportionally with the volume of moisture transported. After attaining the critical moisture content diameter of particle remains constant. Figs. 6.29 to 6.33 show the effects of different parameters on the change of diameter with the drying time.

6.5.3 Porosity Profile

Figs. 6.34 to 6.38 show the porosity profiles. The result is in accordance with our assumption. Up to critical moisture content the porosity remains zero. After attaining critical moisture content the particle starts to become porous. The porosity increases with a decreasing rate.

6.6 Concluding Remarks

In this chapter all results regarding the study of different parameters are represented graphically. In order to solve mathematical model equations, the experimental input data and characteristics of various operating parameters are collected from the literature. The model is solved with these experimental input data. All the computationally estimated results are presented and discussed in this chapter. It is found that all the results are in good agreement with the experimental results. In view of this fact, the developed mathematical model for co-current spray dryer has been validated.

Fig 6.1 Particle Moisture Profile

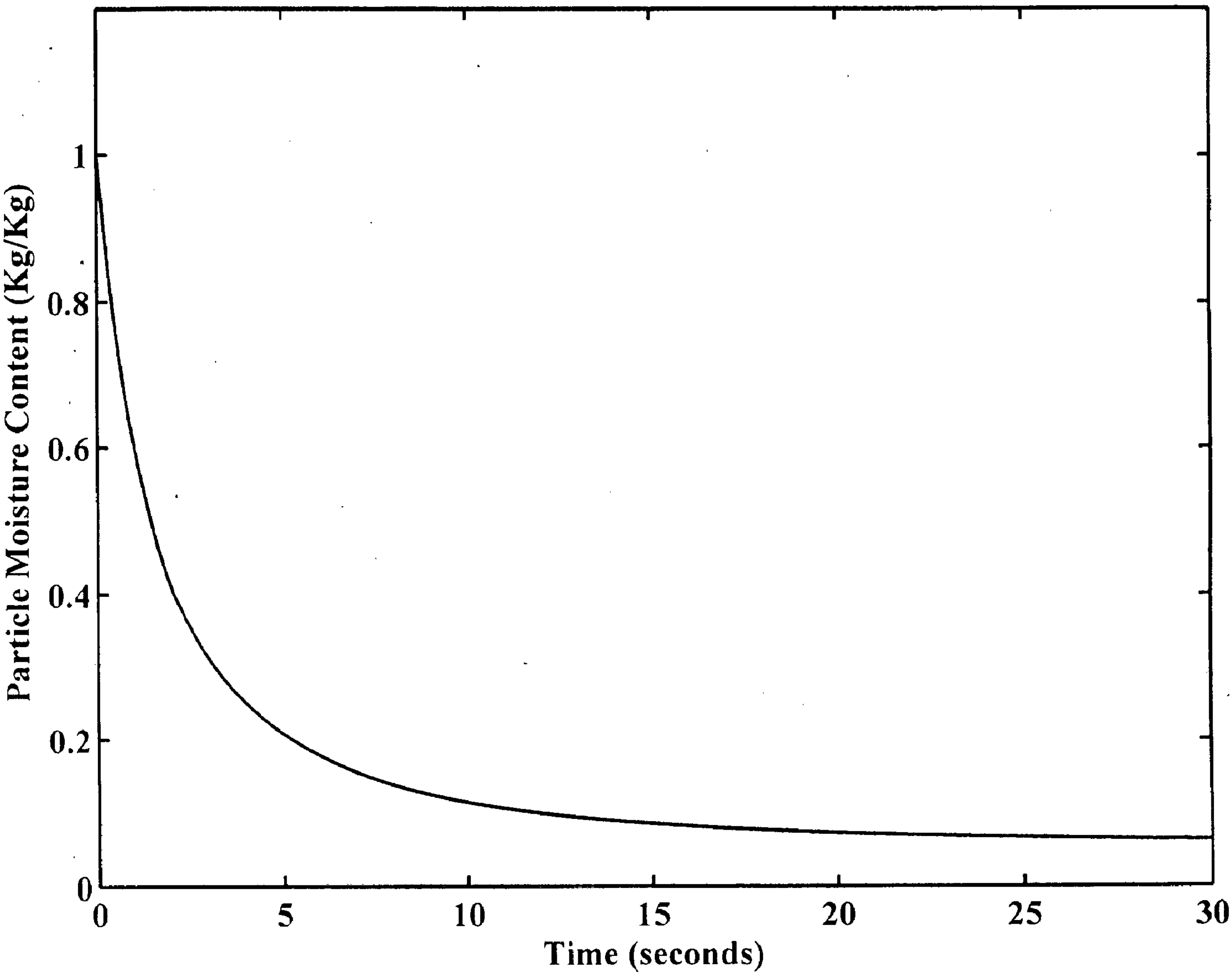


Fig 6.2 Particle Moisture Profile for Different Drop Sizes

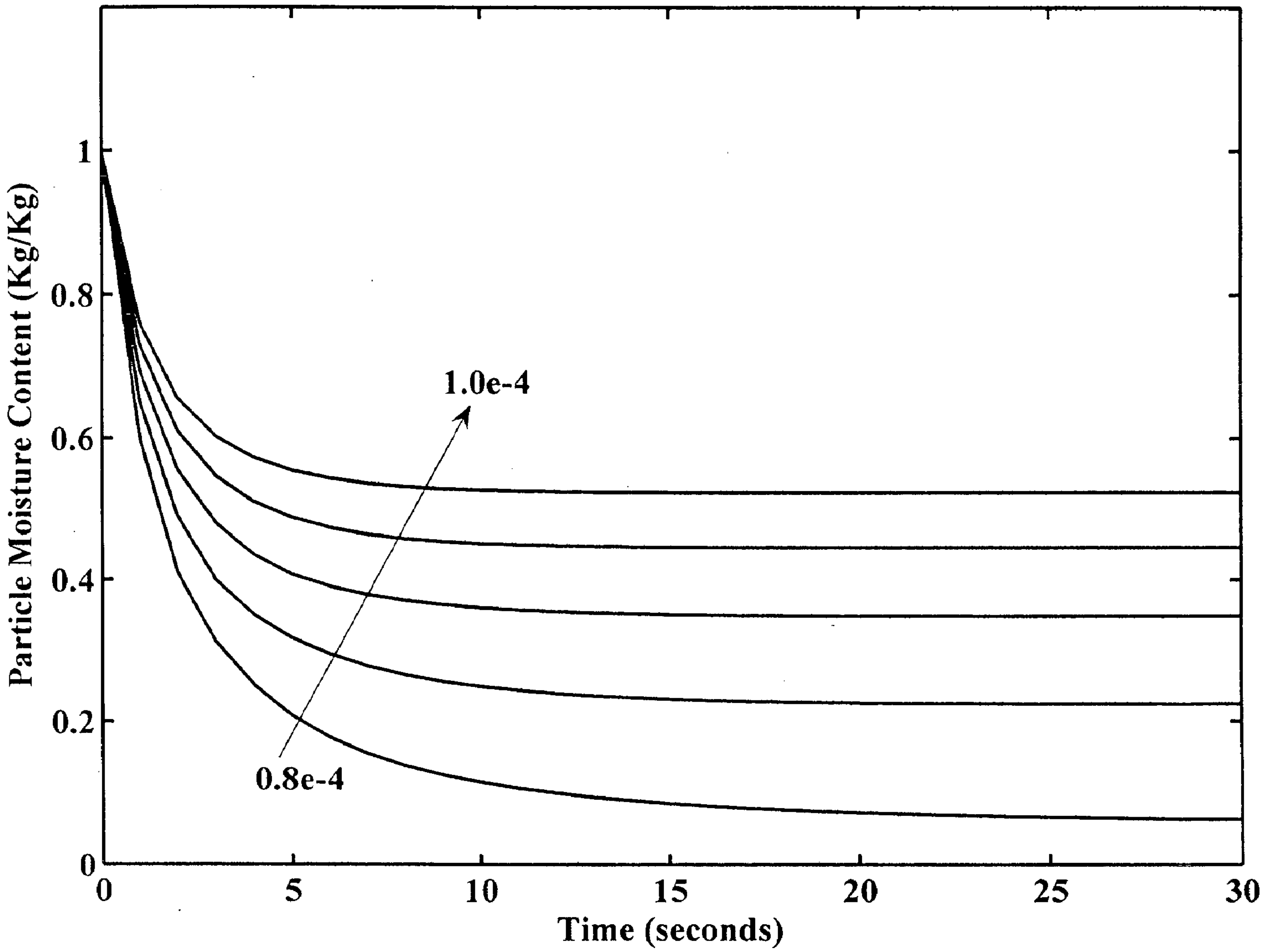


Fig 6.3 Particle Moisture Profile for Different Inlet Humidities of Drying Medium

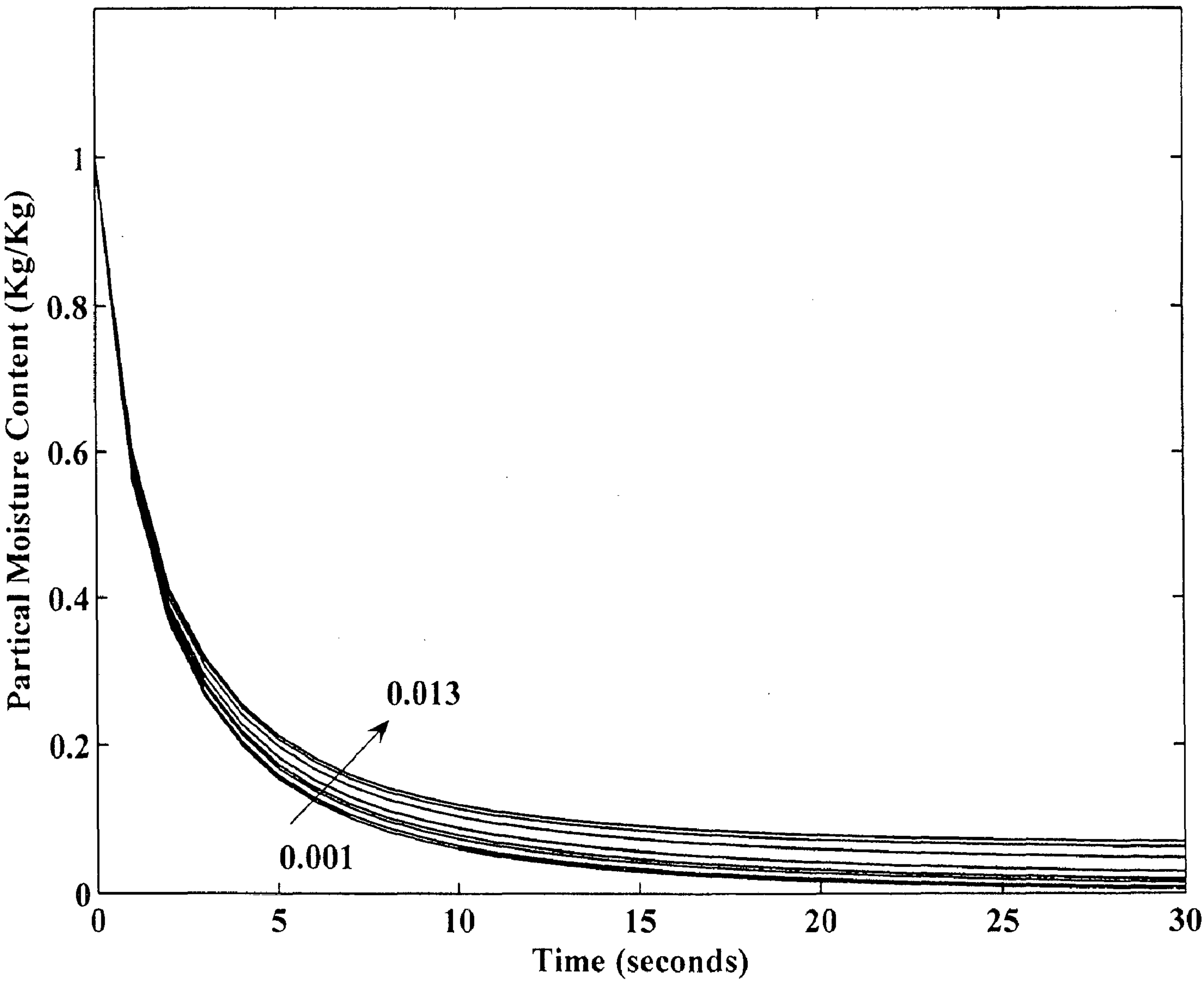


Fig 6.4 Particle Moisture Profile for Different Drying Medium Inlet Temperatures

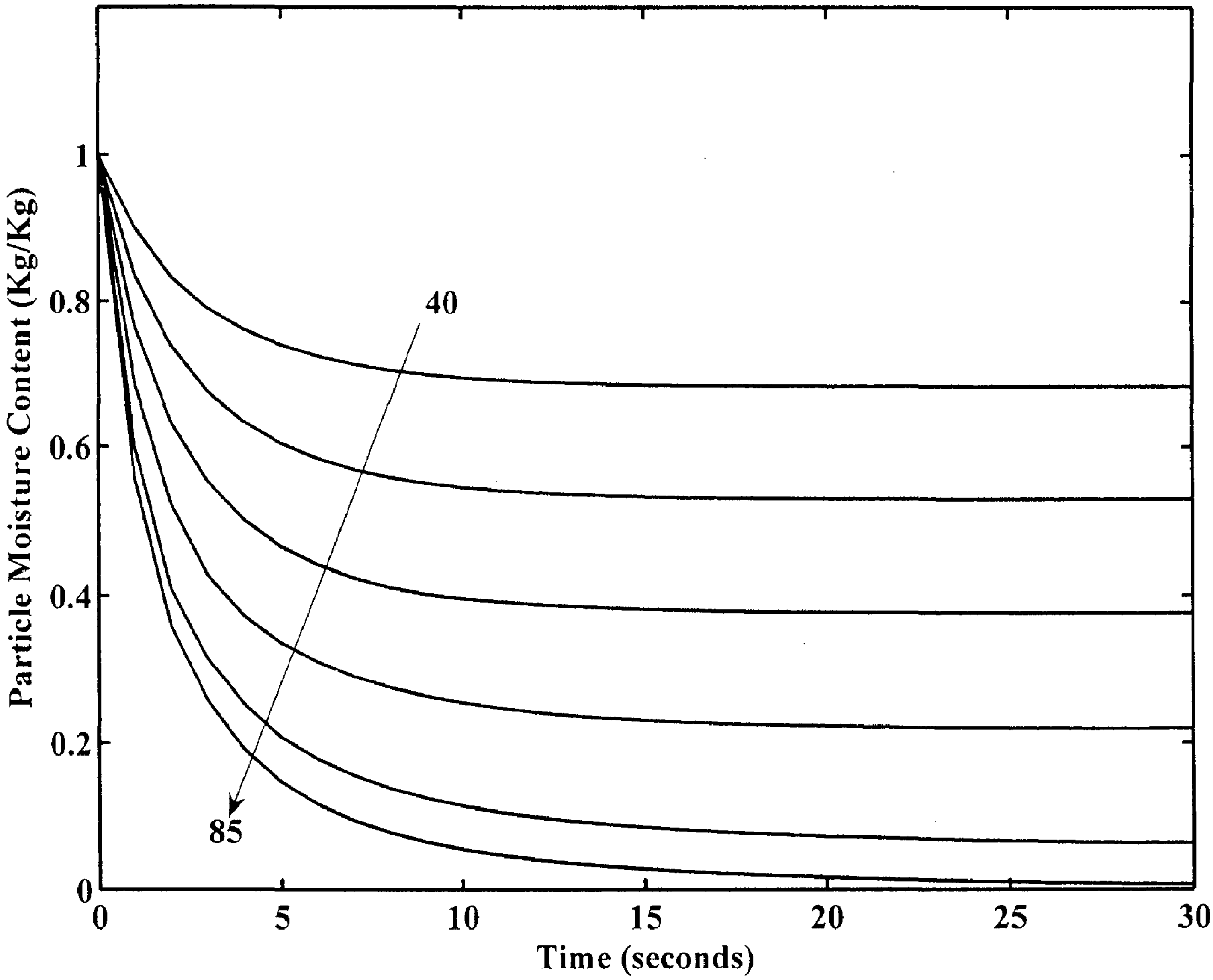


Fig 6.5 Particle Moisture Profile for Different Feed Temperatures

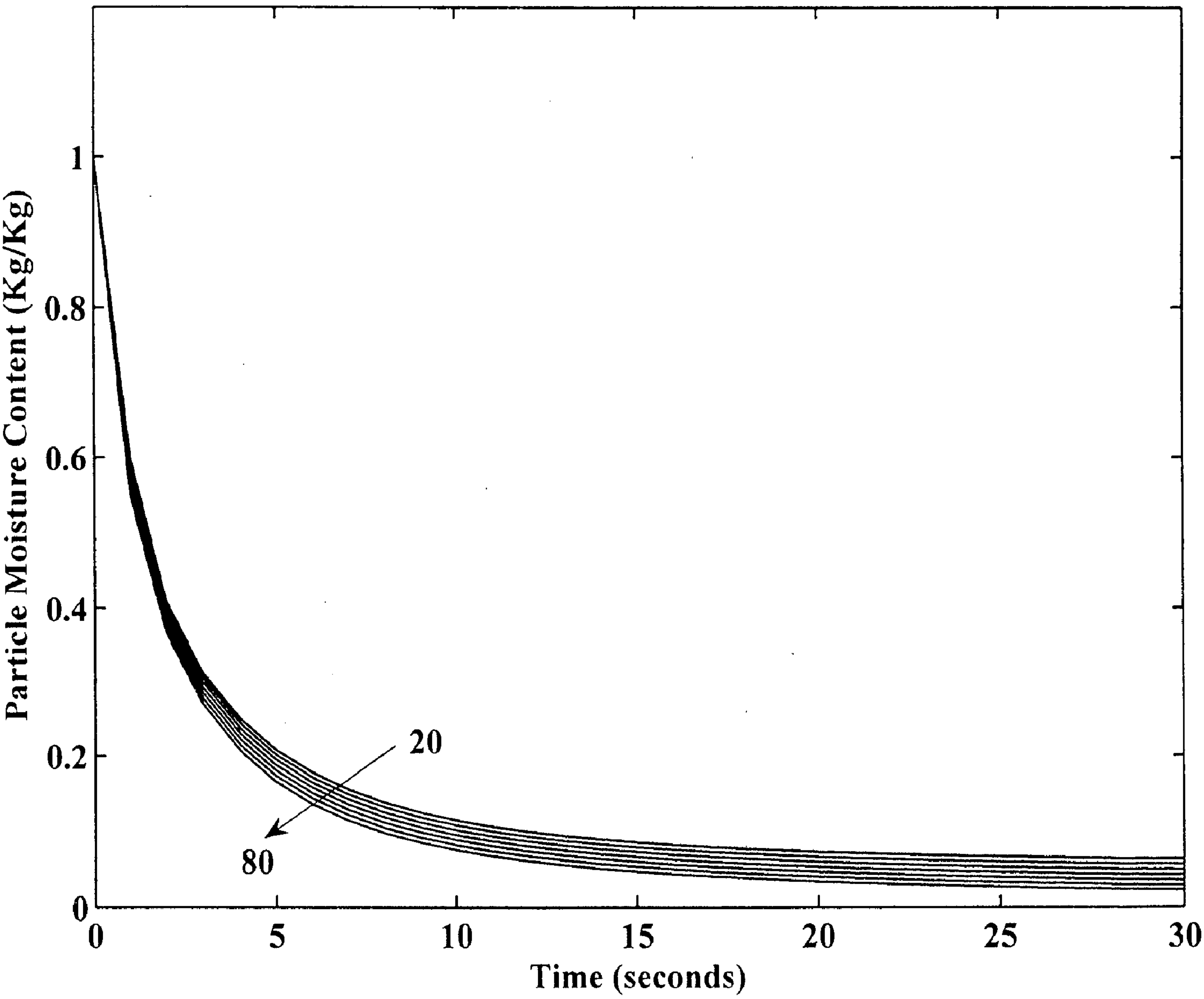


Fig 6.6 Drying Medium Humidity Profile

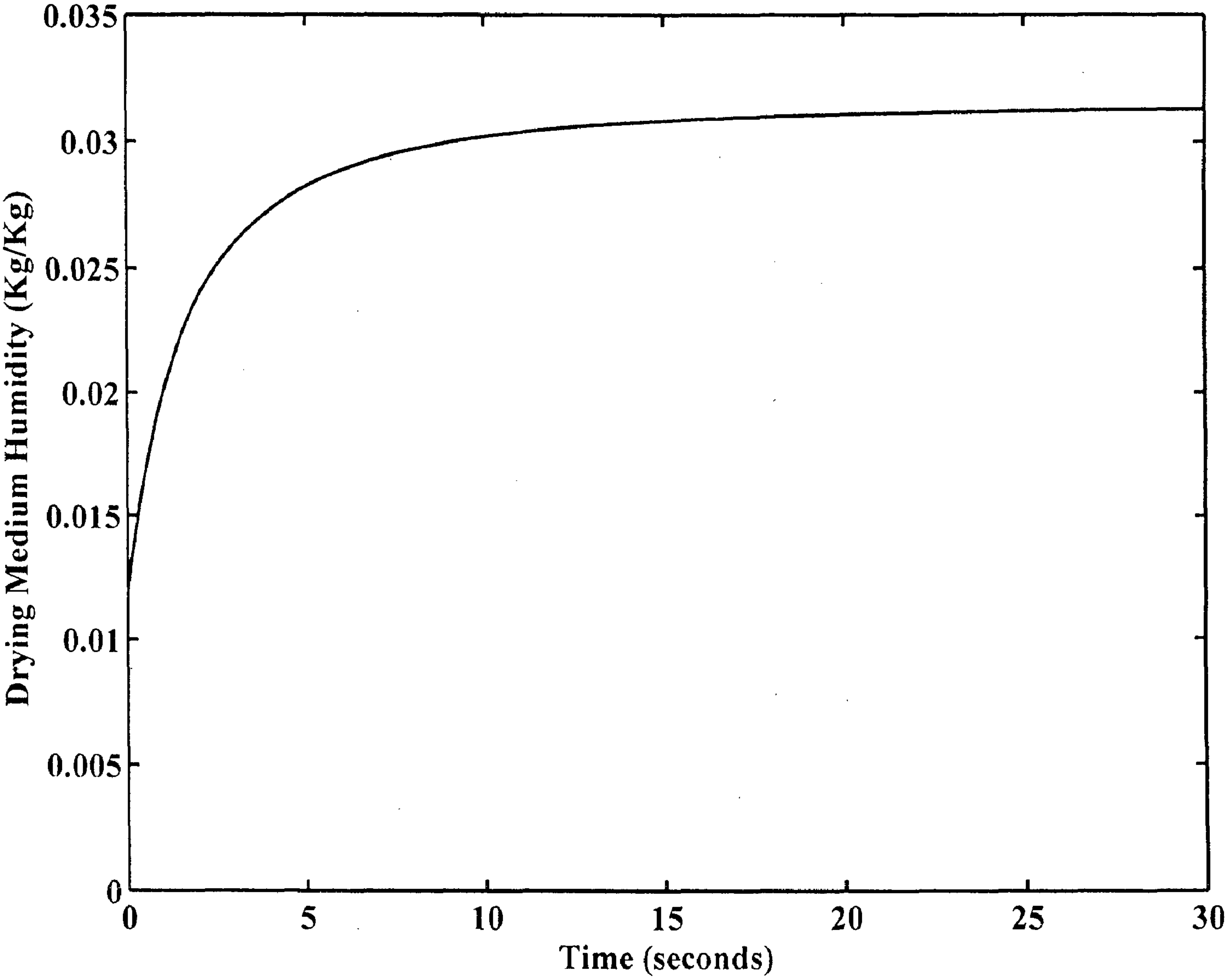


Fig 6.7 Drying Medium Humidity Profile for Different Drying Medium Temperatures

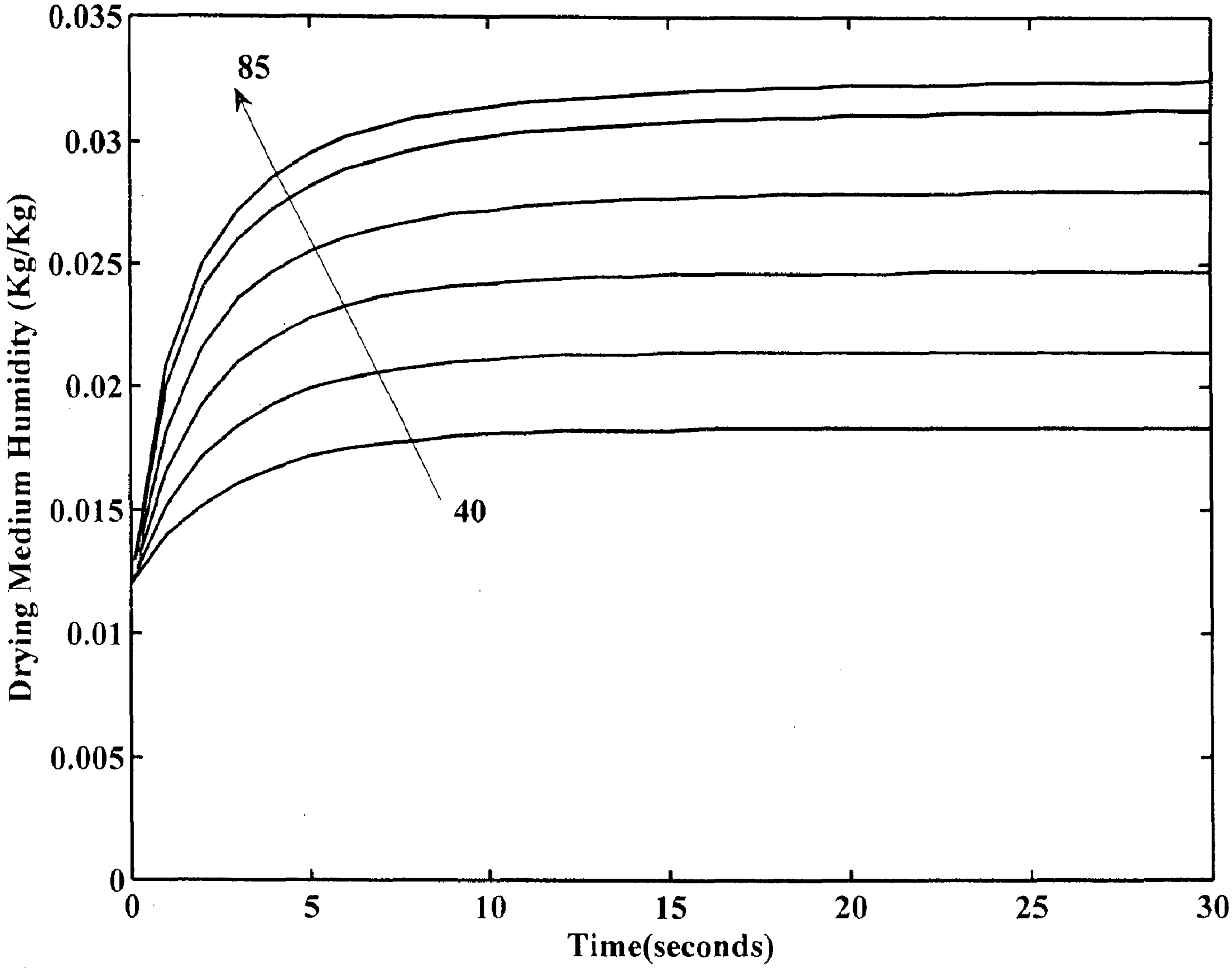


Fig 6.8 Drying Medium Humidity Profile for Different Feed Temperatures

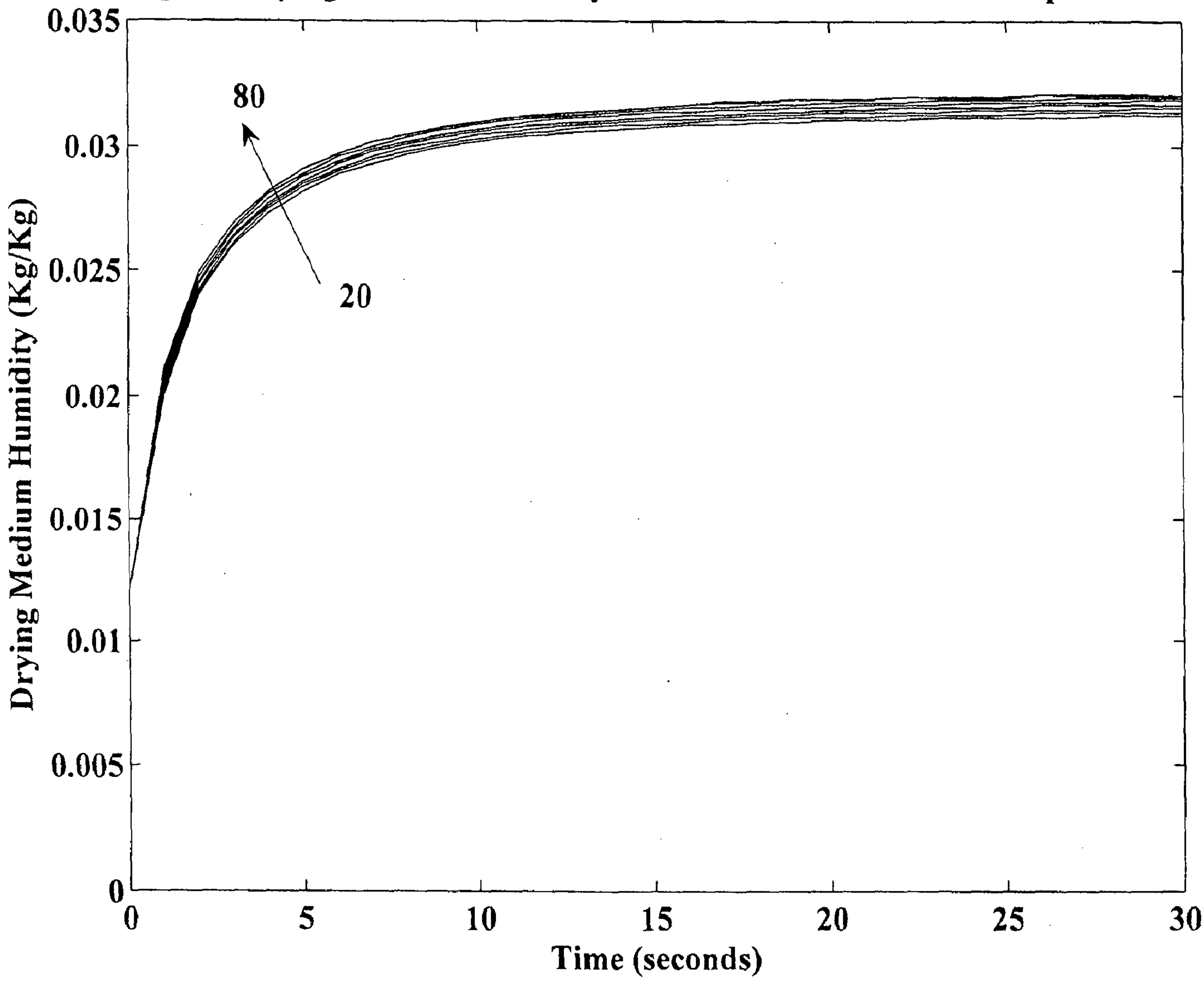
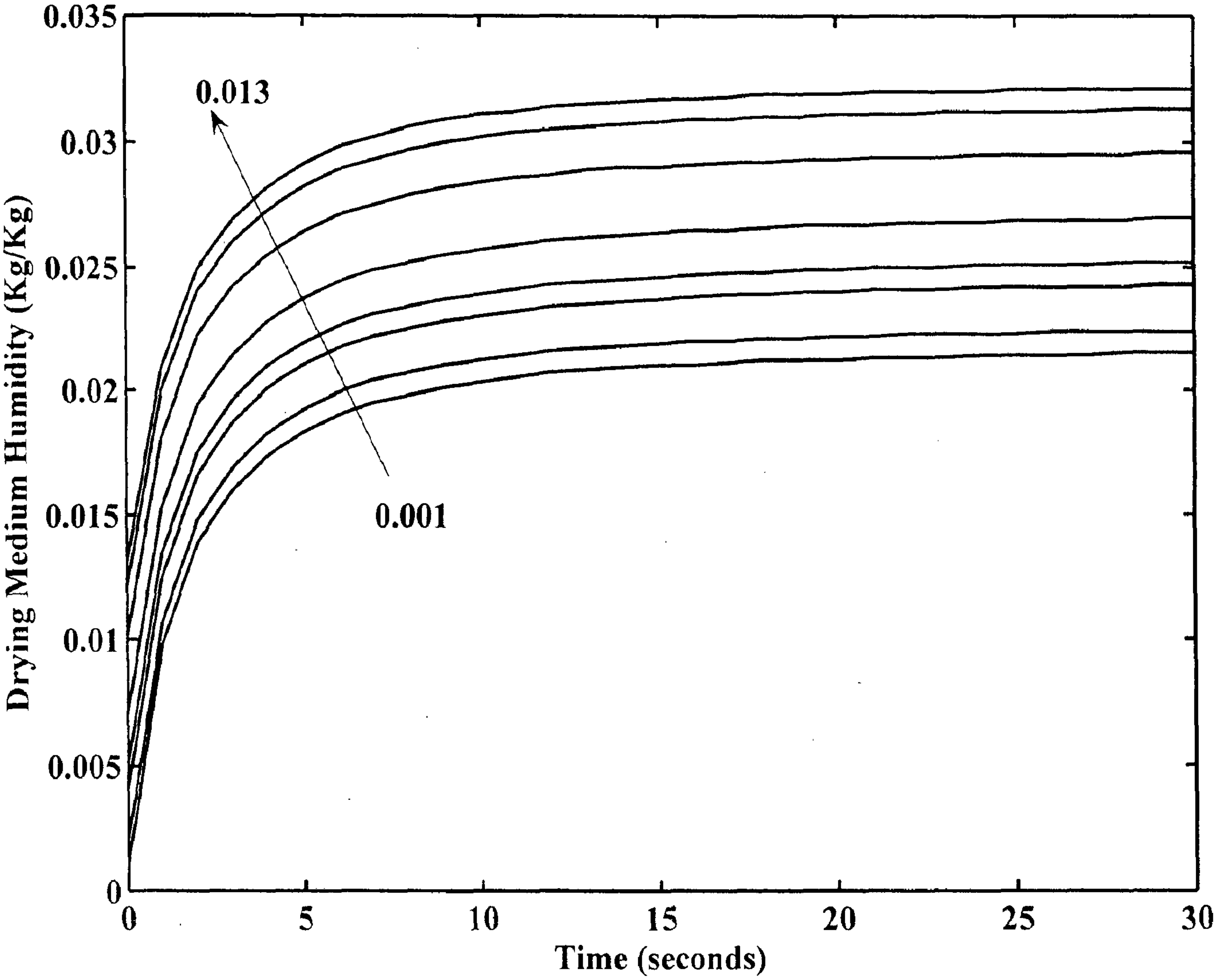


Fig 6.9 Drying Medium Humidity Profile for Different Inlet Drying Medium Humidities



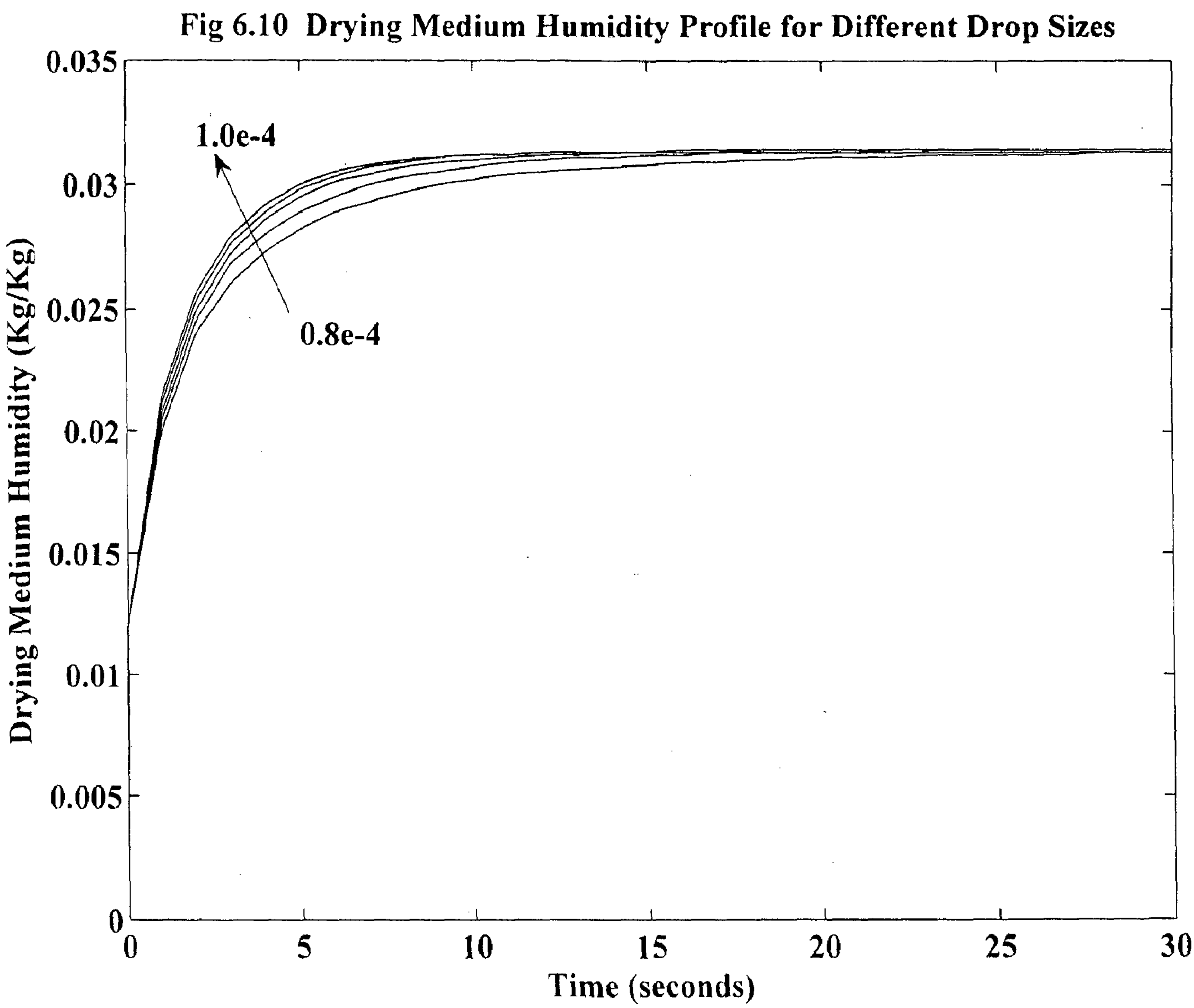


Fig 6.10 Drying Medium Humidity Profile for Different Drop Sizes
1.0e-4
0.8e-4
Drying Medium Humidity (Kg/Kg)
Time (seconds)
0.035
0.03
0.025
0.02
0.015
0.01
0.005
0
5
10
15
20
25
30

Fig 6.11 Particle Temperature Profile

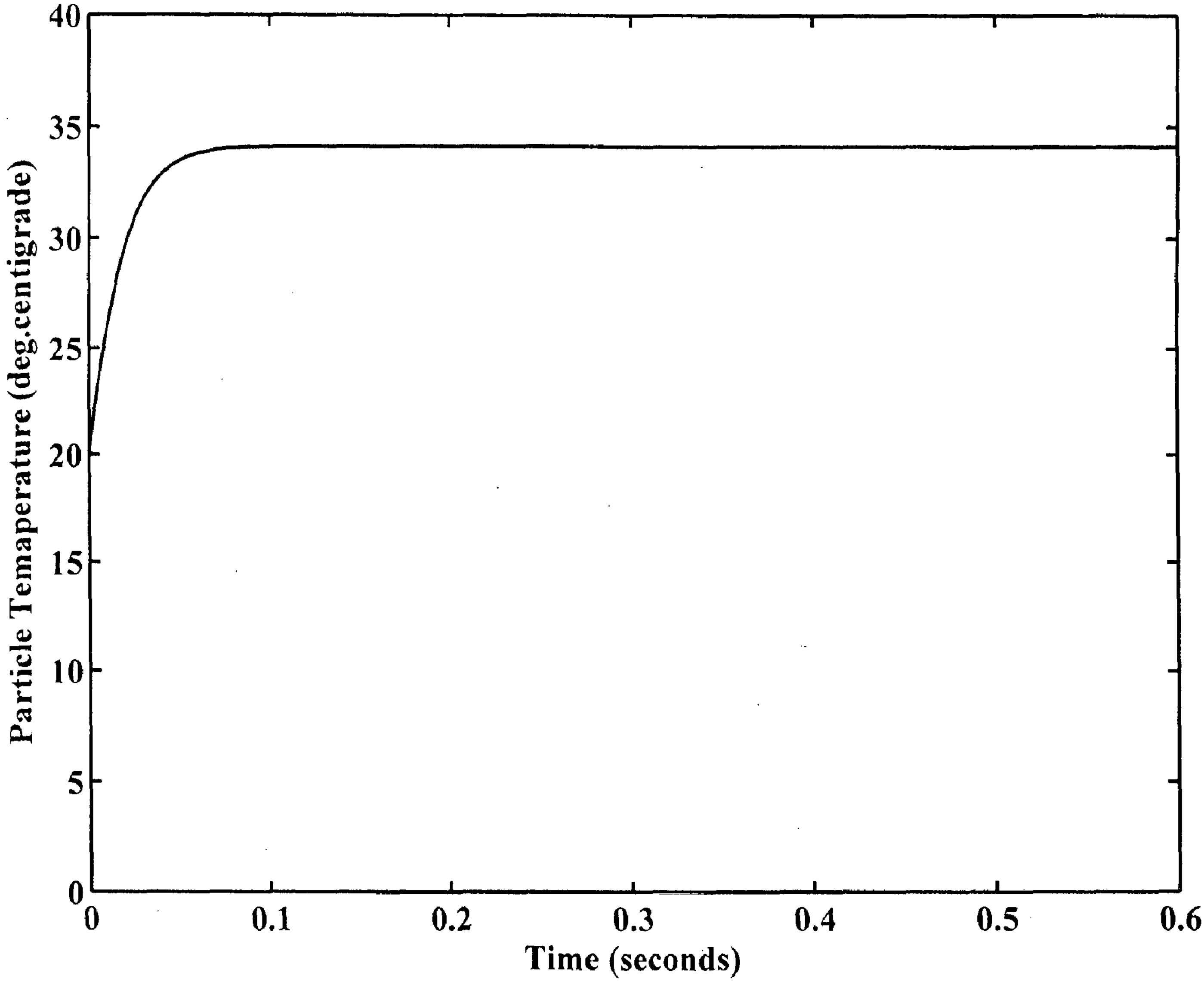


Fig 6.12 Particle Temperature Profile

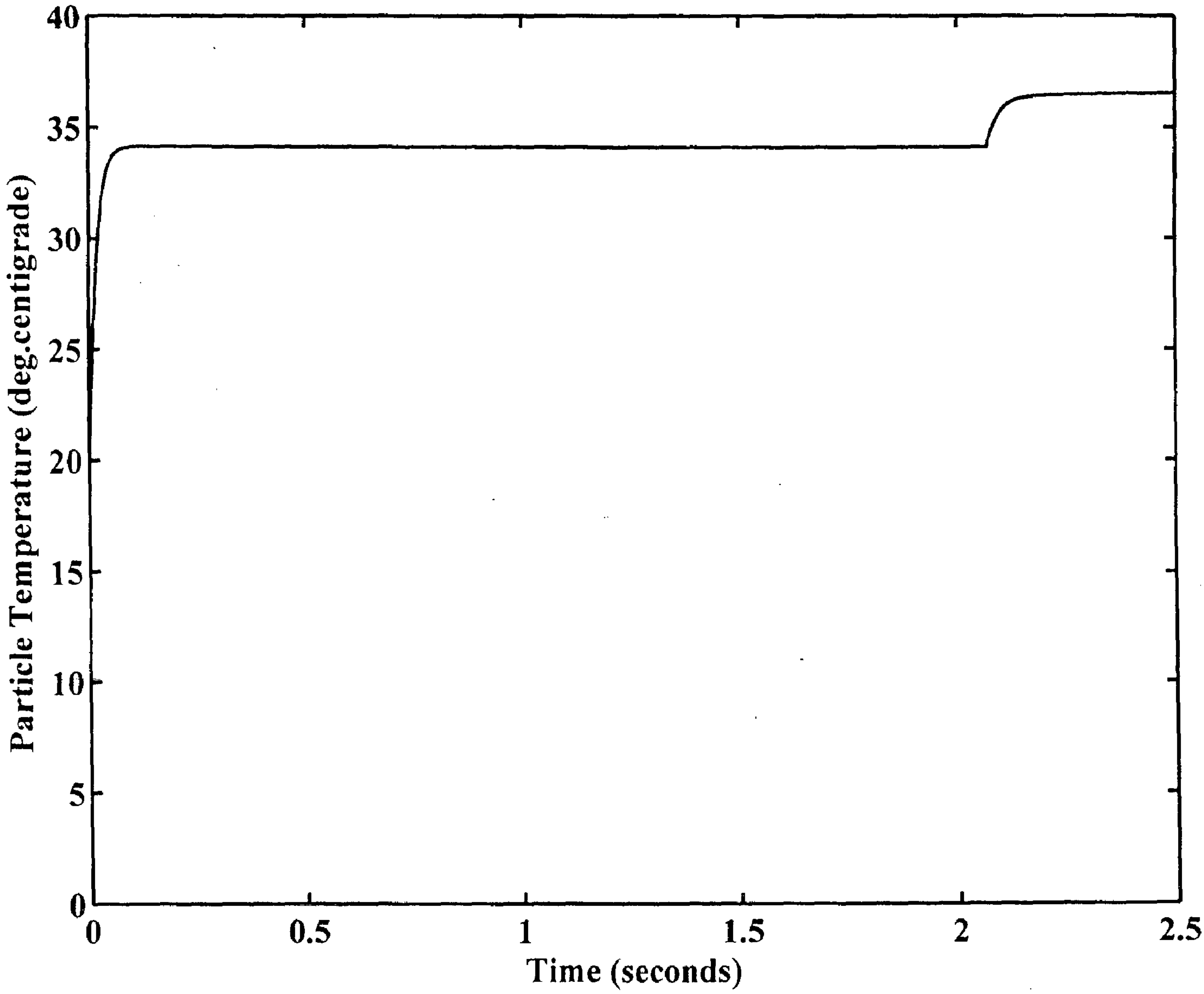


Fig 6.13 Particle Temperature Profile for Different Drop Sizes

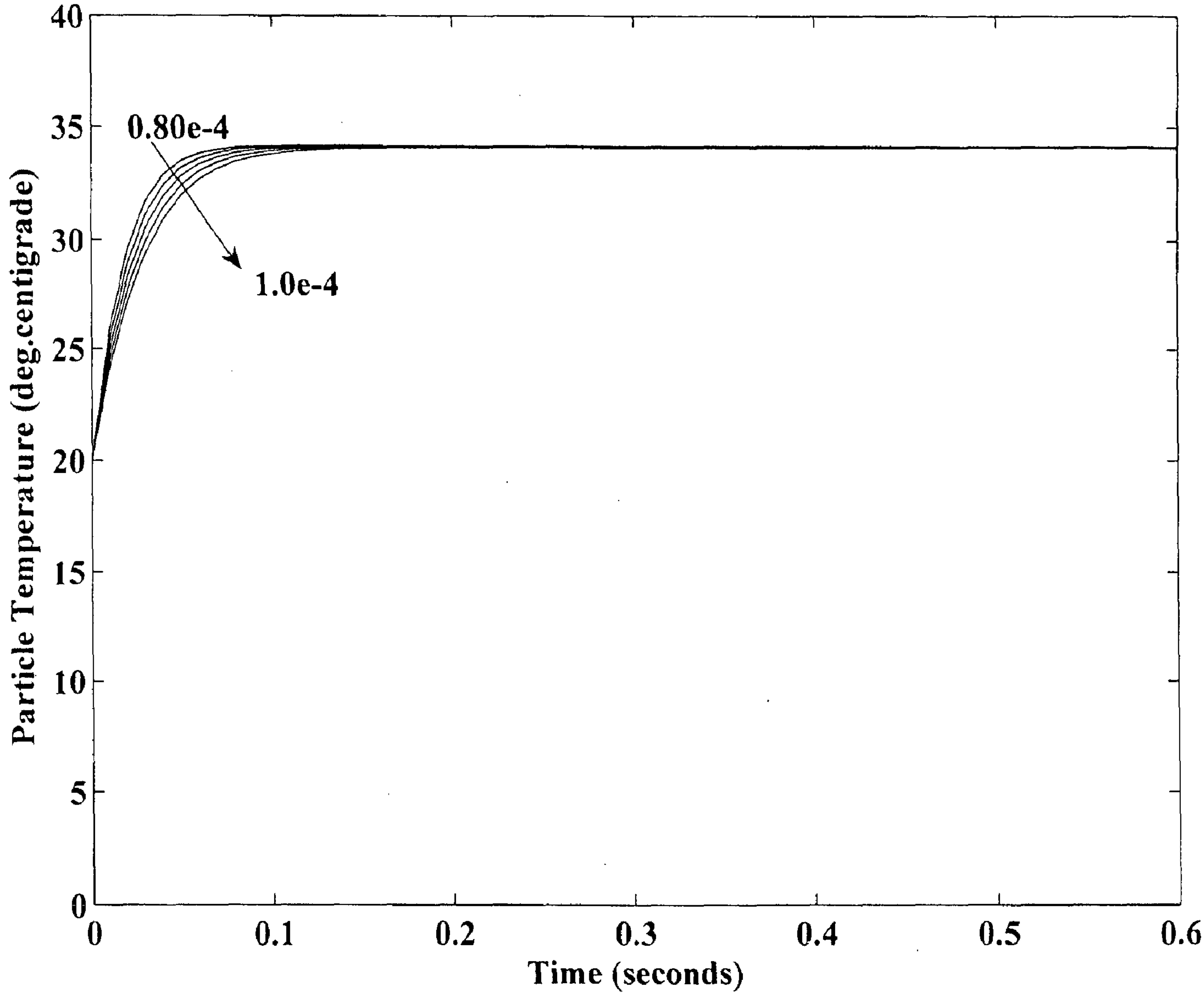


Fig 6.14 Particle Temperature Profile for Different Drying Medium Humidities

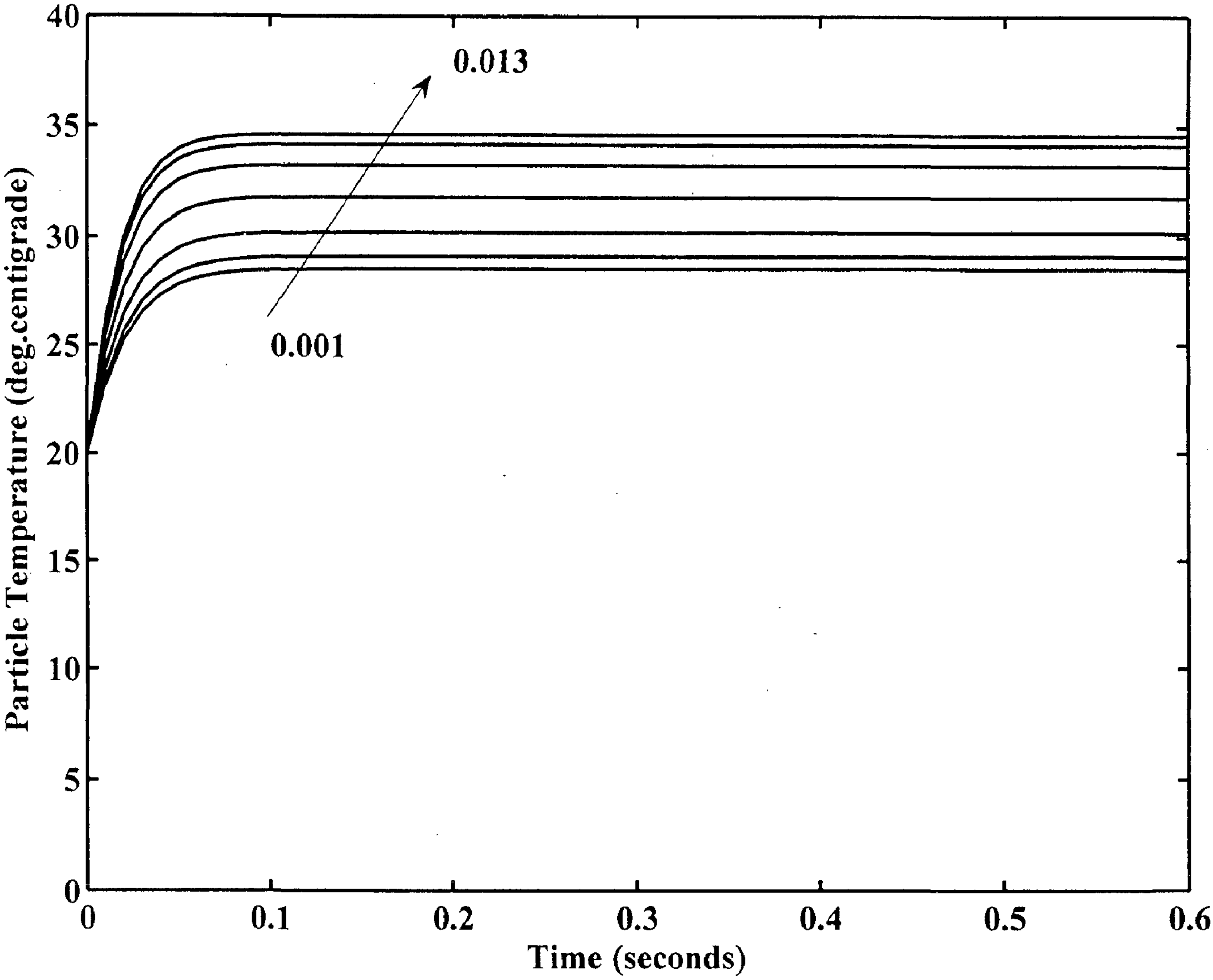


Fig 6.15 Particle Temperature Profile for Different Drying Medium Inlet Temperatures

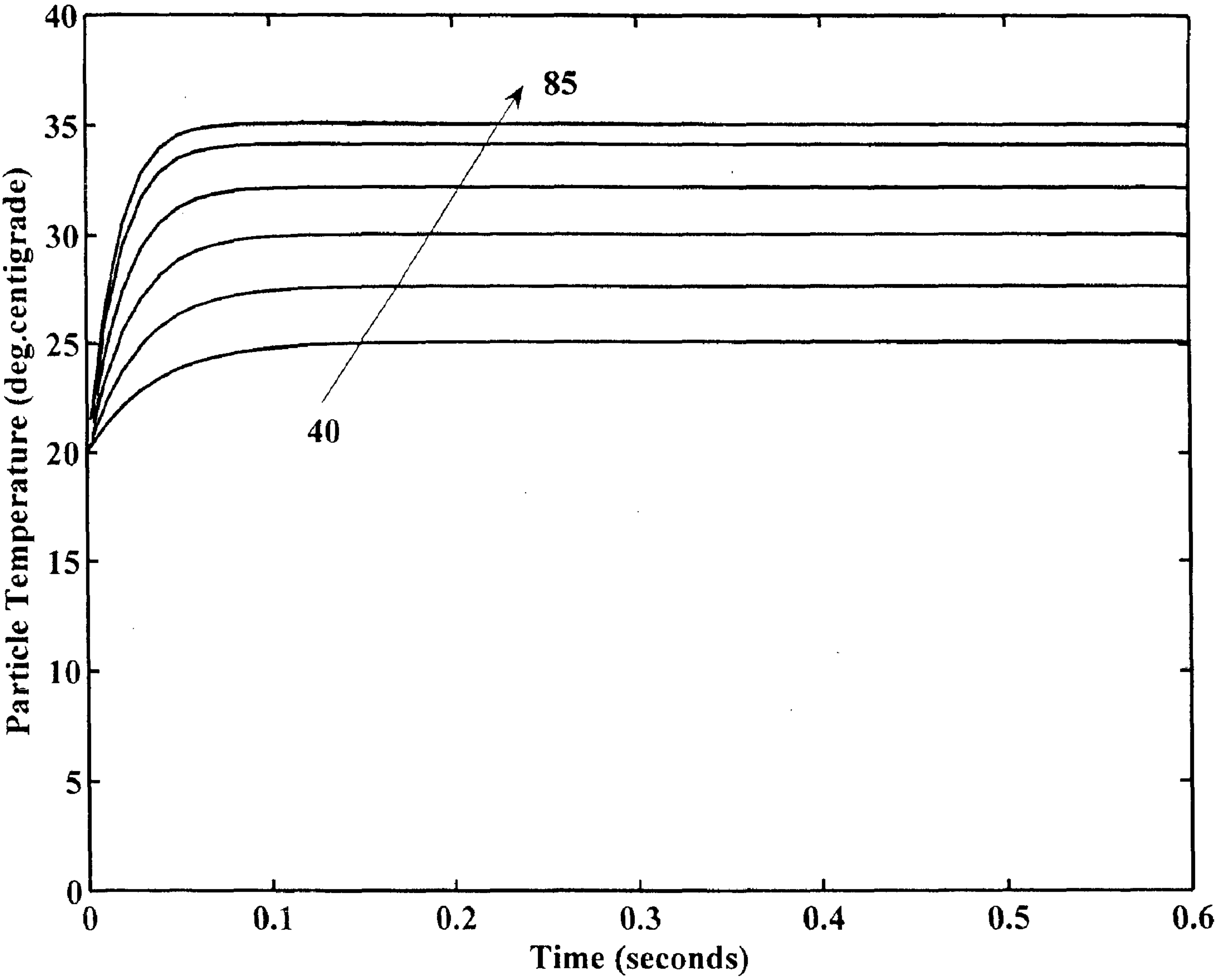


Fig 6.16 Particle Temperature Profile for Different Feed Temperatures

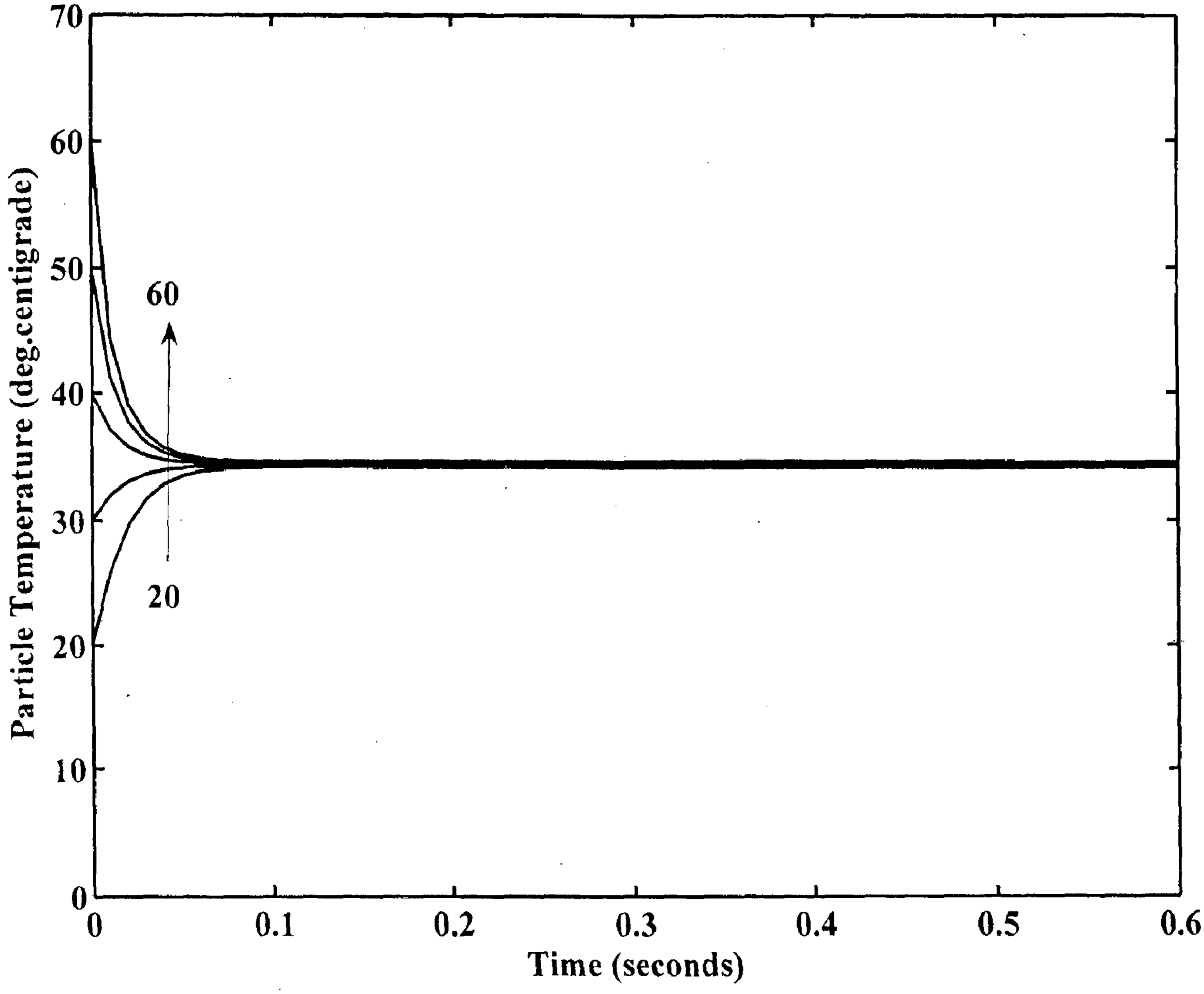


Fig 6.17 Drying Medium Temperature Profile

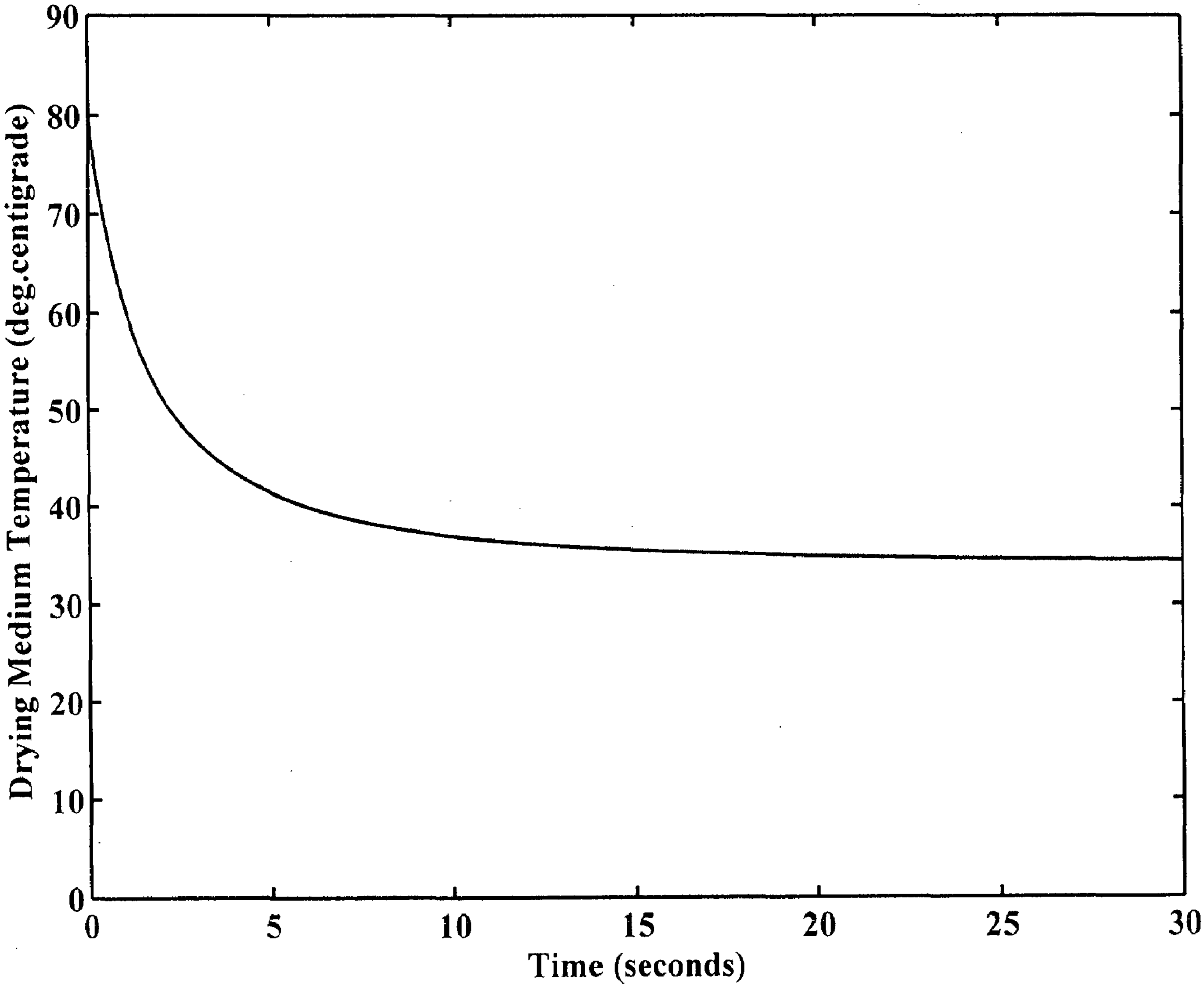


Fig 6.18 Drying Medium Temperature Profile for Different Drop Sizes

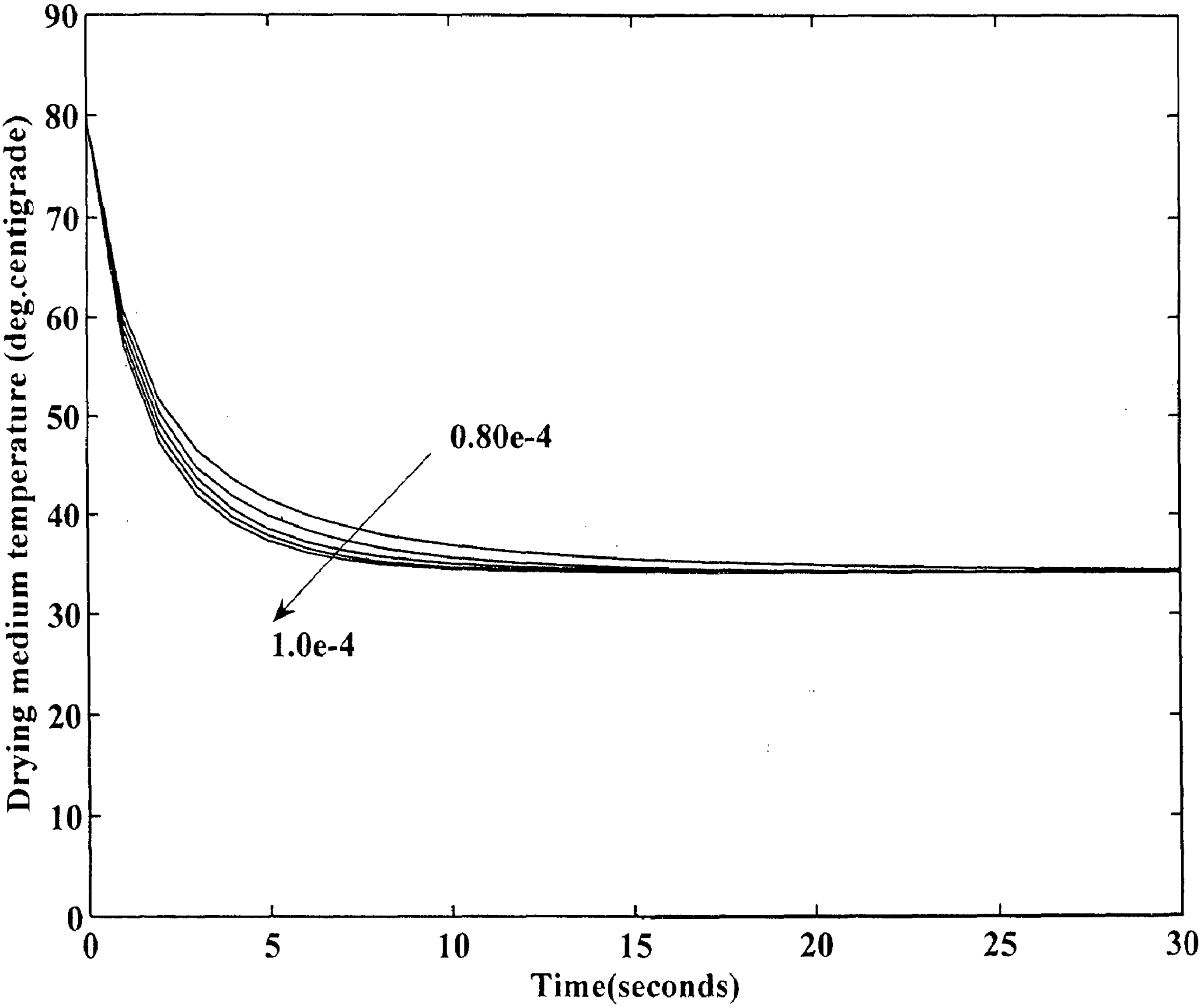


Fig 6.19 Drying Medium Temperature Profile for Different Drying Medium Humidities

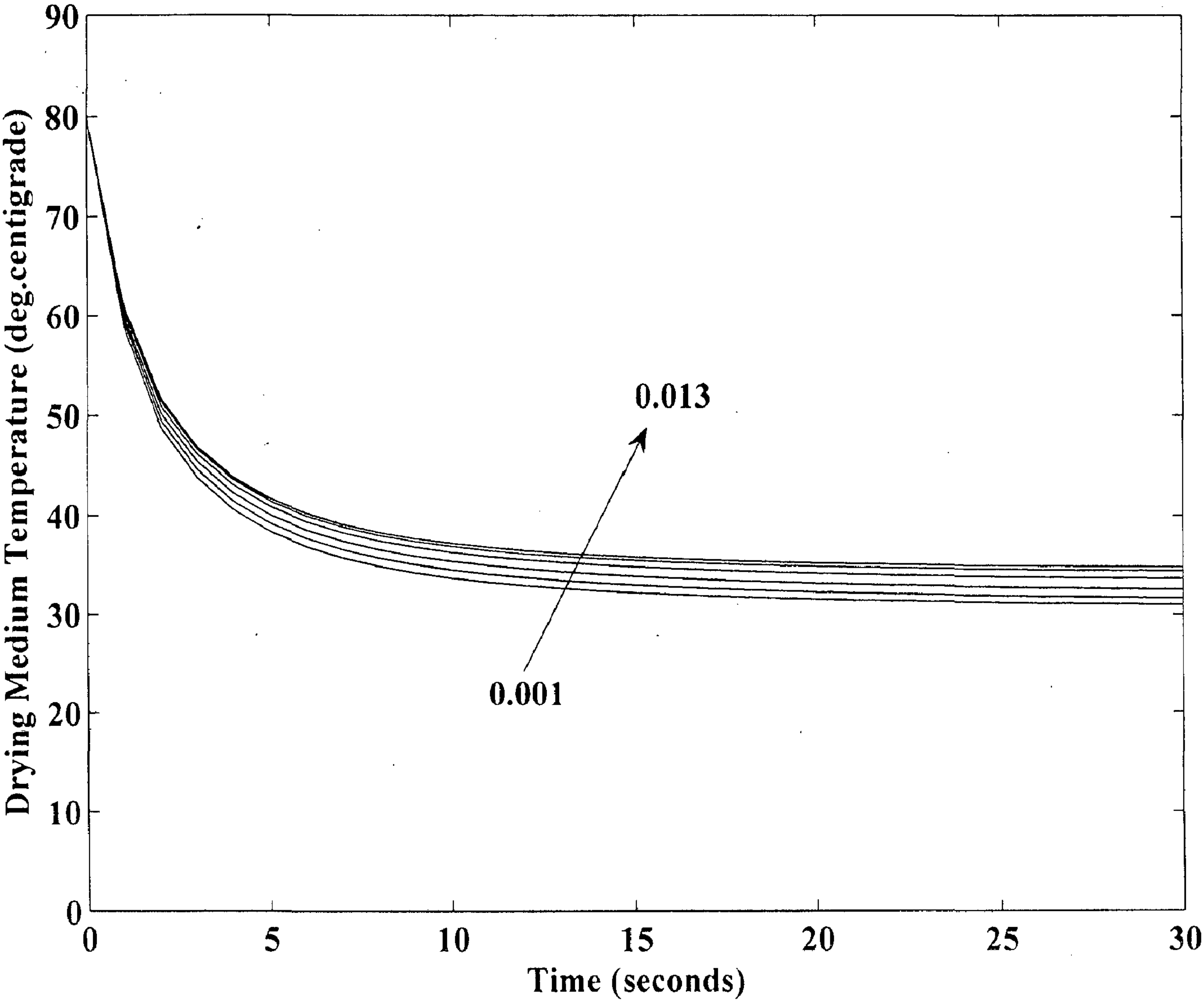


Fig 6.20 Drying Medium Temperature Profile for Different Drying Medium Temperatures

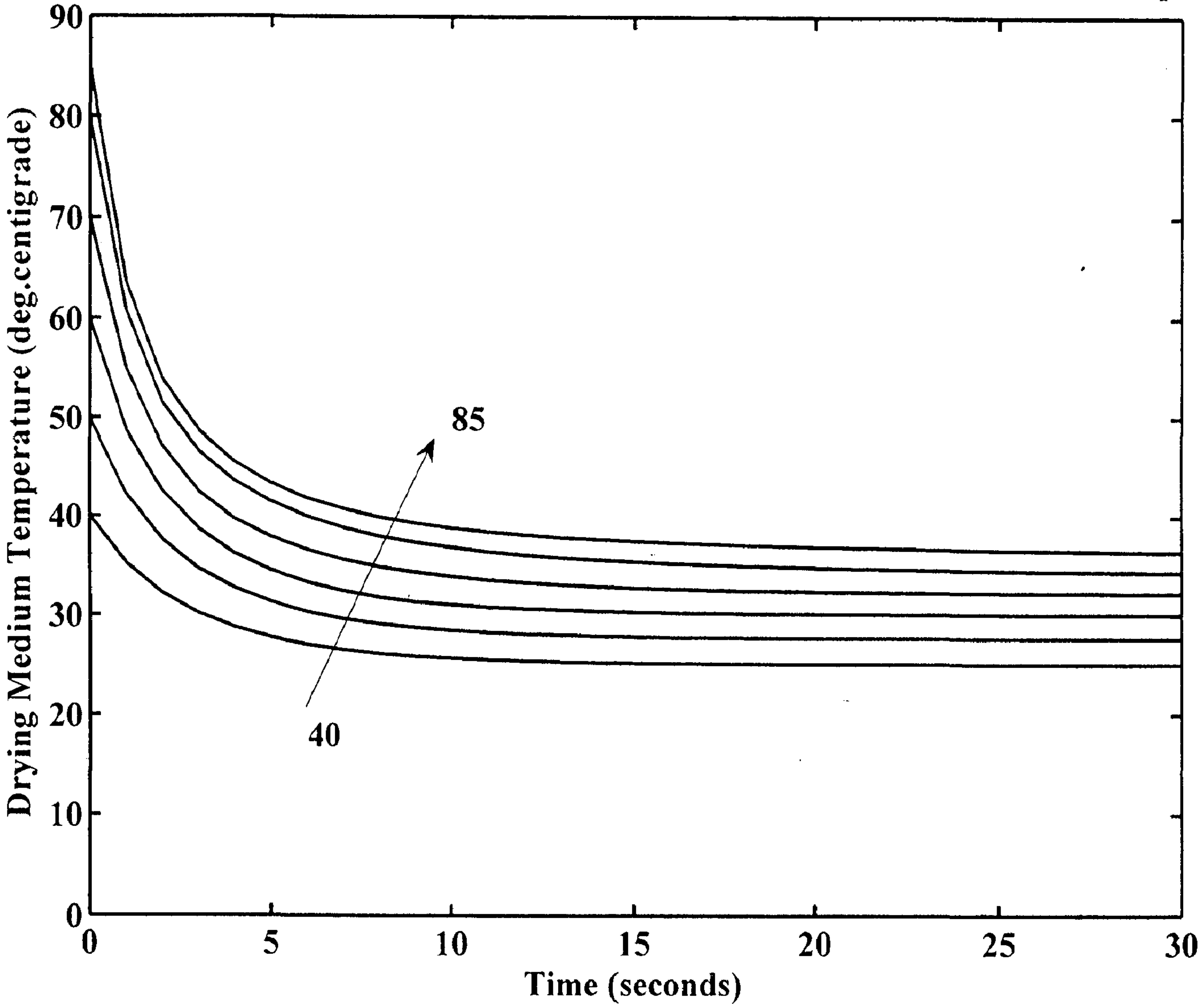


Fig 6.21 Drying Medium Temperature Profile for Different Feed Temperatures

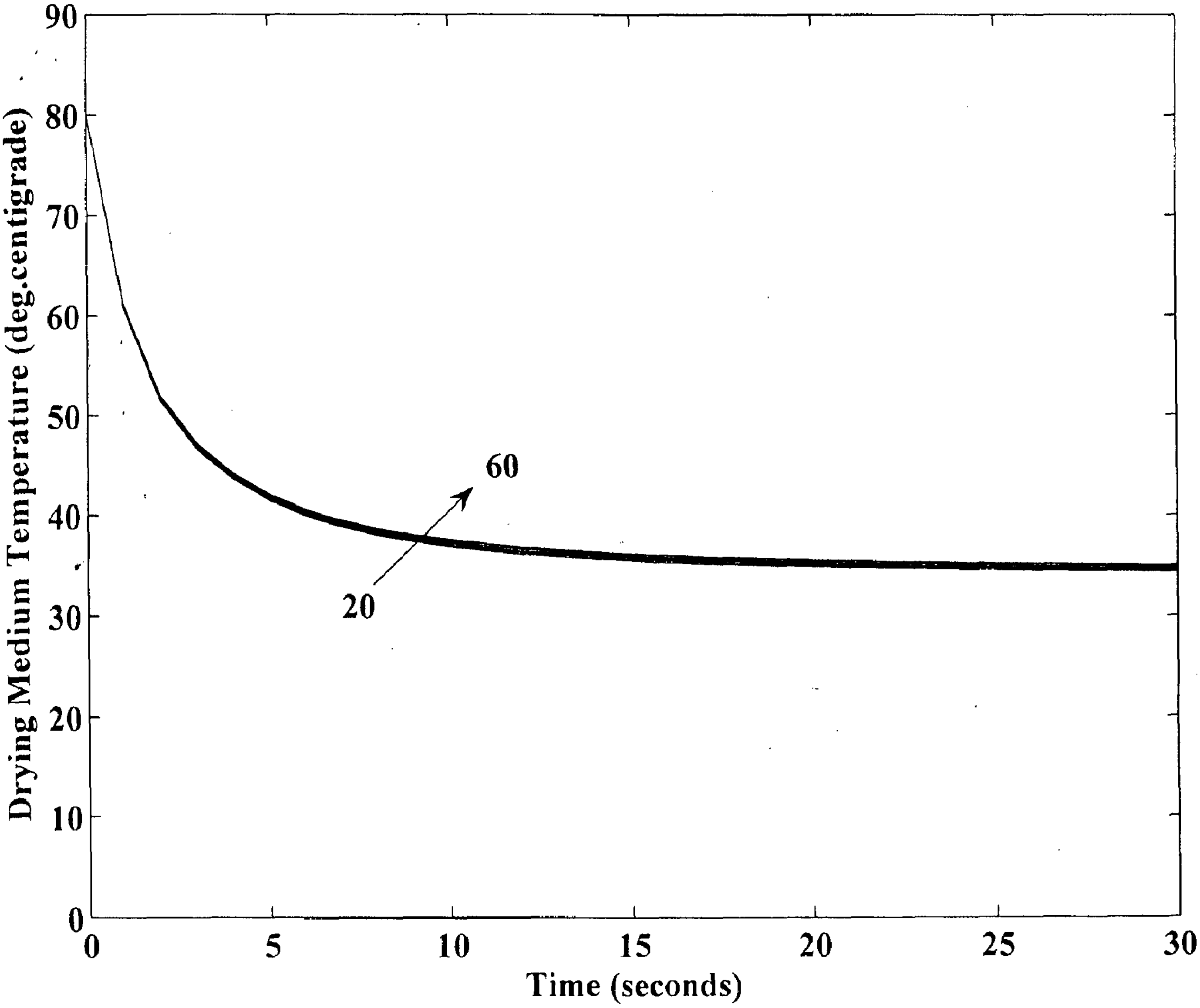


Fig 6.22 Axial Velocity Profile

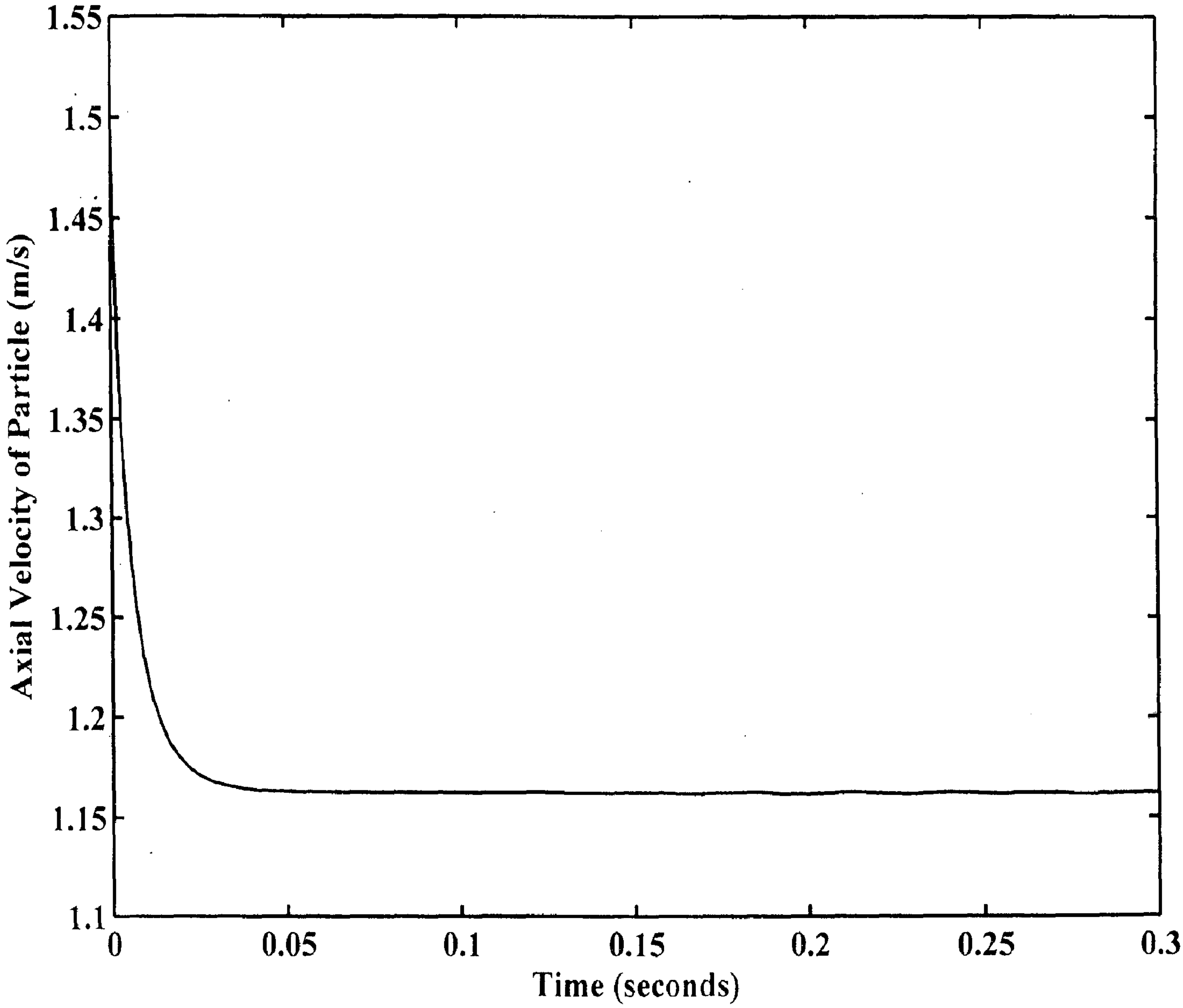


Fig 6.23 Axial Velocity Profile for Different Drop Sizes

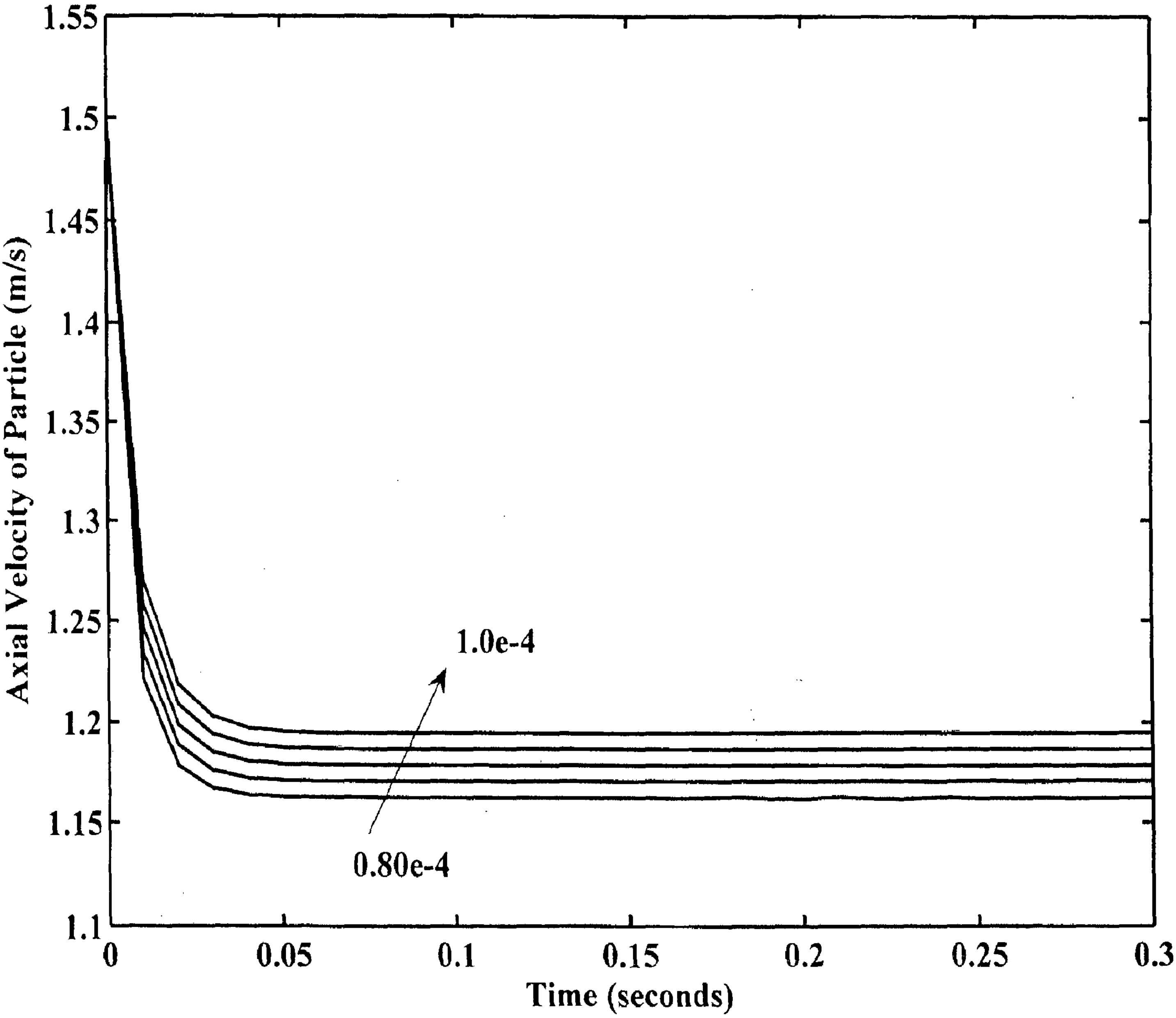


Fig 6.24 Density Profile

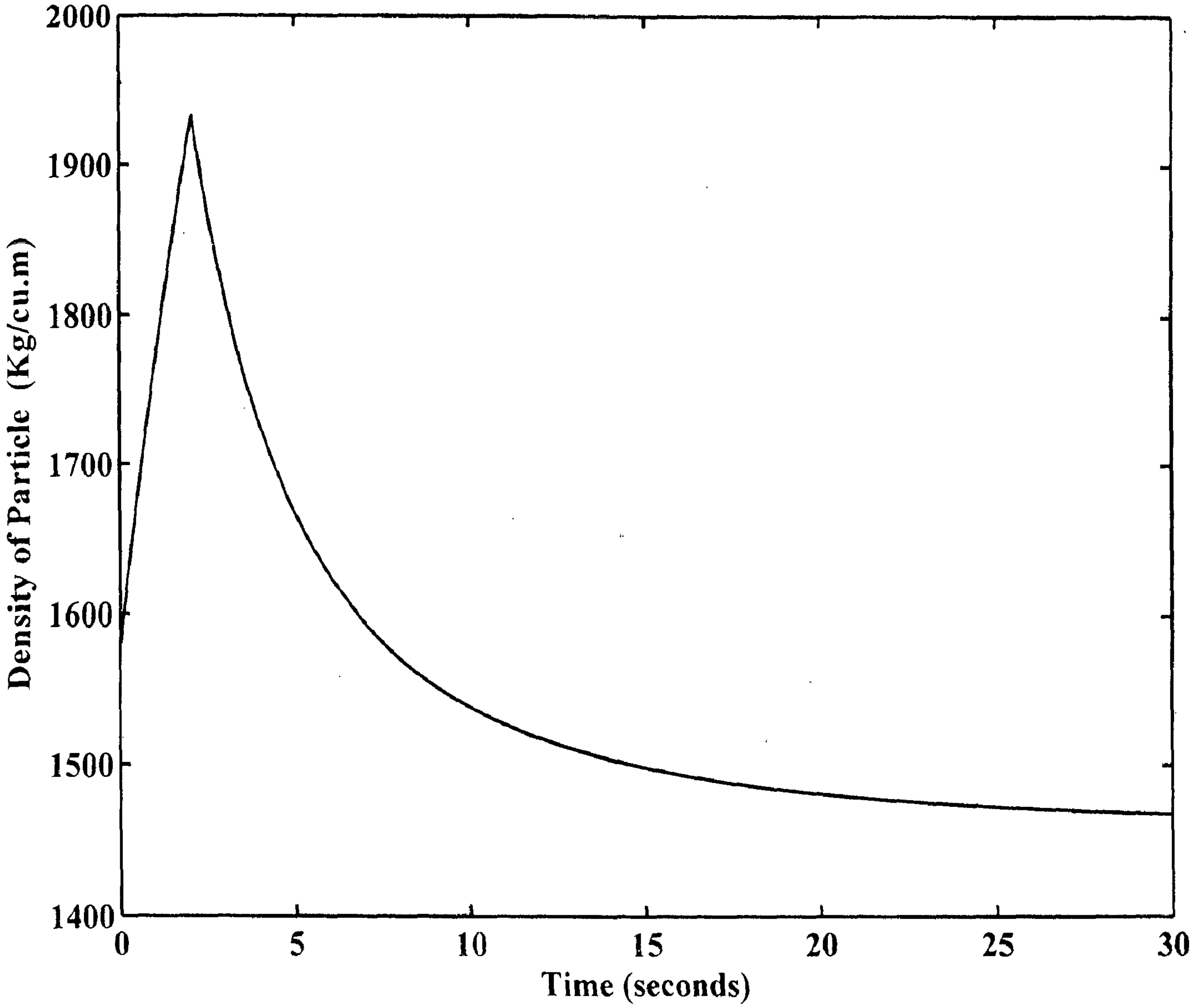


Fig 6.25 Density Profile for Different Feed Temperatures

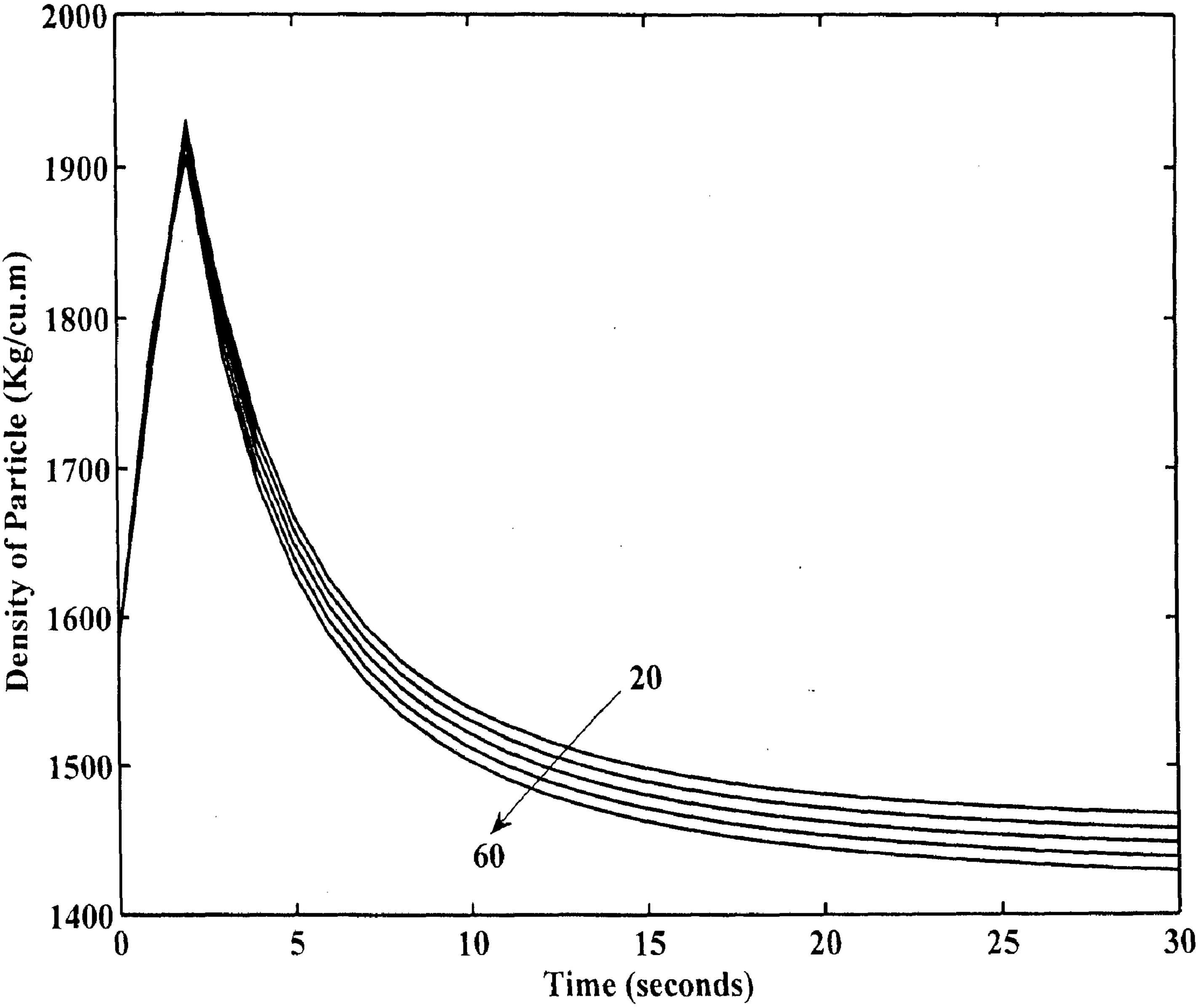


Fig 6.26 Density Profile for Different Drying Medium Humidities

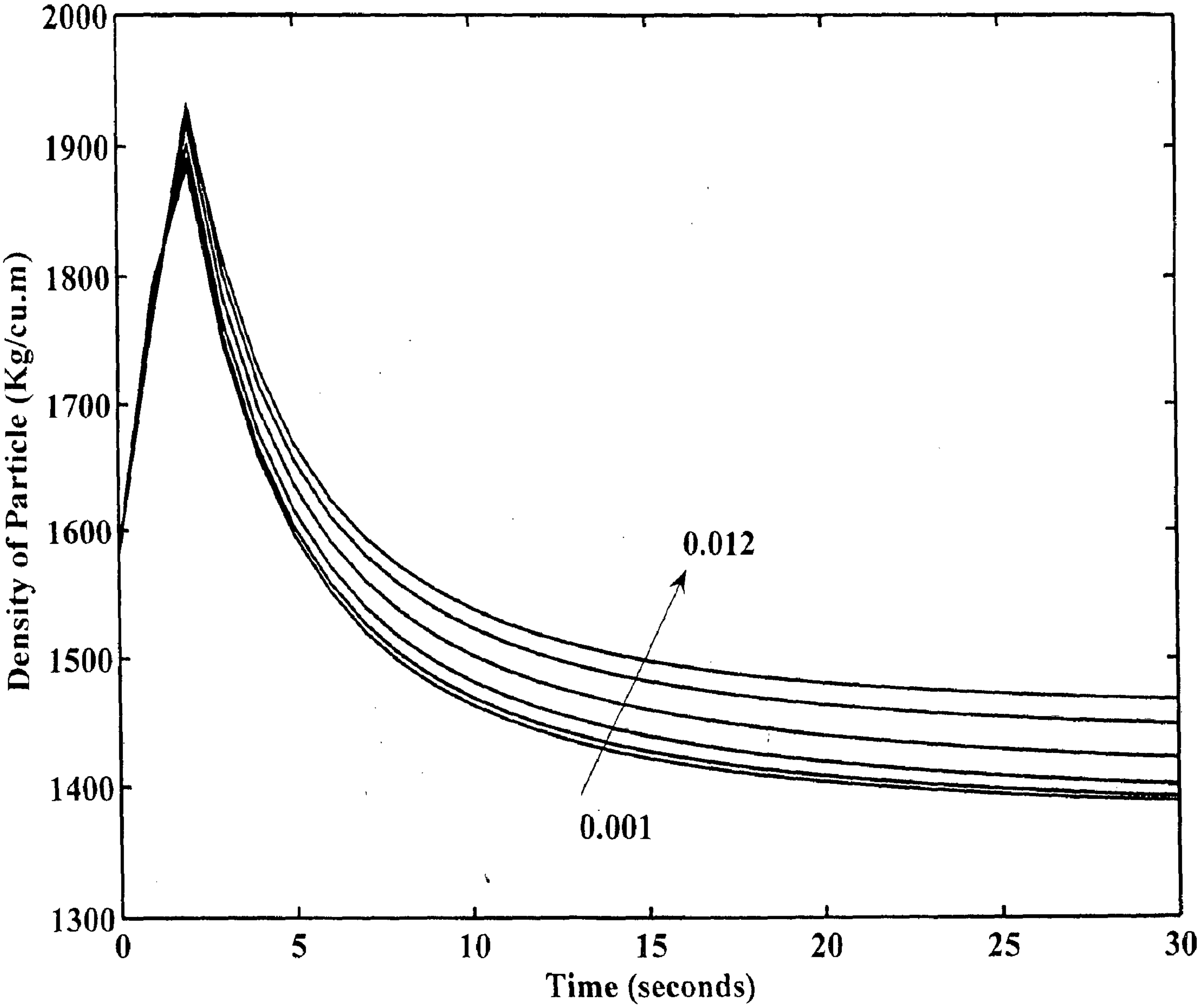


Fig 6.27 Density Profile for Different Drying Medium Inlet Temperatures

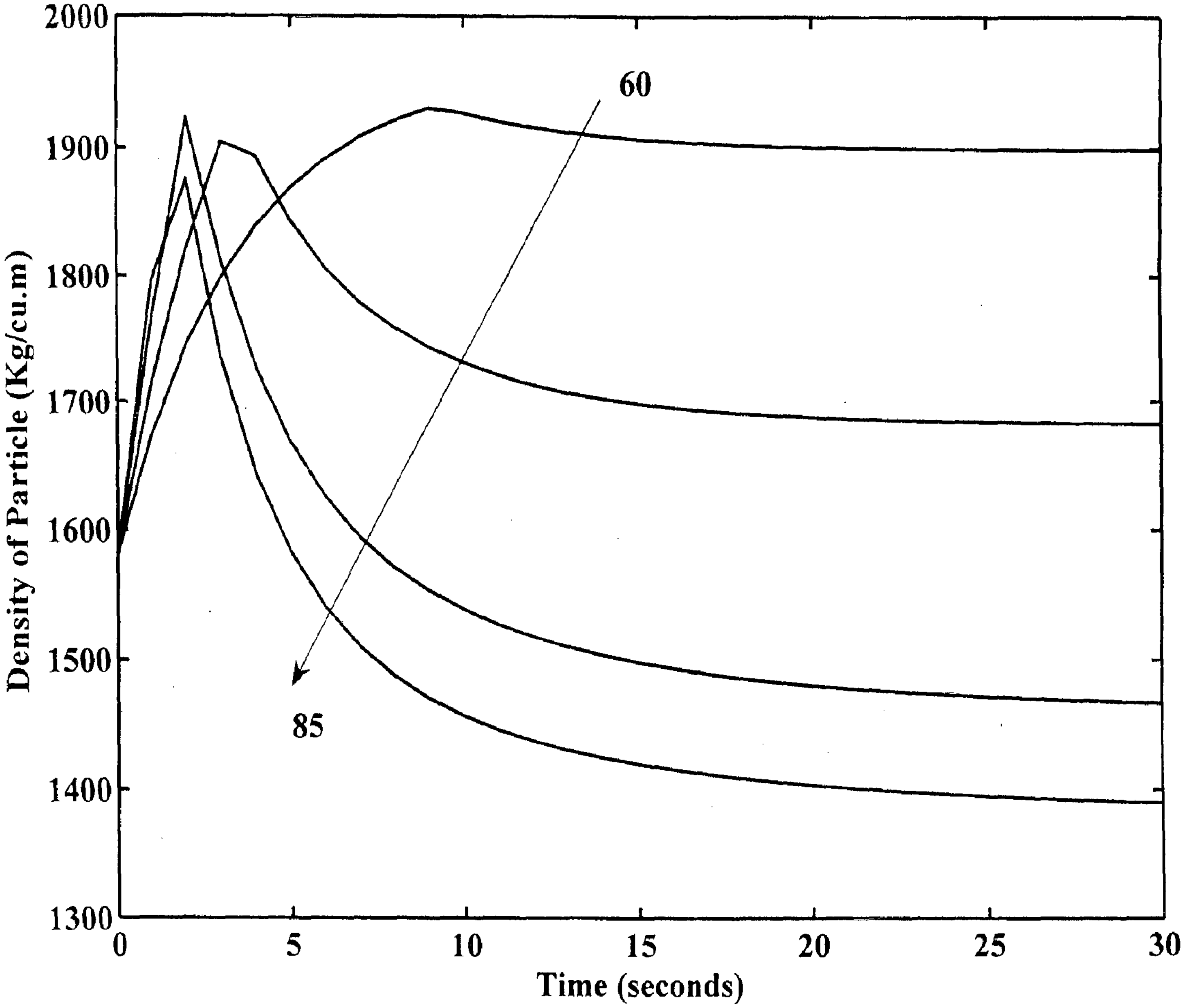


Fig 6.28 Density Profile for Different Drop Sizes

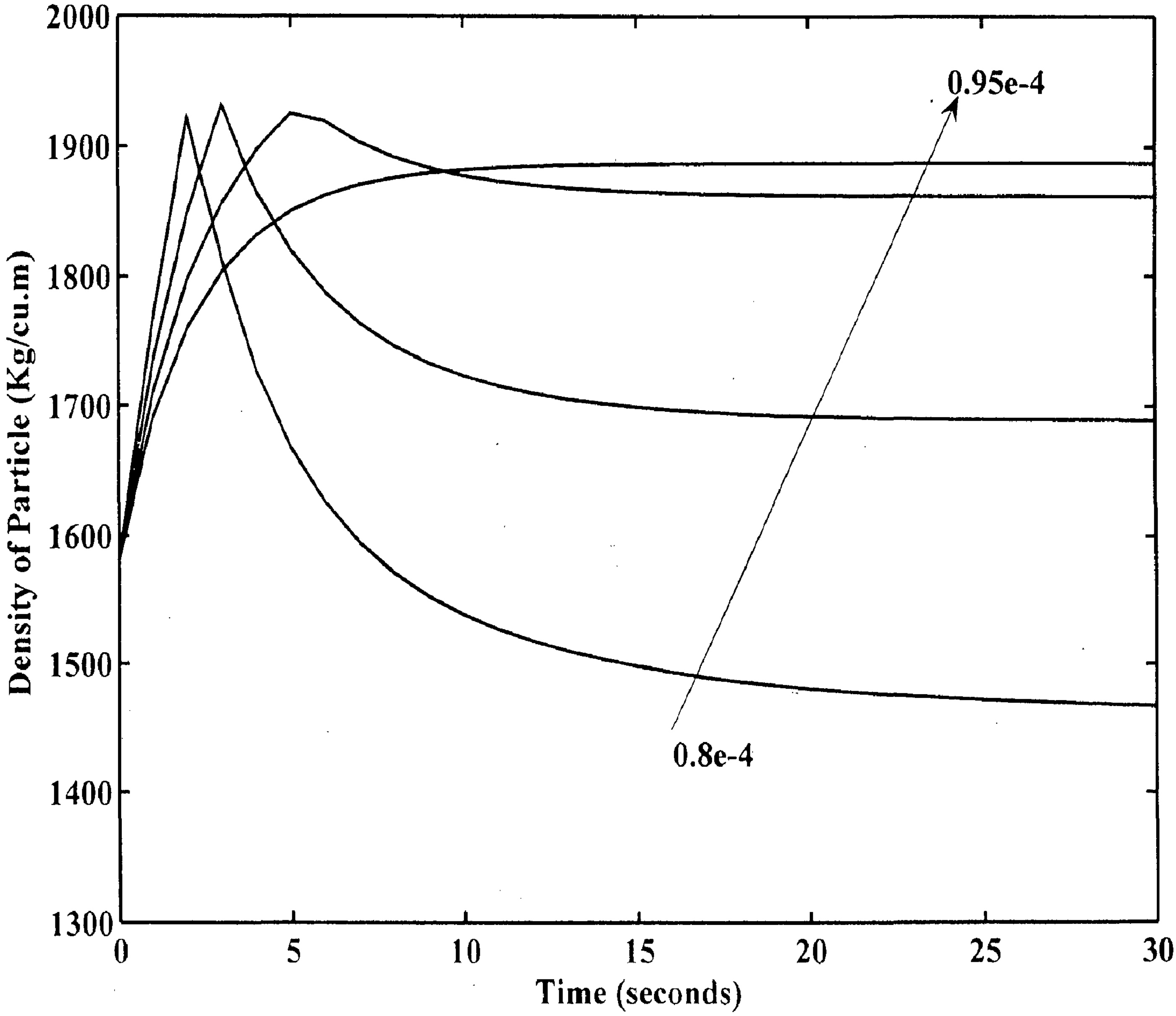


Fig 6.29 Diameter Profile

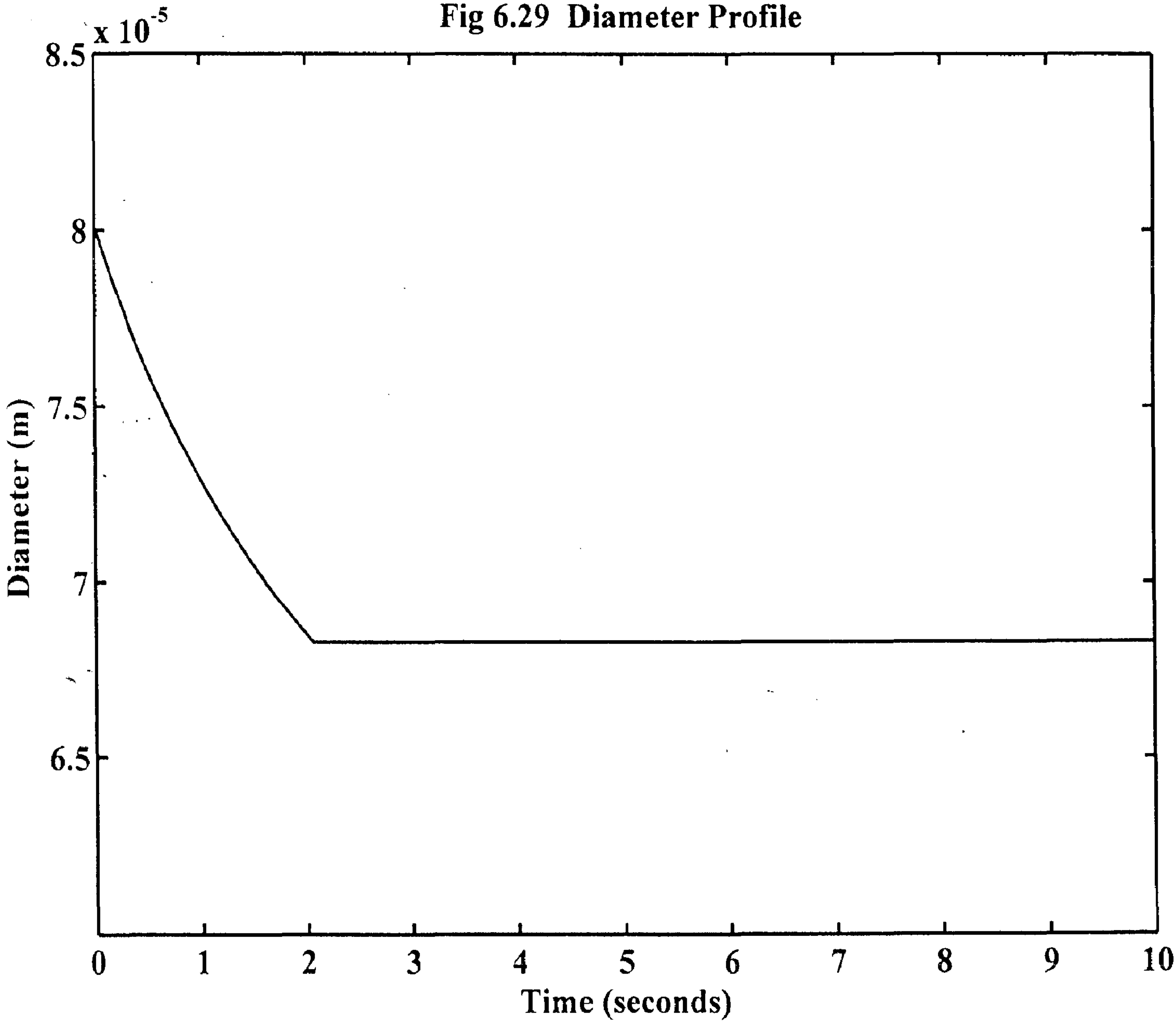


Fig 6.30 Diameter Profile for Different Feed Temperatures

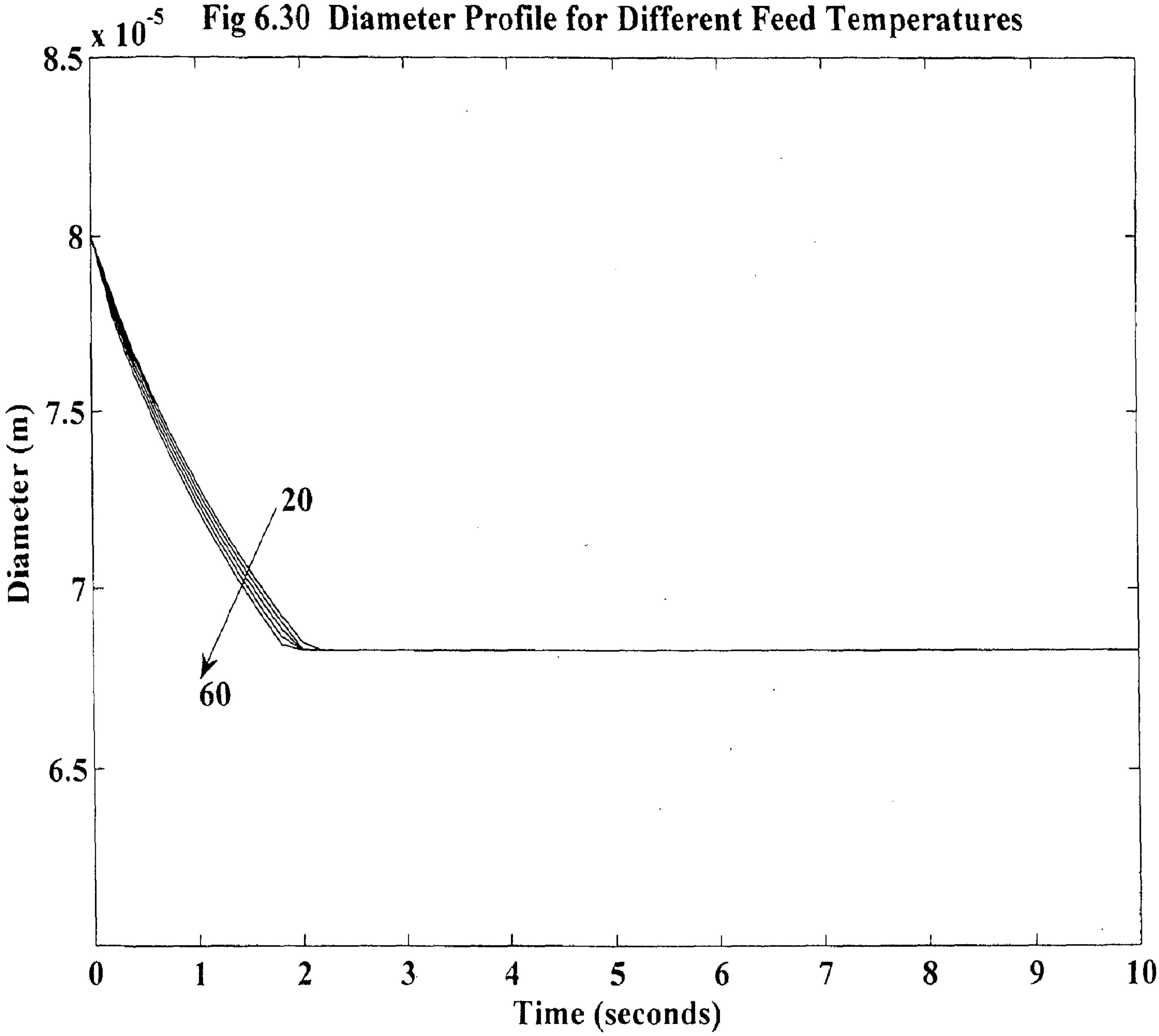
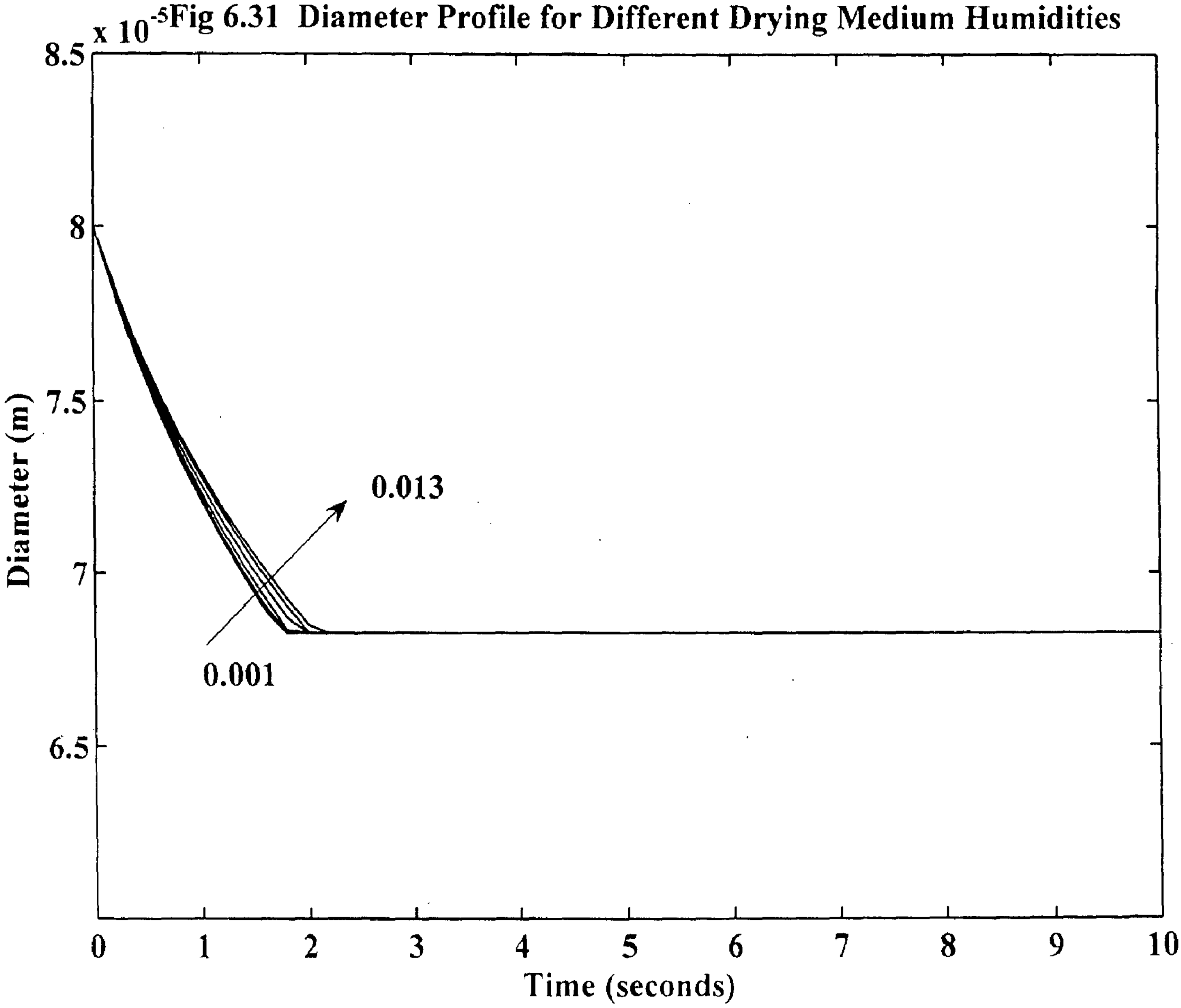


Fig 6.31 Diameter Profile for Different Drying Medium Humidities



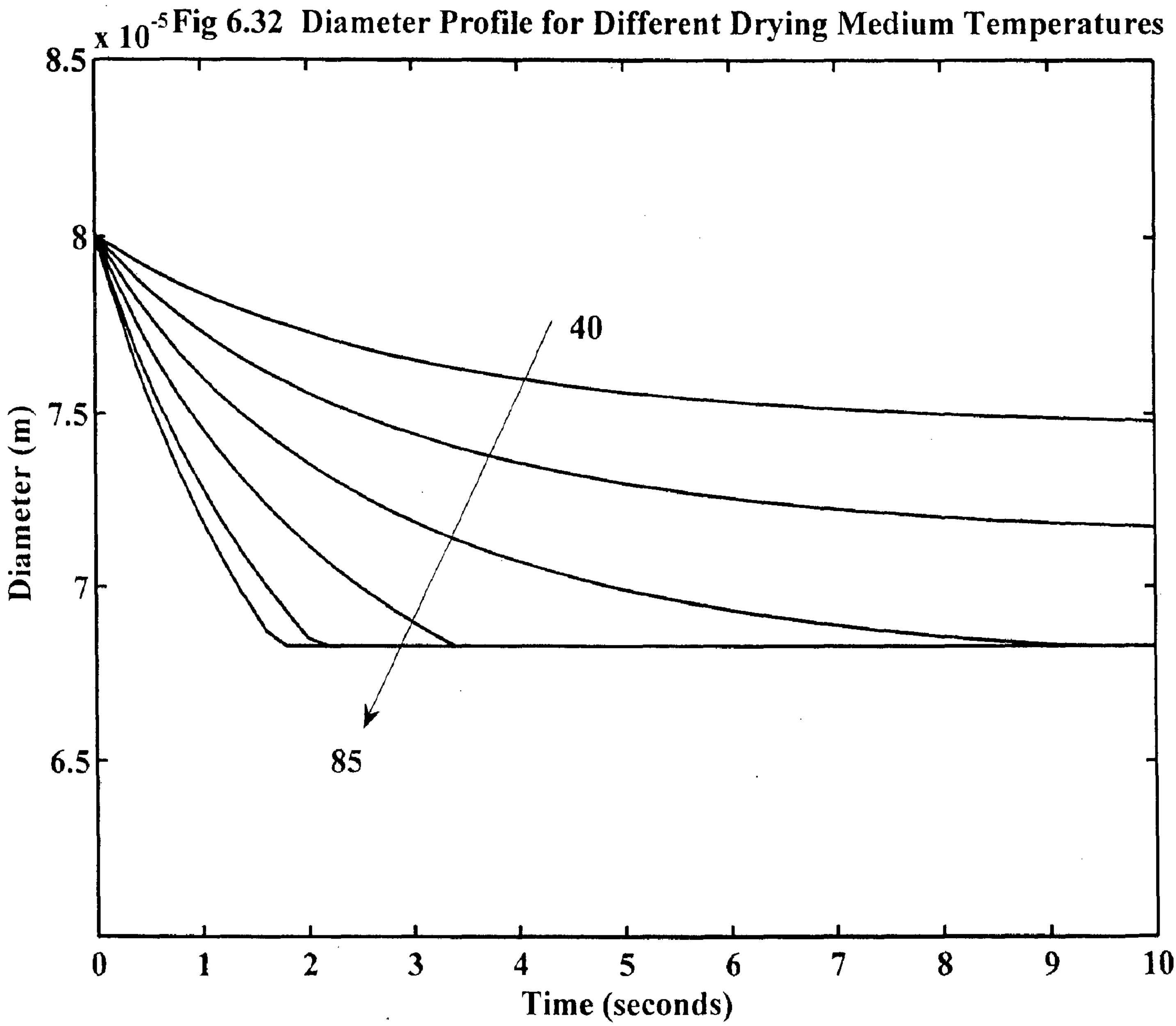

x 10^-5 Fig 6.32 Diameter Profile for Different Drying Medium Temperatures
8.5
8
7.5
7
6.5
40
85
Diameter (m)
0
1
2
3
4
5
6
7
8
9
10
Time (seconds)

Fig 6.33 Diameter Profile for Different Drop Sizes

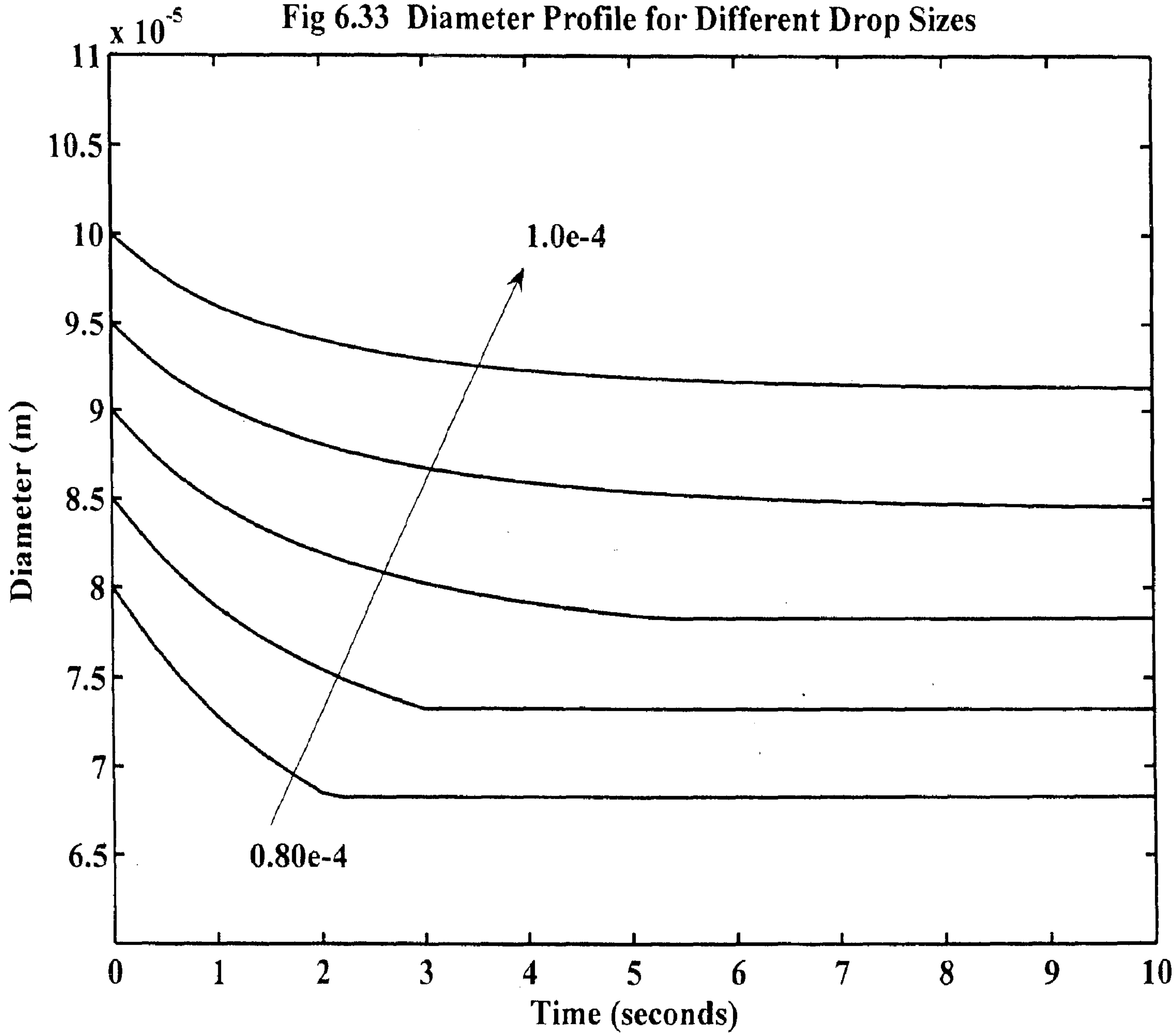


Fig 6.34 Porosity Profile

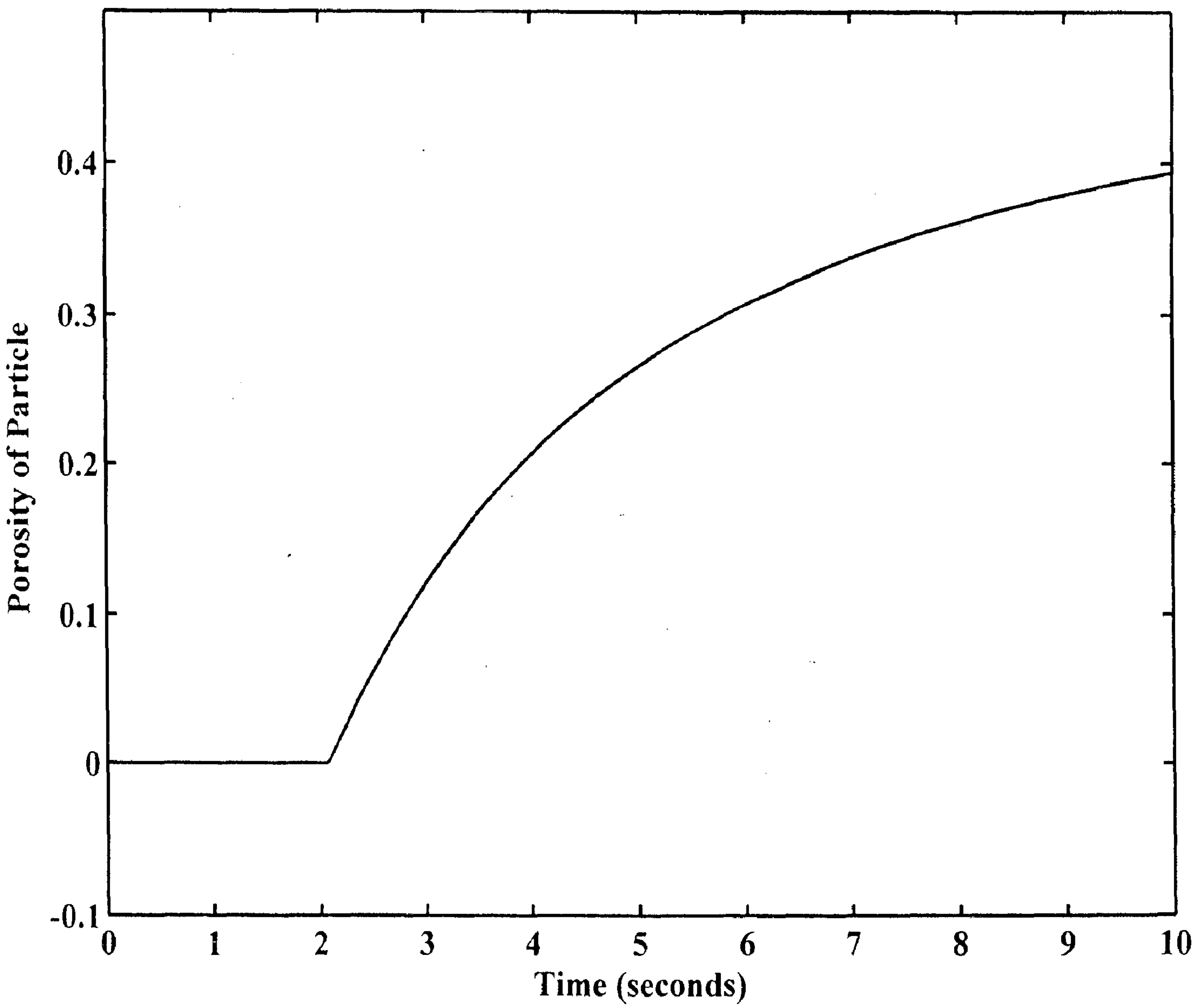


Fig 6.35 Porosity Profile for Different Feed Temperatures

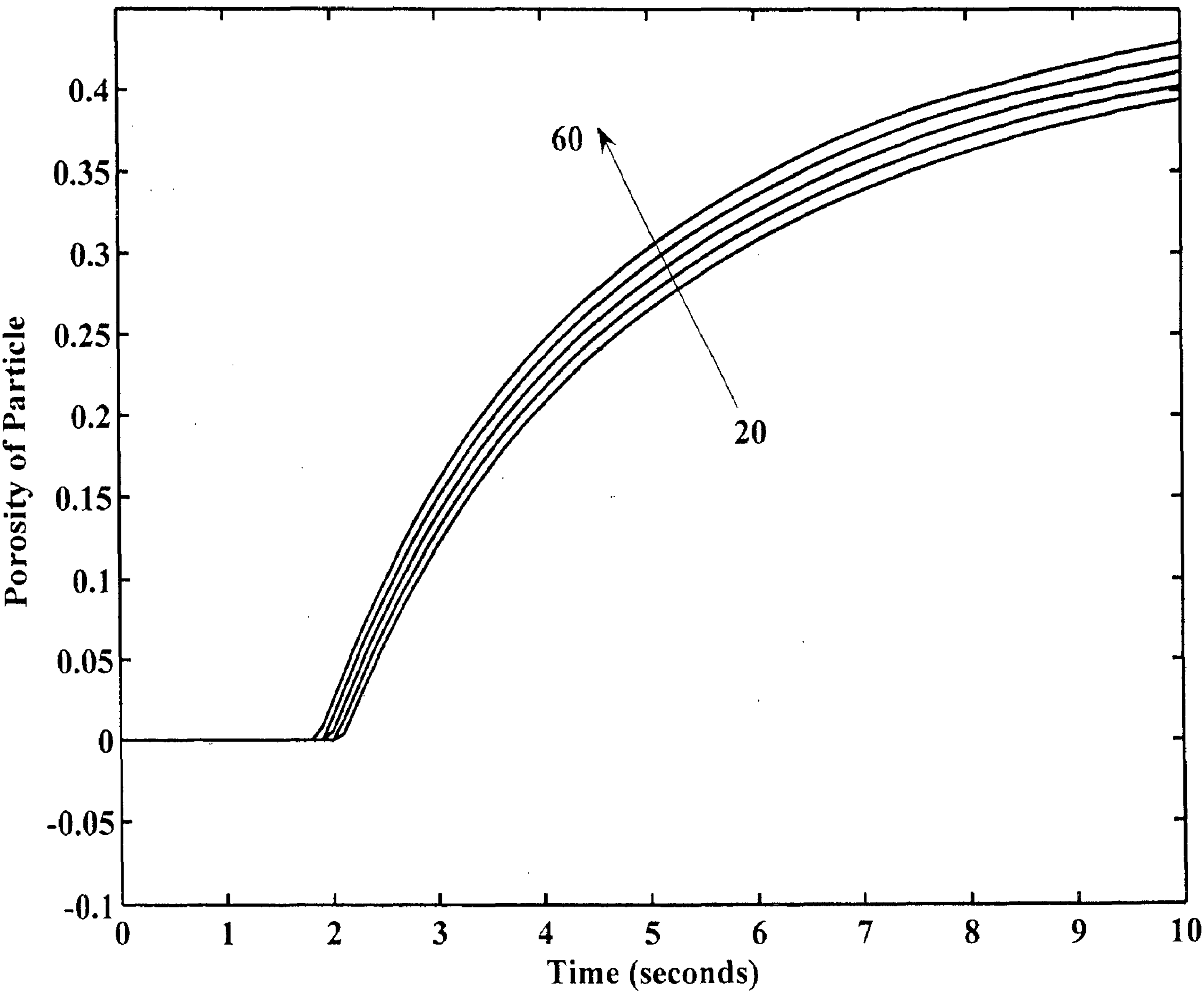


Fig 6.36 Porosity Profile for Different Drying Medium Humidities

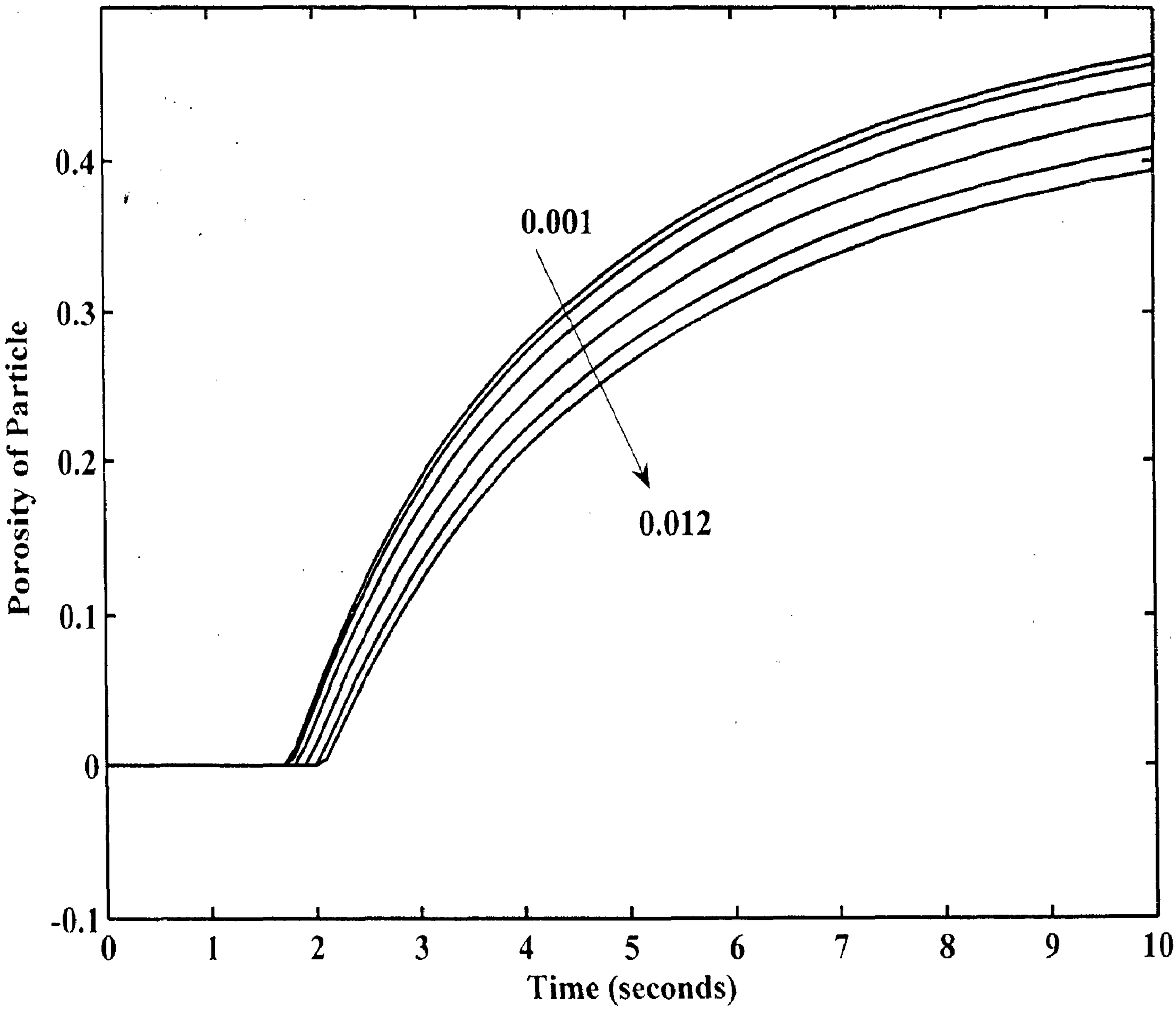


Fig 6.37 Porosity Profile for Different Drying Medium Temperatures

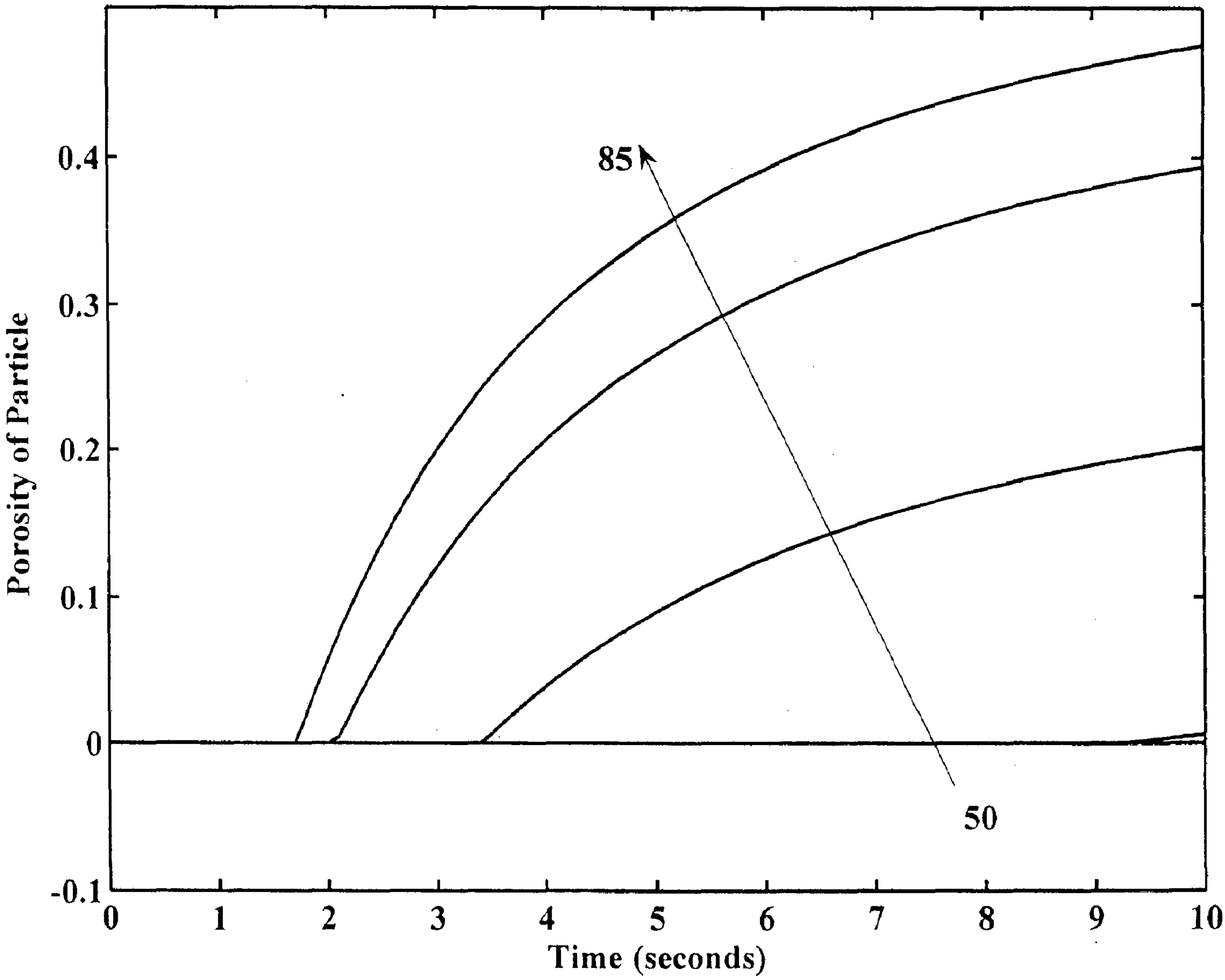
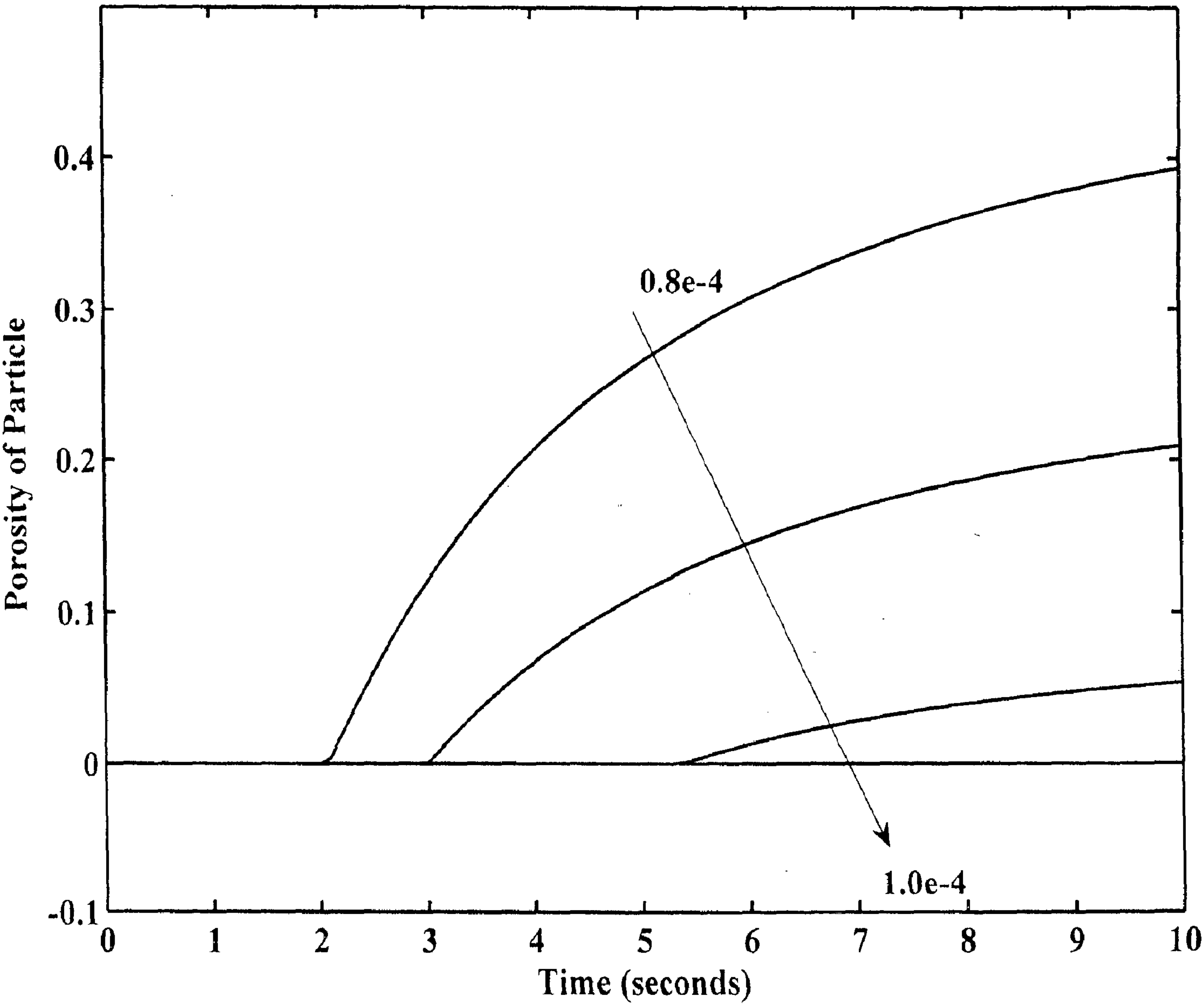


Fig 6.38 Porosity Profile for Different Drop Sizes



CONCLUSIONS AND RECOMMENDATIONS

7.0 Introduction

In this chapter we have discussed concluding remarks and recommendations briefly.

7.1 Conclusions

- A mathematical model for co-current spray dryer has been developed for the spray drying of skim milk.
- Mathematical model consists of a set of coupled ordinary differential equations, which constitute initial value problem. So these are solved in MATLAB 7.0.
- A steady-state simulation has been carried out for the detailed analysis of co-current spray dryer. Effects of different initial parameters on the performance of spray dryer operation have also been studied.
- A polynomial falling rate function has been developed on the basis of data available in the literature. This function is used to estimate the behavior of various parameters for falling rate period.

7.2 Recommendations for future work

- The results obtained from the work in the laboratory are some times quite different from those obtained at the industrial scale. It is therefore recommended that the model developed here to be tested for the data from the industries. This will enhance the applicability of model.
- In this model we did not consider the nozzle zone and the air hydrodynamics. It is recommended that the physics of the nozzle zone and air hydrodynamics should be incorporated into the model for better performance.

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