MASS TRANSFER OF NAPHTHALENE FROM AN IMMERSED SURFACE TO A SAND-AIR FLUIDIZED BED

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Master of Science
in
Chemical Engineering

by

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Summary

Fluidization process is widely used by a great assortment of industries worldwide and represents a trillion dollar industry [6]. They are currently used in separation, classification, drying and mixing of particles, chemical reactions and regeneration processes; one of these processes is the mass transfer from an immersed surface to a gas fluidized bed.

Mass transfer coefficient is a diffusion rate constant that relates the mass transfer rate, mass transfer area, and concentration gradient as driving force. It can be estimated from many different theoretical equations, correlations, and analogies that are functions of material properties.

A Q.V.F. glass column was used in this investigation, with 8 cm inside diameter and 70 cm in long. An empirical correlation was developed for mass transfer of naphthalene vapor into air-sand fluidized bed by experimental studying of many variables. These variables are: temperature (39-66℃), air velocity (1.2-1.4 U_{mf}), and sand particle size (215,165,112.5 μm).

$$Sh = Sh_e + C_1\left[\frac{(\rho_p - \rho)\rho g\,(\psi d_p)}{(G - G_{mf})^2}\right]^{C_2}\left[\frac{(\psi d_p)(G - G_{mf})}{\mu}\right]^{C_3}$$

In which $C_1 = 16.8574$, $C_2 = 0.07497$, and $C_3 = 0.1284$.

With an average absolute error = ± 1.54 %.

Where Sherwood number is: * Inversely proportional with particle size.
*Directly proportional with fluid flow.
* Inversely proportional with temperature.

A mathematical model for mass transfer from an immersed surface to a mass capacity fluidized bed was presented and compared with the experimental results of this work and previous works.

List of Contents

Contents **Page**

Abstract I

List of Contents 5

Notations V

List of Tables X

List of Figures XI

Chapter One: Introduction 1

Chapter Two: Literature Survey

2.1 Fluidization 3

2.1.1 Merits and Demerits of Fluidization Technique 4

2.1.2 Types of Fluidization 5

2.1.3 Types of Gas Distributor 5

2.1.4 Minimum Fluidizing Velocity 6

2.2 Factors Affecting the Rate of Mass Transfer 9

2.3 Mass Transfer from Particle to Gas Stream 11

2.4 Mass Transfer in Fixed Bed 12

2.5 Mass Transfer in Fluidized Bed 14

2.6 Mass Transfer from an Immersed Object to Fluidizing Bed 17

Chapter Three: Model of Mass Transfer from an Immersed Surface to a Mass Capacity Fluidized Bed

3.1 Introduction

3.2 Theories of Mass Transfer from an Immersed Surface 23
to a High Mass Capacity Fluidized Bed 25

3.2.1 Modified Packet Theory 25

3.2.2 Simplified Packet Theory 29

Chapter Four: Experimental Work

4.1 General Description of Flow System 34
4.1.1 Air Compressor 34
4.1.2 Air Flow Meter 34
4.1.3 Fluidization Column 35
4.1.4 Pressure Manometer 35
4.1.5 Bed Material 35
4.1.6 Immersed Work piece 35
4.1.7 Heating Equipment 36
4.1.8 Temperature Measurement Device 36
4.2 Experimental Procedure 39
4.2.1 Determination of Minimum Fluidizing Velocity 39
4.2.2 Mass Transfer in Empty Bed 39
4.2.3 Mss Transfer in Fluidized Bed 40
4.3 Correlation of Experimental Results 41

Chapter Five: Results and Discussion

5.1 Correlation of Results 45

5.2 Comparison of |Experimental Results with Previous Works and model 47

Chapter Six: Conclusions and Recommendations for Future work

6.1 Conclusions. 63

6.2 Recommendations for Future Work. 64

References 65

Appendixes

Appendix A Minimum Fluidization.

Notations

Symbols		Notations	Units
a_s	=	Proportionally factor	m^2/ s
b_s	=	Packing factor	
C_b	=	Bulk concentration.	mole / m^3
C_f	=	Specific heat of gas	J / kg. K
$C_m , \bar{C}_m$	=	Relative and mean relative mass capacity respectively	
C_{mf} ,C_{ms}	=	Specific mass capacity of gas and particles respectively	kg / kg
C_s	=	Concentration at the surface.	mole /m^3
C_p	=	Carbon potential of carburizing gaseous	-
d	=	Diameter of the bed.	m
d_p	=	Fluidizing particle diameter.	m
D , D_b	=	Effective diffusivity iv a packet and in a bubble respectively	m^2 / s
D_f	=	Molecular diffusivity in a gas	m^2 / s
$D_s , \bar{D}_s$	=	Effective and mean effective diffusivity in a particle respectively	m^2 / s
D_t	=	Transverse dispersion coefficient.	-
D_v	=	Diffusivity of transferable component.	m^2 /s
D_{v0}	=	Diffusivity of transferable component at 0 ^{0}C.	m^2/s
e	=	Void fraction.	-
e_{mf}	=	Minimum void fraction.	-

Symbols		Notations	Units
f_o	=	Hydrodynamic parameter	-
Fr	=	Froude number. $u^2 / d_p g$	-
g	=	Gravitational force.	m/s^2
G	=	Gas mass velocity.	$kg /m^2. s$
G_{mf}	=	Gas mass minimum velocity.	$kg/m^2. s$
h	=	Heat transfer coefficient.	$J /s.m^2.K$
h_f	=	Surface-to-gas heat transfer coefficient	$W /m^2. K$
j_d	=	Mass transfer factor.	-
k_{xp}	=	Surface-to-packet mass transfer coefficient related to $1-X_0/X_w$	$kg / m^2. s$
k_y	=	Mass transfer coefficients related to Y_w-Y_0	$kg / m^2.s$
k_y'	=	Surface-to-inert bed mass transfer coefficient	$kg / m^2.s$
k_{yb}'	=	Surface-to-bubble mass transfer coefficient	$kg / m^2.s$
k_{yf}	=	Surface-to-gas mass transfer coefficient	$kg / m^2.s$
k_{yp}	=	Surface-to-packet mass transfer coefficient	$kg / m^2.s$
k_{yp}'	=	Surface-to-packet mass transfer coefficient	$kg / m^2.s$
k_{ys}	=	Gas-to-particle mass transfer coefficient	$kg / m^2.s$
$k_{\theta p}$	=	Surface-to-packet mass transfer coefficient related to $\theta_w - \theta_o$	$kg / m^2.s$
Kg	=	Mass transfer coefficient.	m/s
Kg_e	=	Mass transfer coefficient between the surface and fluid.	m/s
m_s	=	Particle mass $m_s = \pi d_s^3 \rho_s / 6$	kg

M = Molecular weight. g/mole

M_s = Mass capacity of particles m^3 / kg

N = Total surface-to-bed mass flux $kg/m^2.s$

N' = Connective component of N $kg / m^2.s$

N_p, $N_{p\tau}$ = Time-averaged and instantaneous mass flux respectively $kg / m^2.s$

Nu = Nuselt number. $h\ d_s / k$ -

P_i = Partial pressure at the surface. mm Hg

P_s = Saturation partial pressure. mm Hg

R_{mp}, R_{mw} = Mass transfer packet and contact resistance respectively $m^2. s /kg$

Re_p = Reynolds number based on the diameter of the inert particles. $\rho u d_p / \mu$ -

Sc = Schmidt number. $\mu / \rho D_v$ -

Sh = Sherwood number. $k_g\ d_s / D_v$ -

Sh_e = Sherwood number in empty bed. -

t = Temperature. Ĉ

t_0,t_w = Bed and immersed surface temperature respectively Ĉ

t_{ref} = Reference temperature. Ĉ

T_s = Surface temperature. K

u = Superficial gas velocity m/ s

U = Gas velocity. m/s

U_{mf} = Minimum fluidizing velocity. m/s

X = Concentration of solid(mass of transferred substance per unit mass of inert solid) kg/kg

X_0 = Particle concentration in bulk medium of a bed kg/kg

X_r = Local concentration in a particle kg/kg

$X_{r=ds/2}$ = Concentration at a particle surface kg/kg

X_w = Particle concentration in equilibrium with an immersed surface kg/kg

Y = Concentration of gas (mass of transferred substance per unit mass of inert gas kg/kg

Y_0 = Gas concentration in bulk medium of a bed kg/kg

$Y_{r=ds/2}$ = Gas concentration in equilibrium with $X_{r=ds/2}$ kg/kg

Y_w = Gas concentration in equilibrium with an immersed surface kg/kg

Greek Letters

ΔP_b = Bed pressure drop. mm Hg

Δp_d = Distributor pressure drop. mm Hg

Δp_t = Total pressure drop. mm Hg

μ = Viscosity. kg/ s.m

μ_o = Viscosity at 0 ^{0}C. kg/s.m

ρ = Gas density. kg/m^3

ψ = Sphericity. $\pi d_p^2 / a_{ext}$ -

ρ_f, ρ_s	=	Gas and particle density respectively	kg/m^3
δ	=	Carbon transfer coefficient	m/s
α	=	Parameter of the simplified packet model	-
λ_{mp}	=	Effective mass conductivity for a packet	kg / m.s
θ	=	Mass transfer potential (X/X_w)	Potential unit
θ_o	=	Potential in bulk medium of a bed	Potential unit
θ_w	=	Potential at an immersed contact time and its mean value respectively	Potential unit
τ	=	Time	s
$\tau_b, \overline{\tau}_b$	=	Bubble residence contact time and its mean value respectively	s
$\tau_p, \overline{\tau}_p$	=	Packet residence contact time and its mean value respectively	s
ε_p	=	Packet porosity	-

Subscripts

b	Bubble
e	Equilibrium
f	Gas
m	Mass (minimum)
0	Bulk medium of a fluidized bed
s	Solid(particle)
w	Immersed surface

List of Tables

Table	Title	Page
(4.1)	Properties of Sand Particles**.**	33
(4.2)	Conditions of Mass Transfer	36
(5.1)	Comparison of the Orders of Magnitude of The experimental Parameters.	50
(5.2)	Selected of Data Sheet for Experiment of Mass Transfer in Empty Bed, (for experiment no. 1).	52
(5.3)	Selected of Data Sheet for Experiment of Mass Transfer in Empty Bed, (for Experiment no. 7).	53
(5.4)	Selected of Data Sheet for Experiment of Mass Transfer in Empty Bed, (for experiment no. 15).	54
(5.5)	Selected of Data Sheet for Experiment of Mass Transfer in Fluidized Bed, (for experiment no. 1).	55
(5.6)	Selected of Data Sheet for Experiment of Mass Transfer in Fluidized Bed, (for experiment no. 7).	56
(5.7)	Selected of Data Sheet for Experiment of Mass Transfer in Fluidized Bed, (for experiment no. 16).	57

List of Figures

Figure	Title	Page
(2.1)	Types of Fluidization	5
(2.2)	Types of Gas Distributor	6
(2.3)	Minimum Fluidizing Velocity	7
(2.4)	Air-Velocity Bed Pressure Drop for Silica Sand (375 micron)	9
(2.5)	Pressure Drop Across Fixed and Fluidized Bed as a Function of Fluid Velocity	13
(2.6)	(j_d) Factor for the Transfer of Naphthalene Vapor to Air in Fixed and Fluidized Beds	16
(2.7)	Relationship Between j_d and Reynolds Number of the Particles in Packed and Fluidized Beds	21
(4.1)	Diagram of the Experimental Equipment	37
(4.2)	Photographic Picture of the Experiment	38
(5.1)	Comparison of Experimental Data with The Packet Theory Systems.	48
(5.2)	Correlation for Mass Transfer in Empty Bed.	55
(5.3)	Experimental Sh. vs. Temperature at Air Flow Rate = 1.2 Umf.	55
(5.4)	Experimental Sh. vs. Temperature at Air Flow Rate = 1.4 Umf.	56
(5.5)	Experimental Sh. vs. Air Flow rate at Temperature = 51 ^{0}C.	56

(5.6) A Comparison of Equation 4.9 With the Experimental Data. 57

Figure **Title** **Page**

(5.7) A Comparison of Equation 4.10 With the Experimental Data. 57
(5.8) Effect of Temperature on Calculated Sh. No. At Particle Size 215 microns. 58
(5.9) Effect of Temperature on Calculated Sh. No. At Particle Size 165 microns. 58
(5.10) Effect of Temperature on Calculated Sh. No. At Particle Size 112.5 microns. 59
(5.11) Effect of Air Flow Rate on Calculated Sh. No. At Particle Size 215 microns. 59
(5.12) Effect of Air Flow Rate on Calculated Sh. No. At Particle Size 165 microns. 60
(5.13) Effect of Air Flow Rate on Calculated Sh. No. At Particle Size 112.5 microns. 60
(5.14) Effect of Sand Particle Size on Calculated Sh. No. at Temperature = 39 Ċ. 61
(5.15) Effect of Sand Particle Size on Calculated Sh. No. at Temperature = 51 Ċ. 61
(5.16) Effect of Sand Particle Size on Calculated Sh. No. at Temperature = 55 Ċ. 62
(5.17) Effect of Sand Particle Size on Calculated Sh. No. at Temperature = 66 Ċ. 62

Chapter One

Introduction

Fluidization is an important chemical process, and it use opened a wide range of possibilities for improving various industrial technologies especially operations that deal with mass transfer **[1]**.

There were so many studies carried out by various workers on the gas-solid fluidized bed operations field, to study the mass transfer and to find the best conditions of carrying them out, and to find out the effects of the large number of variables on the rate of mass transfer. This led researchers to suggest various modes that control the rate of mass transfer and its behavior. Accordingly, it can be seen that there is a large disagreement between them and that can be seen clearly from the combination of these results that there is a high percentage of disagreement in the results **[2]**.

In early applications, the fluid flowed through a static bed of granules supported on a grid. Provided the material is suitable, great improvement in mixing and contact is achieved if the granule size is properly matched to the upward velocity of the fluid. If they are matched well, the particles of material will be supported by the drag forces. When this occurs, the bed is said to be "fluidized" **[3]**.

Gas solid fluidization has a wide range of industrial applications like catalytic reactions, combustion, gasification, fluid bed coating of wall or dipped surface, dehumidification of large surfaces, granulation, and drying of wood etc.. In a number of these applications, the feed is not of uniform size and also there could be reduction in size due to attrition and during operation. These lead to entrainment and also limitation of operating velocity **[4]**.

The transfer rate, mass and heat transfer, from an object to a gas stream affected by the thickness of the boundary layer around the objects, so the transfer rate can be increased by immersing the object in a fluidized bed in which the solid particles will disrupt the boundary layer.

A number of correlations for mass transfer in fluidized beds have been proposed. Most of these involve a single-line relationship between the Reynolds number and the product of Sherwood number multiplied by some power of the Schmidt number. Earlier correlations define the Sherwood and Schmidt numbers arbitrarily; therefore the variables in bed geometry are not properly account for. Recent correlations are in good agreement with most of the reliable liquid-solid data reported thus far. For gas-solid systems, however, they leave much y to be desired **[5]**.

A study of the gas-solid system presupposes knowledge of all factors involved, such as fluid properties, fluidized properties, and nature of the flow, as well as the effect of each on the others.

The objective of this work is: determine the dependence of mass transfer coefficient on fluidized bed variables. The mass transfer coefficient can be predicted from the knowledge of mass transfer coefficient in the absence of fluidizing particles, plus a term that describes the effect of fluidizing solid particles on transfer rate coefficients.

Chapter Two
Literature Survey

2.1 Fluidization

Fluidization can be defined as the process by which solid particles are transformed into a fluid like-state via suspension in a fluid (gas or liquid) **[6]**.

If a fluid passes upwards through the bed, pressure drop across the bed will be directly proportional to the rate of gas at low flow rates. But when the fractional drag on particles becomes equal to their apparent weight particles rearrange so that they offer less resistance to the flow fluid and the bed starts to expand **[7]**.

If the velocity is increased still further, the individual particles separate from one another and become free supported in fluid and the bed is said to be fluidized. Further increase in the velocity causes the particles to separate still further from one another and that results in different bed behavior depending on whether the fluidizing agent is a liquid or gas **[8]**.

Fluidized bed processes can be broadly characterized in to two types: physical or chemical. Physical processes are concerned with heat or mass transfer such as the drying of solids. Chemical processes deal with catalytic or transformation reactions such as combustion processes or catalytic cracking. The immersed bodies can be classed as a physical fluidized bed.

Kaneko *et al.*, 1999**[9],** and Rhodes *et al*, 2001**[10],** Kafui *et al.*, 2002 **[11],** studied the general characteristics of a fluidized bed, such as the gradual change in particle characteristics and size distribution in the bed, and also studied the impact of inter particle forces on fluidization.

The size of solid particles which can be fluidized varies greatly from less than 1 micron to 2.5 inch. It's generally concluded that particles distributed in size between 65 mesh and 10 microns are the best for smooth fluidization (least formation of large bubbles) **[12]**.

2.1.1 Merits and Demerits of Fluidization Technique

In this work the advantages and disadvantages of the technique of fluidization was summarized **[1]**:

The advantages are:

1- Ease of handling.

2- Nearly isothermal behavior due to excellent solids mixing.

3- High heat and mass transfer rates.

4- Excellent suitability to large-scale operations.

The disadvantages are:

1- Fluid throughput rates are limited to the range over which the bed is fluidized. If the velocity is much higher than U_{mf}, there can be excessive loss of material carried out from the bed and there may also be acceptable particle damage due to excessive operating velocity.

2- The pumping power supplied to fluidize the bed can be excessive for very large, deep beds.

3- Size and type of particles which can be handled by this technique are limited.

4- Due to the complexity of fluidized bed behavior, there are often difficulties in attempting to scale-up smaller scale to industrial units.

2.1.2 Types of Fluidization

Depending of the type of material being fluidized, six distinct fluidization types can occur. These are, homogeneous, bubbling, slugging, turbulent, jetting and spouting as described by Kunii and Levenspiel 1991**[13]**. Show Fig. (2.1).

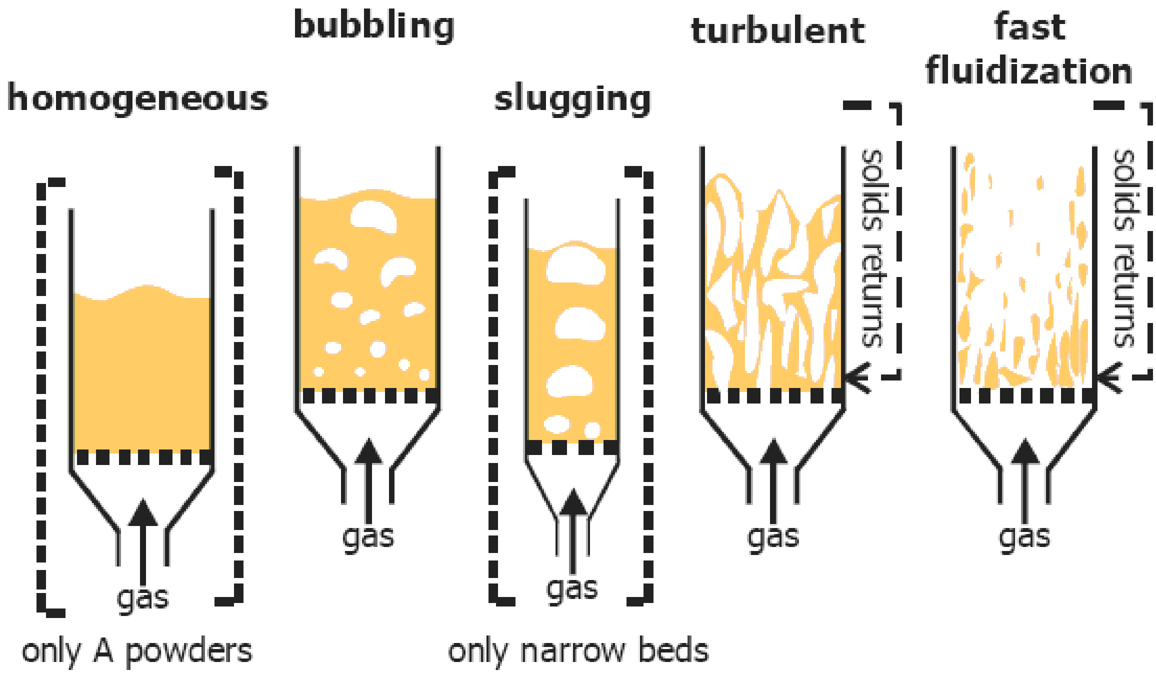
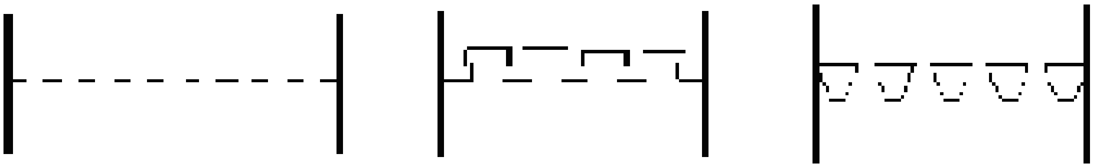


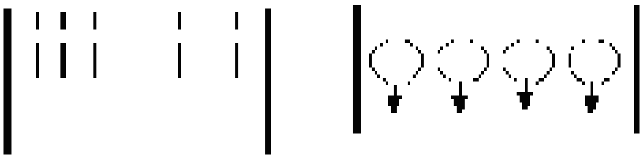
Figure (2.1): Types of Fluidization **[13]**

2.1.3 Types of Gas Distributor

The distributor is the devise designed to insure that the fluidizing gas is always evenly distributed across the cross-sectional of the bed. It's a critical part of the designed of fluidized bed system. Good design is based on achieving a pressure drop which is a sufficient fraction of the bed pressure drop. Many operating problems can be traced back to poor distributor design. Some distributor designs in common use are shown in Fig. (2.2)**[14]**.



(a) drilled plate (b) cap design (c) continuous horizontal slots



(d) standpipe design (e) spurge tube with holes pointing downwards

Figure (2.2): Types of Gas Distributor **[14]**

2.1.4 Minimum Fluidizing Velocity

When the gas is passed upwards through a fluidized bed unrestrained at its upper surface, the pressure drop increases with gas velocity as shown in Fig.(2.3), the drag on an individual particle excess the force exerted by gravity, the pressure drop across the bed equals the weight of the bed per unit area. Then an excess pressure is required to free the particles that are interlocked at the fluidized state and theoretical pressure drop.

The velocity at the point that the pressure drop falls back is called the minimum fluidizing velocity (U_{mf}) **[15].**

From Carman and Kozeny for spherical particles:

$$U_{mf} = 0.0055 \frac{e_{mf}^3}{(1-e)} \frac{\rho_s - \rho}{\mu} d^2 g \qquad \text{--- (2.1)}$$

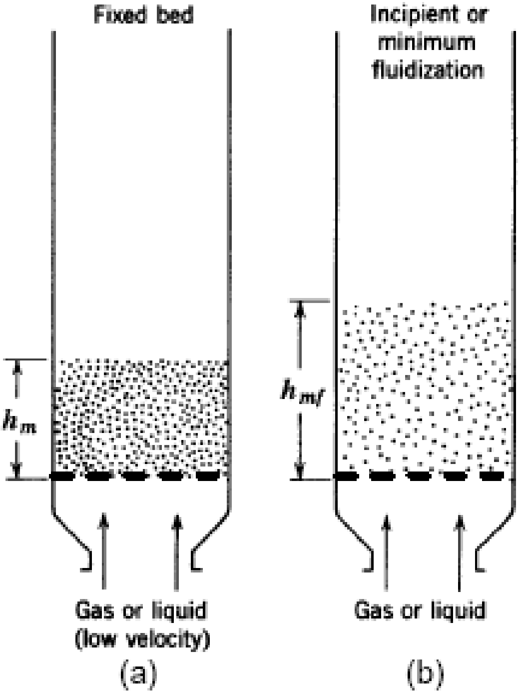


Figure (2.3): Minimum Fluidizing Velocity **[15]**

Leva, 1959**[16]**, worked with round and sharp sands of 0.05-0.40 mm using 0.1 m diameter with various depths fluidized by air. He noted that the smaller particles require an extra of energy for fluidization.

Davidson *et al.*, 1966**[17]**, has shown that the U_{mf} is a function of the square of particle diameter; as a result, the quantity of air required for the U_{mf} changes as the products particle size change.

Wen and Yu, 1966**[18],** produced an empirical correlation for U_{mf} for gas fluidization the Wen and Yu correlation is often taken as being most suitable for particles larger than 100 μm, where as the correlation of Baeyens in 1974 **[19]**, shown below in equation (2.2), is best for particles less than 100μm.

$$U_{mf} = \frac{d_p^{1.8}(\rho_p - \rho)^{0.934} g^{0.934}}{110\mu^{0.87}\rho^{0.066}} \qquad \text{--- (2.2)}$$

Arai *et al.*, 1974**[20]**, studied experimentally the effect of bed height on the behavior of gas-solid fluidized bed of gas beads and silica sand, a gas column was charged with different particles weight range from (120-300)g. They concluded that the effect of bed height on the U_{mf} may be neglected except when it is very low.

Grace, 1982**[21]**, found a correlation to predict the minimum fluidization velocity for gas-solid systems:

$$\frac{\rho u_{mf} d_p}{\mu} = \left(C^2 + 0.0408Ar\right)^{1/2} - C \qquad \text{--- (2.3)}$$

Where *Ar* is the Archimedes (or Galileo) number that equal to $Ar = \rho g d_p^3(\rho_p - \rho)/\mu^2$ and *C* is the constant which equal to 27.2 and others using the value 33.4 and 33.7. It would seem reasonable to use a mean value of 30.

Gupta and Sathiyamoorthy, 1999**[22]** and Kunii and Levenspiel, 1991**[13]**, studied the pressure drop vs. velocity relationships during fluidization and defluidization. They plotted curves for different sand sizes for increasing flow rates (fluidization) and decreasing flow rates (defluidization). In a typical curve showing the variation of bed pressure drop with superficial air velocity for the sand size of 375 microns is shown in Fig. (2.4), they showed that the pressure drop was higher on fluidization than on defluidization. It was evident because initially the bed was fewer perms.

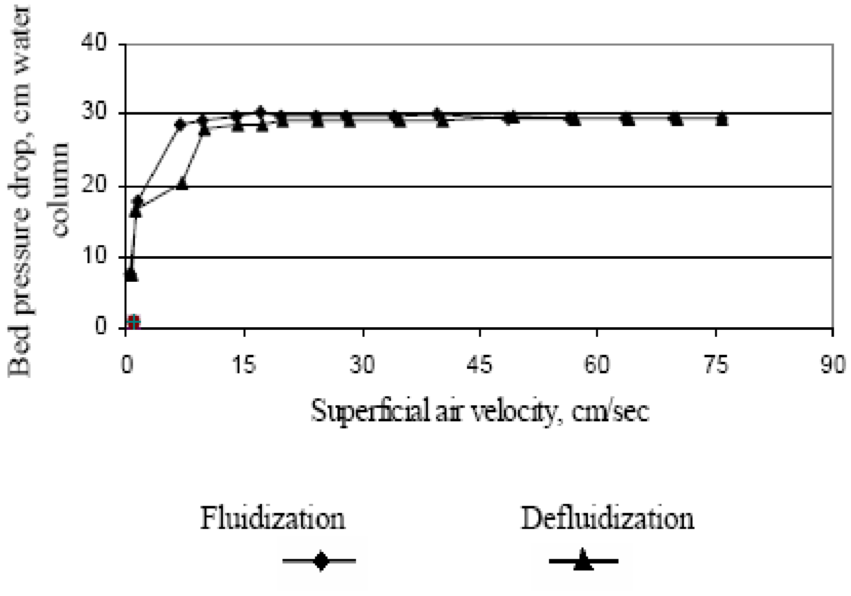


Figure (2.4): Air Velocity-Bed Pressure Drop
for Silica Sand (375 microns)

2.2 Factors Affecting the Rate of Mass Transfer

Markova *et al.*, 1965**[23]**, studied the effect of fluidized particle size on mass transfer coefficient with particle diameter of 0.565, 0.488 and 0.347 mm and they concluded that the dependence of mass transfer coefficient on the size of particle is worthy of note, decreasing the particle size will increase the coefficient. They also studied the effect of gas velocity on the mass transfer coefficient using a glass apparatus with a diameter of 0.15m and 0.05m height. The bed temperature was 50 C̊. They concluded that increasing the air velocity will increase the mass transfer coefficient.

Vanecek *et al.*, 1966**[24]**, found that the effect of particle size is small but a diffusion process is largely affected by the particle size.

Kim, 1995 **[25]**, studied possibilities of improving substantially the hydrodynamic behavior and the mass transfer characteristics in a fluidized bed, thus increasing the gas loading capacity of the fluidized bed column , reducing the specific pressure drops and increasing the mass transfer rate. The object of the investigation is achieved in that the fluidized bed bodies forming the fluidized bed obtain a different geometric shape than is hitherto customary. Unlike the hitherto customary cone shapes or ellipsoidal shapes, fluidized bed bodies are used whose shape has a characteristic asymmetry. He found that the change of fluidized beds bodies such as (ellipsoidal, cone, hens egg-shaped) have the following advantages:

1-Reduction in the column volume in height and diameter.
2-Lower specific pressure drops, particularly at high gas rates.
3-Use of the fluidized bed bodies for mass transfer processes (absorption, dsorption, dust separation, and catalytic reactions in the fluidized bed).

Wenyuan Wu. *et al.*, 2002**[26]**, studied the effect of gas temperature on the mass transfer coefficient .They found that:

1- Changing the solubility of substances. It may increase or decrease the solubility and thus increase or decrease mass transfer rates.

2- Increasing the rate of molecular diffusion thus increasing the rate of movement of molecules through boundary layers.

3- Decrease of mass transfer coefficient with decreases the bed temperature, and increase with increasing particle size.

Nevenka *et al.*, 2004**[27]**, studied experimentally the influence of fluid velocity and particle size, on the mass transfer in packed beds and fluidized beds. They found:

- The mass transfer in the presence of particles is more intensive, hence the values of mass transfer coefficient are greater in both the two - phase systems (packed and fluidized bed) than the fluid flow around the single sphere. When the fluid flows around a single sphere only one part of the area is exposed to transfer. In the presence of particles the whole area become active and the boundary layer become thinner, thus increase the mass transfer.
- With increasing interstitial velocity in the fluidized bed, the mass transfer coefficient decreases slightly reaching the value of the mass transfer coefficient for flow around a single sphere.

2.3 Mass Transfer from Particle to Gas Stream

Methods of predicting the evaporation rate of single drop (particles) and the phenomena associated with the evaporation process are of importance in the analysis of chemical engineering operations involving dispersions in gas **[28]**.

Any body immersed in a fluid is surrounded by a boundary layer. A boundary layer is simply a region where the movement of molecules is slow and determined solely by the diffusion force **[29]**. Boundary layer theory predicts that the rate of transfer is a maximum on the front side of the drop (particles) facing the on coming gas stream, decreasing to a minimum value near the separation point, and increase to another but lower maximum rate on the trailing face which experiences velocities in the reverse direction. Such distribution of mass transfer rate is shown by Frossling for the sublimation of naphthalene bead **[30]**.

The evident that the rate of mass transfer will be largest on the side facing the gas stream where the concentration gradient is steepest. Since the physical situation in finely dispersed systems is such the relative velocity and Reynolds number become vanishingly small.

The value of *Sh* and *Nu* number in this limiting condition at zero Reynolds number (theoretical minimum value) is **[28]**:

$Sh=Nu=2.0$

For finite velocities, experimental data on mass-transfer rate for spheres may be correlated by an empirical equation of the form used by Frossling.

$$Sh = 2.0 + C_1 (Sc)^{C_2} (\mathrm{Re}_p)^{C_3} \qquad \text{--- (2.4)}$$

Where $C_1=0.6$, $C_2=1/3$, $C_3=1/2$.

At high values of *Re No.*, the constant term becomes less significant so that equation (2.4) may be converted to the familiar *J-factor* equation of Chilton and Colburn.

2.4 Mass Transfer in Fixed Bed.

If the gas is pumped upwards through a granular bed at low flow rate the fluid percolates through the pores with no perturbation of the bed height, it's called a fixed bed.

For fluid-solid reactions, the pressure drop for flowing through the fixed bed is higher than for a fluidized bed with a same flow rate. A schematic of the pressure drop versus flow rate is shown is shown in Fig. (2.5). At low fluid velocities the pressure drop is approximately liner with flow rate, this is expected behavior for packed beds. After achieving incipient fluidization increasing the fluid flow velocity doesn't result in any significant increase in the pressure drop as the bed expands to reduce resistance to flow **[1]**.

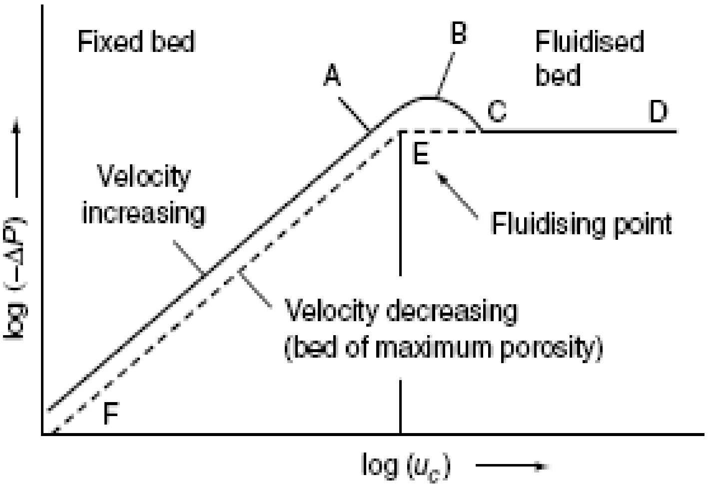


Figure (2.5): Pressure Drop across Fixed Bed as a Function of Fluid Velocity

Mass transfer studies with beds of coarse solids show reasonable agreement for both flowing liquids and gases. Based on studies by Ranz 1952**[28]**, the finding may be summarized by:

$$Sh = 2 + 1.8 Sc^{1/3} \mathrm{Re}_p^{1/2} \qquad \text{--- (2.5)}$$

For

$$\mathrm{Re}_p = \frac{d_p u \rho}{\mu} > 80$$

Wakao and Funazkri, 1978**[31],** correlated the published mass transfer data, of particle to fluid mass transfer coefficient in packed bed, for the axial fluid effective dispersion coefficient.

The corrected Sherwood numbers in the range of Reynolds number from about 3 to 10000 are correlated by:

$$Sh = 2 + 1.1 Sc^{1/3} \mathrm{Re}_p^{0.6} \qquad \text{--- (2.6)}$$

They found that the reevaluated gas-phase data are considerably higher than those obtained under the assumption that the axial fluid effective dispersion equal to zero.

Gunn, 1987**[32]**, studied the mass transfer in gas-solid fixed and fluidized beds operating in a wide range of velocities and porosities. He developed a theoretical correlation that expresses the mass transfer between the particles and fluids processes.

2.5 Mass Transfer in Fluidized Bed

Quite a number of papers deal with transfer problem. A chaos of correlations, statements and conclusions are found in the literature, due to the fact that an increase fluidizing velocity increase the bed expansion, which in many cases is not measured. The consequence, that in the proposed correlations not only the influence of gas velocity, u, on the fluid to particle mass transfer coefficient, k_g, is hidden but also that of the void fraction, e, on k_g **[33]**.

Several correlations for the calculation of the Nusselt and Sherwood dimensionless numbers are reported. Since these correlations are mainly based on experimental investigations performed under laboratory conditions, they may be different to the situation in large scale reactors.

Hurt, 1934**[34]**, measured the height of transfer unit for gas film controlled system in fluidized beds with different sizes and shapes of packing. No spherical particles were used, but the cylindrical particle data showed close agreement between heat and mass transfer factors as Chilton and Colburn 1934 had suggested. No such agreement between heat and mass transfer data was obtained for other packing shapes. Hurt did not report fraction voids or surface area of the beds he used.

McCune and Wilhelm, 1949**[35]**, attempted to relate the mass transfer group ej_d with the modified Reynolds [d_pG/eu], but found that the group obtained from their data for fluidized beds did not correlate with the corresponding groups for packed beds.

Resnick and White, 1949 **[36],** calculated the mass transfer coefficient of naphthalene crystals of five different size ranges (250-1000microns) in air, hydrogen, and carbon dioxide at a temperature of 298K and rates between (0.01 and 1.5 $kg/m^2.s$). They expressed (J-factor) and plotted against Reynolds number as shown in Fig. (2.6).

Gamson, 1951**[37],** utilizing the available mass transfer data for packed and fluidized beds related the mass transfer modulus $[\, j_d /(1-e)^{0.2}]$ to the modified Reynolds group.

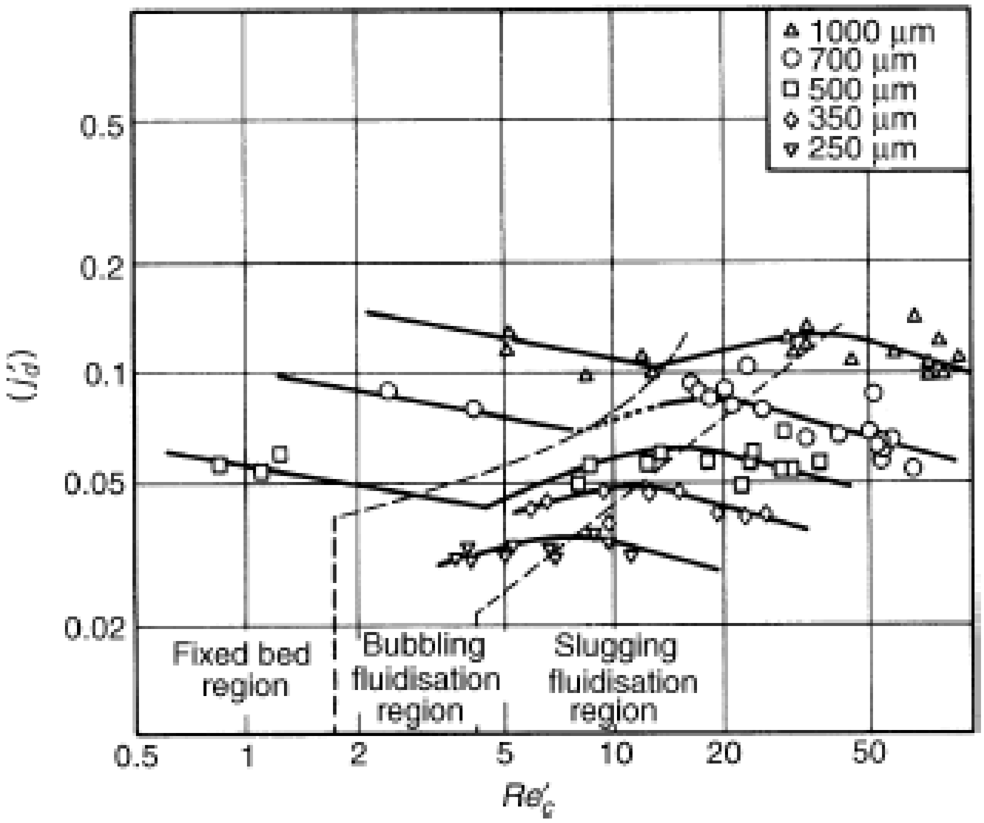


Figure (2.6): (j_d) Factor for The Transfer of Naphthalene Vapor to Air in Fixed and Fluidized Beds

Gupta and Thodas, 1962**[38],** attempted to correlate mass transfer factor with the conventional Reynolds number d_pG/u, utilizing all the available data. They found that the relationship representing the data best can be expressed by the equation:

$$ej_d = 0.01 + \frac{0.863}{\mathrm{Re}_p^{\,0.58} - 0.483} \qquad \text{--- (2.7)}$$

$\mathrm{Re}_p > 1$ For

In fluidized bed, the solid particles are sufficiently separated so that in effect there is mass transfer between a gas and single particles. The most widely used correlation for this purpose is the equation of Froessling for mass transfer to single sphere:

$$Sh = 2 + 0.6\,\mathrm{Re}^{1/2} Sc^{1/3} \qquad \text{--- (2.4)}$$

Mass transfer coefficient obtained from this relationship may then be combined with mass transfer among the various phases in the fluidized bed to yield the overall behavior with regard to the transport of mass. Owing to the small particle sizes and high surface area per volume of solids used in fluidized beds, the mass transfer from the gas to the solid surface is usually quite rapid, and consequently, it seldom limits the reaction.

2.6 Mass Transfer from an Immersed Object to Fluidizing Bed

Mass transfer can involve the movement of mass through a fluid, movement of mass through catalyst pores or cellular or movement of mass between phases. Mass transfer is an important consideration in downstream processing where mass must be moved between phases.

Shirai, Yoshitome and Shoji, 1966 **[39],** studied heat and mass transfer between fluidized bed and surface of single sphere fixed in the bed. For mass transfer study, sand was employed as fluidizing particles, the solid sphere is made of brick and the system used is air-water system. They found that the value of Sherwood number is only 1.5 times that for mass transfer between particles and fluid. They concluded that the mechanism of heat is not analogous to that for mass transfer, because heat may be transported by both gas and the fluidizing particles but mass can be transferred only by the gas phase, when the particles are non-adsorptive.

Ziegler and Holmes, 1966 **[40]**, studied mass transfer from fixed surface to gas fluidized beds, mass transfer coefficients were measured for the diffusion of water vapor from a saturated porous sphere into various air-fluidized beds of solid particles. Naphthalene diffusion from coated flat plate into fluidized beds was also studied. They found that for the case in which particle adsorption is negligible, the Sherwood number could be correlated by:

$$Sh = Sh_e + f(y) \qquad \text{--- (2.8)}$$

In which Sh_e is the predicted Sh forced convection in the absence of particles but at the increase Reynolds number of the fluidized bed, and $f(y)$ describes the effect of particle motion on transfer rate. They chose $f(y)$ as a power function of y, that is

$$\text{--- (2.9)} \quad f(y) = C_1 y^{C2}$$

From experimental data, the value of C_1 and C_2 found are 33.7 and 0.4 respectively with an average standard deviation of 16.5%.

For Naphthalene diffusion, unfortunately adsorption on the particles increased the transfer rate, for which reason data are inconsistent **[33]**.

Ciborowski and Kopec, 1985**[41],** reported previous works, for mass transfer from an immersed object to a gas fluidized bed, which can be summarized as follows:

1- The ratio of surface-to-inert bed mass transfer coefficient to that between the surface and a gas; (k_g / k_{ge}), varies from 1.1 to 10.
2- (k_g) is a slightly increasing function of the superficial gas velocity and for large velocities is independent of u.
3- (k_g)rises with an increase in fluidizing particle size, while h diminishes as d_p grows.

R. Joulie, *et al.*, 1986**[42]**, showed that the rate of sublimation in fluidized beds is far higher than in air alone. It increase with increasing bed temperature, decreasing particle size, increasing powder mass capacity, and roughly various as a parabolic function of time. They found that the temperature difference between the bed and the object surface, depends on the fines characteristics as well as on bed temperature, but is independent of gas velocity when good solid mixing conditions are achieved.

Choudhary *et al.*, 1976 **[43]**, Lerou and froment, 1977 **[44]**, Vortmeyer and Winter, 1982 **[7]**, and delmas and Froment, 1988 **[45],** studied the effect of radial variations of porosity and velocity on axial and radial transport of mass in packed beds.

R. Joulie, *et al.*, 1993**[46]**, studied the heat and mass transport phenomena between a large sublimely object and a gas-solid fluidized bed of either inert or adsorbent fine panicles. Heat (*h*) and mass transfer (*k*) coefficients are related to the size and adsorption capacity of the fine particles, as well as to the diameter of the spherical immersed object and the bed temperature. Convective and conductive components are also identified, and correlations are proposed to predict all of them. They stated that the significant differences which appear for the various working conditions may be accounted for considering the particular mechanisms that prevail in every case.

Guedes de Carvalho *et al.*, 1999**[47]**, found an equation for gas flow predict the coefficient of mass transfer from large active particles in beds of small inert. The equation is:

$$\frac{Sh'}{\varepsilon}=\left[4+\frac{4}{5}(Pe')^{2/3}+\frac{4}{\pi}Pe'\right]^{1/2}\left[1+\frac{1}{9}\left(\frac{d}{d_1}\right)Pe'\right]^{1/2} \qquad \text{--- (2.10)}$$

Philipp Schlichthaerle *et al.*, 2000**[48]**, studied the sublimation of large solid carbon dioxide particles inside fluidized beds of fine particles. A model which takes the surface area of the sublimely particles into account is used to describe the sublimation kinetics. Based on this model, the results of different experiments, namely single particle experiments using a precision scale, batch experiments in a laboratory-scale fluidized bed and continuous experiments in a larger circulating fluidized bed are compared. The main focus of the study is to evaluate the influences of the particle size, of the inert bed material, of the bed temperature and of the superficial gas velocity, respectively**.**

Delgado and Guedes de Carvalho, 2001**[49]**, showed that there is a significant dependence between D_t and Sc in the range $Sc<550$. Since the rate of mass transfer around an immersed sphere exposed to a flowing fluid is strongly determined by D_t , it may be expected that this mass transfer will show a significant independence on Sc.

Nevenka *et al.*, 2004**[27]**, studied the mass transfer coefficient between fluid and an immersed sphere in packed and fluidized beds of inert spherical particles experimentally using a column 40 mm in diameter and inert glass particles 0.5-2.98 mm in diameter.

They obtained a new relationship between j_d and Re number of the particles in packed and fluidized beds, that shown in Fig.(2.6). The data for the mass transfer factor (j_d) in fluidized bed can be separated into two groups, depending on the particle diameter. The data for each group fall in straight lines, the slopes of which are approximately the same but different from that for the packed beds. The mass transfer coefficient for packed and fluidized bed can be correlated with a single equation:

$$j_d \varepsilon = 0.64 \left(\frac{\mathrm{Re}_p}{1-\varepsilon} \right)^{-0.4} \qquad \text{--- (2.11)}$$

With a mean absolute deviation of (16.88 %) and a mean relative deviation of (- 3.71 %).

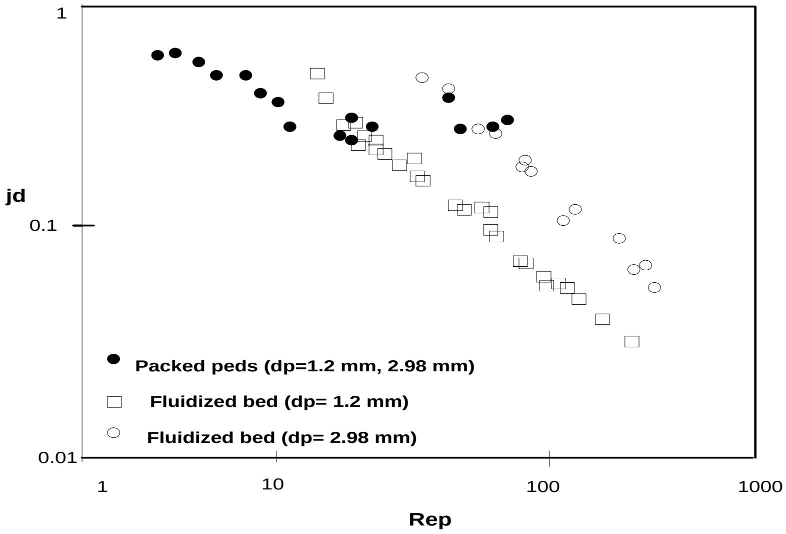


Figure (2.7): Relationship between j_d and Reynolds Number of The Particles in Packed and Fluidized Beds

Weimin GAO *et al.*, 2004 **[50]**, studied the mass transfer in fluidized bed. A steel work piece covered with carbon was used in their investigation. The carbon transfer coefficient was determined from the carbon distribution within the diffusion layer of the sample. An empirical relationship of the carbon potential as a function of carburizing atmosphere, bed temperature (-90°, -55°, -30°, 0°, 30°, 60°) and fluidizing velocity ($1.32U_{mf}$, $2.12U_{mf}$) was determined, based on the understanding of the mass transfer mechanism and analysis of the experimental results.

$$\delta = 1.5022\delta_o\left(1-f_B\right)\frac{1}{d_p}\mathrm{Re}^{0.8452}\left(1-\gamma\right) \qquad \text{--- (2.12)}$$

$$\delta_o = \left(2.54\times10^{-5} + 3.3\times10^{-7}C_p^{6.22}\right)\exp\left(-\frac{40.365}{RT}\right) \qquad \text{--- (2.13)}$$

$$\gamma = \exp\left(-\frac{8.805}{\rho_p\left(U - U_{mf}\right)^2}\right) \qquad \text{--- (2.14)}$$

Chapter Three

Model of Mass Transfer from an Immersed Surface to a Mass Capacity Fluidized Bed

3.1 Introduction

The process of mass transfer from an immersed surface to a gas fluidized bed has not yet been intensively investigated. Only a few papers on this process are available and the majority of them deal with non-adsorbing (inert) fluidized particles.

To describe the process mathematically Baskakov and Supurn**,** 1970 **[51]**, Markova, 1972**[52]**, and Prozorov, 1976 **[53],** assumption that mass is transmitted from the surface by packets of particles and by gas bubbles as follows:

$$k_{y'} = (1 - f_o) k_{yp'} + f_o k_{yb'} \qquad \text{--- (3.1)}$$

Where

$$f_o = \frac{\bar{\tau}_b}{\bar{\tau}_b + \bar{\tau}_p} \qquad \text{--- (3.2)}$$

In contrast to heat transfer theory **[54]**, where the heat within a packet is transferred through gas and particles and the accumulation of heat within particles plays a dominant role, these workers assumed that mass within a packet is transferred only by gas between particles occurs. Thus the mass transfer coefficient to a packet was found **[51, 52]** to be

$$k_{yp'} = 2\rho_f \left(\frac{D_p}{\pi \bar{\tau}_p} \right)^{1/2} \quad \text{--- (3.3)}$$

$$k_{yb'} = 2\rho_f \left(\frac{D_b}{\pi \bar{\tau}_b} \right)^{1/2} \quad \text{--- (3.4)}$$

It must be remembered that all the above considerations apply to an inert fluidized bed.

If adsorption of a transported substance onto the particles takes place the mass transfer coefficient rises **[40,51,55]** and the ratio (k_y/k_{yf}) may then reach values from 3 to 15 **[40,55].** For such cases, on the basis of the packet theory and allowing for mass accumulation on particles, Yokota, 1975**[55],** derived the following expression:

$$k_{yp} = \rho_f \left(\frac{\varepsilon_p D_f M_s \rho_s (1-\varepsilon_p)}{\bar{\tau}_p} \right)^{1/2} \quad \text{--- (3.5)}$$

Where $Sh_p = \frac{k_{yp} L}{D_f \rho_f}$ transformed into the dimensionless form:

$$Sh_p = \left(\frac{\varepsilon_p (1-\varepsilon) M_s \rho_s L^2}{D_f \bar{\tau}_p} \right)^{1/2} \quad \text{--- (3.6)}$$

For particles with large mass capacities, the mass transfer coefficients become higher and greater similarity between surface-to-bed mass and heat transfer mechanism.

In this chapter, the mass capacity process was investigated and described on the basis of the modified packet model including the mass contact resistance. For the contact resistance control region the alternative simplified packet model is developed.

3.2 Theories of Mass Transfer from an Immersed Surface to a Mass Capacity Fluidized Bed

For strongly adsorbing particles mass "conduction" into a packet may take place through both gas and particles. Thus some mass transfer potential (θ) common for the gaseous and solid phases should be used.

Under isothermal conditions the specific mass capacities will then be defined as follows:

$$C_{ms} = \left(\frac{\partial X}{\partial \theta}\right)_e \quad , \quad C_{mf} = \left(\frac{\partial Y}{\partial \theta}\right)_e \qquad \text{--- (3.7)}$$

For further manipulation it is convenient to define the relative mass capacity as:

$$C_m = \frac{C_{ms}}{C_{mf}} = \left(\frac{\partial X}{\partial Y}\right)_e \qquad \text{--- (3.8)}$$

This is the slope of an adsorption isotherm.

3.2.1 Modified Packet Theory

An assumption consider a packet of emulation which suddenly comes into contact with a mass exchange surface with potential (θ_w) and after a time (τ_p) is replaced by fresh packet from the core of the bed with potential (θ_o).

The different equation which describes non-steady-state mass transfer into the packet is:

$$\varepsilon_p \rho_f \frac{\partial Y}{\partial \tau} + \left(1-\varepsilon_p\right)\rho_s \frac{\partial X}{\partial \tau} = \frac{\partial}{\partial z}\left(\lambda_{mp}\frac{\partial \theta}{\partial z}\right) \qquad \text{--- (3.9)}$$

or, after the substitution of equation (3.7) into equation (3.9)

$$\left\{\varepsilon_p \rho_f C_{mf} + (1-\varepsilon_p)\rho_s C_{ms}\right\}\frac{\partial \theta}{\partial \tau} = \frac{\partial}{\partial z}\left(\lambda_{mp}\frac{\partial \theta}{\partial z}\right) \qquad \text{--- (3.10)}$$

The initial and boundary conditions for this situation are:

$$\theta(\tau = 0, z) = \theta_o \quad , \ \theta(\tau, z \to \infty) = \theta_o \ , \quad \theta(\tau, z = 0) = \theta_w \qquad \text{---(3.11)}$$

For simplest case of non-adsorbing ($C_{ms}=0$) and non-diffusion ($D_s=0$) particles and taking ($\theta=Y$), the effective mass conductivity for a pocket results of equation (3.10):

$$\lambda_{mp} = \varepsilon_p D_f \rho_f C_{mf} \qquad \text{--- (3.12)}$$

For adsorbing particles the first term in curly brackets in equation (3.10) can be neglected, since ($\rho_s C_{ms} >> \rho_f C_{mf}$).

When C_{ms} , C_{mf} and λ_{mp} are independent of θ an analytical solution of equation (3.10) is possible by assuming a complete analogy between heat and mass transfer within packet **[56]**. An expression for effective mass conductivity of a packet will then be identical to that given for effective thermal conductivity

$$\frac{\lambda_{mp}}{\lambda_{mf}}=\frac{(1-\varepsilon_p)(1-1/B)}{\frac{1}{B}+0.28\varepsilon_p^{0.63B^{0.18}}}+1 \qquad \text{--- (3.13)}$$

Where $B=\frac{\lambda_{ms}}{\lambda_{mf}}$

Knowing that $\lambda_{ms}=D_s\rho_sC_{ms}$, $\lambda_{mf}=D_f\,\rho_f\,C_{mf}$ and $C_m=C_{ms}/C_{mf}$, so equation (3.13) can be written in a form more appropriate for further discussion as

$$\lambda_{mp}=\beta\varepsilon_pD_f\,\rho_f\,C_{mf} \qquad \text{--- (3.14)}$$

Where

$$\beta=\frac{(1/\varepsilon_p-1)(1-1/B)}{\frac{1}{B}+0.28\varepsilon_p^{0.63B^{0.18}}}+\frac{1}{\varepsilon_p} \qquad \text{--- (3.15)}$$

and $B=\frac{D_s\rho_sC_m}{D_f\,\rho_f}$

The solution of equation (3.10) for constant λ_{mp} , C_{ms} and C_m and for boundary conditions given in equation (3.11) has the form:

$$\frac{\theta-\theta_w}{\theta_o-\theta_w}=erf\left[\frac{z}{2}\left(\frac{(1-\varepsilon_p)\rho_sC_{ms}}{\lambda_{mp}\tau}\right)^{1/2}\right] \qquad \text{--- (3.16)}$$

From which the instantaneous mass flux calculated as:

$$N_{p\tau} = -\lambda_{mp}\left(\frac{\partial\theta}{\partial z}\right)_{z=0}$$

$$= \left(\frac{\left(1-\varepsilon_p\right)\rho_s C_{ms}\lambda_{mp}}{\pi\tau}\right)^{1/2}\left(\theta_w - \theta_o\right) \quad \text{--- (3.17)}$$

In order to get the time-averaged mass flux N_p information about the packet residence time distribution is necessary. In this work an exponential distribution is assumed, so that

$$N_p = \frac{1}{\bar{\tau}_p}\int_0^\infty N_{p\tau}(\tau)\exp(-\tau/\bar{\tau}_p)d\tau \quad \text{--- (3.18)}$$

After substitution of equation (3.17) into equation (3.18) and integration, the packet mass transfer resistance is calculated as:

$$R_{mp} = \frac{N_p}{\theta_w - \theta_o} = 2\left(\frac{\bar{\tau}_p}{\left(1-\varepsilon_p\right)\rho_s\lambda_{mp}C_{ms}}\right)^{1/2} \quad \text{--- (3.19)}$$

Where, according to equations (3.7), (3.8) and (3.14)

$C_{ms} = C_m$ $\qquad C_{mf} = 1$

$$\lambda_{mp}C_{ms} = \beta\varepsilon_p D_f\,\rho_f C_m \qquad \text{for} \quad \theta = Y \quad \text{---(3.20)}$$

$C_{ms} = X_w$ $\qquad C_{mf} = X_w / C_m$

$$\lambda_{mp}C_{ms} = \beta\varepsilon_p D_f\,\rho_f C_{ms}^2 / C_m \qquad \text{for} \quad \theta = X / X_w$$

In assumption that there is an additional contact mass resistance R_{mw} in the packet zone adjacent to the surface which is proportional to the particle diameter d_s. The formula for mass transfer coefficient will then be:

$$k_{\theta p} = (R_{mw} + 0.5R_{mp})^{-1} \qquad , \quad R_{mw}\,\alpha d_s \qquad \text{--- (3.21)}$$

From equations (3.19) and (3.22) it follows that the mass transfer coefficient for packets ($k_{\theta p}$) is a function of the driving force $(\theta_w - \theta_o)$ if $\theta = Y$ or $\theta = X / X_w$. Hence the driving force should be selected experimentally to assure the smallest variation of $k_{\theta p}$.

From equations (3.19) and (3.22) found that for a negligible contact resistance, i.e. for sufficiently small values of $(d_s^2 C_{ms} / \bar{\tau}_p)$, the mass transfer coefficient is independent of particle size.

3.2.2 Simplified Packet Theory

In order to derive the simplified packet model equations the following additional assumptions are made.

1- For sufficiently short packet contact times which correspond to vigorous fluidization and for relatively large particles only the first layer of particles, i.e. those in contact with the surface, participate in surface-packet mass transfer.

2- During the time that a packet remains at the surface a particle in the first layer adsorbed to the surface.

3- A diffusion model of mass "conduction" within a particle may be plausible.

Treating the particles as spheres of uniform diameter and constant diffusivity, the equation describing the concentration profile (X_r) can be written as:

$$\frac{\partial X_r}{\partial \tau} = D_s \left(\frac{\partial^2 X_r}{\partial r^2} + \frac{2}{r} \frac{\partial X_r}{\partial r} \right) \qquad \text{--- (3.22)}$$

With boundary and initial conditions

$X_r(\tau = 0, r) = X_o$

$$D_s \rho_s \left(\frac{\partial X_r}{\partial r} \right)_{r=d_s/2} = k_{ys} \left(Y_w - Y_{r=ds/2} \right) \qquad \text{---} \qquad (3.23)$$

Bearing in mind that

$k_{ys} = Sh_s D_f \rho_f / d_s$

$X_w - X_{r=d_s/2} = C_m (Y_w - Y_{r=d_s/2})$

By transforming equation (3.23) will give Biot number:

$$Bi_{ms} = Sh_s / (2B) \qquad \text{--- (3.24)}$$

The well known solution of equation (3.22) leads to the following expression describing the time dependence of the mean concentration *X*.

$$\frac{X_w - X}{X_w - X_o} = 1.5 Bi_{ms}^2 \sum_{i=1}^{\infty} \eta_i \exp(-\mu_i^2 Fo_{ms}) \qquad \text{--- (3.25)}$$

Where

$\eta_i = \frac{\sin^2 \mu_i}{\mu_i^3 \left(\mu_i - 0.5 \sin 2\mu_i' \right)}$ and μ_i is the root of (1-Bi_{ms})

After deriving the result of different of equation (3.25) by equation (3.25):

$$\frac{\partial X}{\partial \tau}=\frac{4D_s}{d_s^2}\frac{\sum_{i=1}^{\infty}\eta_i\,\mu_i^2\exp(-\mu_i^2 Fo_{ms})}{\sum_{i=1}^{\infty}\eta_i\,\exp(-\mu_i^2 Fo_{ms})}(X_w - X) \qquad \text{--- (3.26)}$$

When ($Bi_{ms} < 10$, and $Fo_{ms} < 10^{-3}$) this expression can be approximated with an accuracy of + 10 % by the form resulting from $Fo_{ms} = 0$.

$$\frac{\partial X}{\partial \tau}=\frac{\hbar D_s}{d_s^2}(X_w - X) \qquad \text{--- (3.27)}$$

Where $\hbar = 4\sum_{i=1}^{\infty}\eta_i\,\mu_i^2 / \sum_{i=1}^{\infty}\eta_i$ is a constant while a packet is in contact with the surface. Note that this approximation should be confused with that well known for large Fourier numbers.

Equation (3.27) may be extended to the more general case of non-spherical particles with an unknown or undetermined diffusivity as follows:

$$\frac{\partial X}{\partial \tau}=\frac{a_s}{d_s^2}(X_w - X) \qquad \text{with} \quad \frac{\partial a_s}{\partial \tau}=0 \qquad \text{---(3.28)}$$

or ,after integration

$$\frac{\partial X}{\partial \tau}=\frac{a_s}{d_s^2}\exp\left(-\frac{a_s\tau}{d_s^2}\right)(X_w - X_o) \qquad \text{--- (3.29)}$$

The instantaneous mass flux from the surface to a packet is expressed as:

$$N_{p\tau}=\frac{b_s}{d_s^2}m_s\frac{\partial X}{\partial \tau} \qquad \text{--- (3.30)}$$

Where m_s is the particle mass and (b_s / d_s^2) is the number of particles per unit area of the surface.

For non-spherical particles only an approximate value of the packing factor ($b_s \approx 1$) can be found. While for spherical particles ($b_s = 1 \to 4/\sqrt{3}$).

The time-averaged mass flux results from equations (3.18), (3.29) and

(3.30) as follows:

$$N_p = \frac{\pi a_s b_s \rho_s / d_s}{6(1 + a_s \bar{\tau}_p / d_s^2)}(X_w - X_o) \qquad \text{--- (3.31)}$$

For very small values of $(a_s \bar{\tau}_p / d_s^2)$, this equation simplified to a form independent of the mean packet residence time.The packet mass transfer coefficient k_{xp} will then be described by the simple formula :

$$k_{xp} = \frac{N_p}{1 - X_o / X_w} = \frac{\alpha \rho_s X_w}{d_s} \qquad \text{--- (3.32)}$$

Where $\alpha = \pi a_s b_s / 6$ --- (3.33)

It follows that both coefficients (a_s and b_s) cannot be determine directly. Hence, the parameter of the simplified packet model (α) cannot be calculated from equation (3.33) and must be found on the basis of experimental data.

The total mass flux from an immersed surface to a mass capacity fluidized bed should be calculated from the expression analogous to heat transfer:

$$N = (1 - f_o) k_{\theta p} (\theta_w - \theta_o) + N' \qquad \text{--- (3.34)}$$

Where $N' = k_y' (Y_w - Y_o)$

N' is the so-called convective component of the mass flux including all effects not connected with particle mass capacity (note that $C_{ms}=0$ implies that $k_{\theta p} = 0$). Hence, k_y' in equation (3.34) is interpreted as the total mass transfer coefficient for an inert fluidized bed and can be predicted from equation (3.1).

Chapter Four

Experimental Work

Sand-air-naphthalene system is used in this investigation. After careful review of substances which exist in the solid state at room temperature and which are also considerably volatile, naphthalene was selected for this study. Spheres (made of wood) were coated with hard smooth surface of naphthalene by dipping the spheres into a bath of molten naphthalene (at about 90°C).

Sand was employed as fluidizing particles, which can be regarded as a non absorptive material. Three different particle sizes of sand were used, with sizes distributed between 75-250 microns in order to get a smooth fluidization. Table (4.1) gives the properties of sand particles used in this work.

A cylinder made of Q.V.F glass was used as a fluidization column that contained the particles supported by the distributor. The distributor used In this work is the drilled plate.

Table (4.1)

Properties of Sand Particles

Range of Particle Size (*micron*)	**Average Particle Size (*micron*)**	**Particle Density (Kg/m^3)**
75-150	112.5	2600
150-180	165	2600
180-250	215	2600

4.1 General Description of Flow System

A schematic diagram of the equipment is shown in Fig.(4.1), and a photograph picture of it in Fig.(4.2).The flow system consisted of the following:

1- Air compressor.
2- Air flow meter.
3- Fluidization column.
4- Manometer (U-tube).
5- Bed material (sand).
6- Immersed work piece.
7- Heating equipment (heating element, variac).
8- Temperature measurement device.

4.1.1 Air Compressor

A piston type compressor supplied by (Tecquipment, Ltd. Notingham, England), was used, with two pistons to supply the air with a surge tank to store the air to minimize the fluctuation through flow meter.

An automatic regulator in the compressor was used to regulate the pressure of the air inside the tank. The amount of air that left the compressor was controlled by the use of the tank valve.

4.1.2 Air Flow Meter

A calibrated air flow meter was used to measure the air flow rate that entered the column. The range of the air flow meter was 0-16 (m^3/hr).

4.1.3 Fluidization Column

A Q.V.F glass column of (8 cm) internal diameter, and (70 cm) in high was used as fluidization column. A porous material was placed at the bottom of the column to support the packing material.

4.1.4 Pressure Manometer

The pressure drop across the bed was measured by the use of (U-tube) manometer made of glass with total length of (0.75 m). The manometer was placed on a wide sheet of wood with a measuring tape for the measurement of the level difference of the liquid (water) inside the tube.

4.1.5 Bed Material

A sand bed material was employed in this investigation with a different particle sizes. (75-250 microns) were used in this investigation in order to get a smooth fluidization.

4.1.6 Immersed Work piece

The immersed object used in this investigation has a spherical shape (2.9 cm in diameter) made of wood, which is coated with naphthalene, is fixed in the center of the column by suspending it with a steel rod. The spherical shape was used in order to minimize the dead zone around the immersed object, and because spherical shapes have many applications in the industrial.

4.1.7 Heating Equipment

The heater has been used was an electrical heater placed inside 2" Q.V.F. glass tube. The variation in the heat supplied from the heater was controlled by the use of a variac connected directly to the heater.

4.1.8 Temperature Measurement Device

Two thermocouples (K type) were used for temperature measurement; each thermocouple is with a range of (0-250 C̊). The thermocouples were located in two different locations in order to measure the temperature (about 3 cm) under and above the sphere. These thermocouples were connected to the digital readers that show the value of any temperature selected.

Table 4.2

Conditions of Mass Transfer

Range of Particle Size(μm)	**Range of Temperature (C̊)**	**Range of Air Flow Rate(m^3/hr)**
215	39-66	4.0-5.2
165	39-66	3.0-4.0
112.5	39-66	2.4-3.2

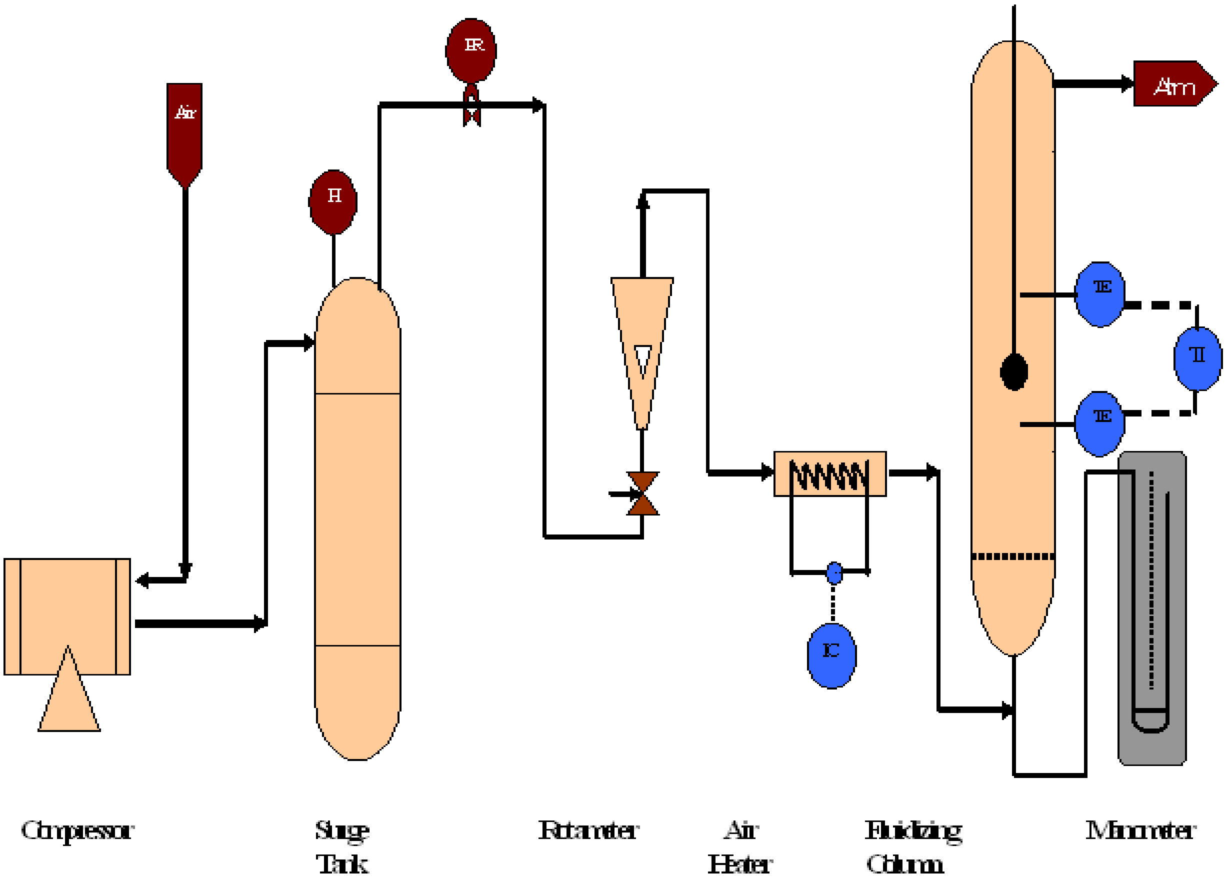


Figure (4.1): Diagram of The Experimental Equipment

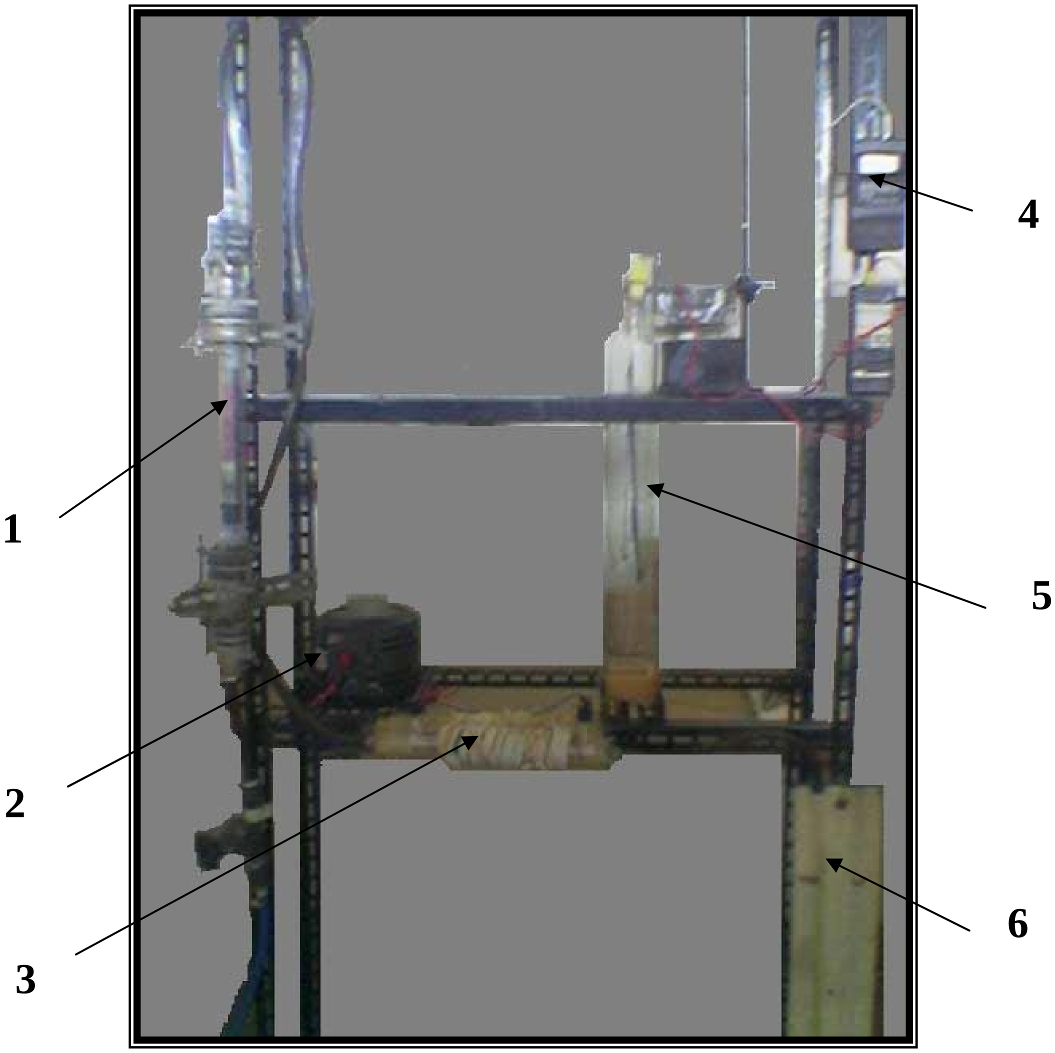


Figure (4.2)

Photo of the Experimental Equipment

1. **Rotameter**
2. **Variac**
3. **Heating Equipment**
4. **Temperature Reader**
5. **Fluidization Column**
6. **Manometer**

4.2 Experimental Procedure

4.2.1 Determination of Minimum Fluidizing Velocity

The pressure drop of the bed (Δp_b) can be determined with the range of gas velocities by subtracting the pressure drop of distributor (Δp_d) that find out for a range of superficial gas velocities, from total pressure drop (Δp_t) that find out for a range of superficial gas velocities after loading known weight of sand particles having known diameter into the bed to a static level of 30 cm.

A curve of pressure drop across the bed (Δp_b) versus superficial gas velocity is determined from this plot (see appendix A).

4.2.2 Mass Transfer in Empty Bed

To determine the experimental value for mass transfer coefficient in empty bed, the following procedure was carried out:

1- Two thermocouples and other devices were placed in their place and connected to the column.
2- The compressor started blowing air into the tank until it reached the desired pressure to turn the compressor off by the automatic regulator. The tanks valve was turned on. The air flowed through the rotameter to the bed until rotameter read a constant desired value of the air flow rate.
3- At the same time the heater was turned on for the desired power that was controlled by the use of the variac.

4- The measurements of the pressure drop across the bed were made by the use of the U-manometer.
5- When a conditions reach to steady state (constant flow rate and constant temperature), the coated sphere was lowered inside the column 15 cm above the distributor surface.
6- Every 5 minutes, the sphere was taken out of the bed and the change of weight was measured by digital balance.
7-The previous steps 1-6, were repeated for arrange of air superficial velocities and a range temperatures.

4.2.3 Mass Transfer in Fluidized Bed

To determine the value of mass transfer coefficient from the sphere to the fluidized bed, the following steps were carried out:

1-A quantity of sand was weighted and poured into the column from the top for a known and constant height (30 cm) for all runs carried in the work.
2-Two thermocouples and other devices were placed in their place and connected to the column.
3-The compressor started blowing air into the tank until it reached the desired pressure to turn the compressor off by the automatic regulator.
4-The tanks valve was turned on. The air flowed through the rotameter to the bed until rotameter read a constant desired value of the air flow rate.
5-At the same time the heater was turned on for the desired power that was controlled by the use of the variac.
6-The measurements of the pressure drop across the bed were made by the use of the U-manometer.

7-When conditions reach to steady state (constant flow rate and constant temperature); the coated sphere was lowered inside the column 15 cm above the distributor surface.

8-Every 5 minutes, the sphere was taken out of the bed and the change of weight was measured by digital balance.

9-The previous steps 1-8, were repeated for arrange of air superficial velocities and a range bed temperatures.

4.3 Correlation of Experimental Results

A relationship of many variables that influence on mass transfer coefficient must be known to determine. These variables are the diffusivity of the active component through the fluid, the superficial flow rate of the fluid, the density and the viscosity of the fluid, and the shape and the size of the spaces between the particles in the fluidized bed.

To get accurate results, numbers of assumptions were made:

1- Neglect the abrasion effects and assume the weight loss of naphthalene is mainly due to evaporation.

2- Void fraction of fluidizing sand particles was equal to the void fraction at minimum fluidizing velocity.

3- Partial pressure of naphthalene at the surface (P_i) was everywhere equal to the saturation partial pressure of vapor at (T_s) The partial pressure of the naphthalene at the bulk of air stream was equal to zero. Change in surface area of the sphere along the experiment is negligible. Surface temperature of the solid sphere (T_s) was everywhere equal to the average value of the temperature reading of the thermocouples below and above the sphere.

The experimental results must be correlated by:

1-The viscosity of air can be calculated from the equation **[57, 36]**:

$$\frac{\mu}{\mu_o} = \left[\frac{T}{273}\right]^n \quad \text{--- (4.1)}$$

Where μ_o is the viscosity of the air at 0Č which is equal to 0.017 in centipoises and *n* is equal to 0.677.

2- Experimental value of mass transfer coefficients were calculated from the equation:

$$N = k_g (C_s - C_b) \quad \text{--- (4.2)}$$

In which C_b is equal to zero.

3- The correlation for diffusivity of naphthalene vapor in air with temperature is made by the equation **[36]**:

$$\frac{Dv}{Dv_o} = \left[\frac{T}{273}\right]^m \quad \text{--- (4.3)}$$

Where the diffusivity of naphthalene vapor in air at 0Č is taken equal to 0.0513 cm^2/s **[57]**, and m= 1.823.

4- Vapor pressure of solid naphthalene is given by following equation for the range of (0-80 Č).

$$\log Ps = -\frac{3729.3}{Ts} + 11.450 \quad \text{--- (4.4)}$$

Where P_s in mHg and T_s in K.

5- Values of Sherwood number for mass transfer from the sphere to a bed of fluidized particles were calculated by the equation:

$$Sh = Sh_e + f(y) \quad \text{--- (4.5)}$$

In which *f(y)* describes the effect of particles motion on transfer rate, and *y* is a dimensionless group determined by dimensionless analysis.

6- The value of Sherwood number for mass transfer in empty bed was calculated from the equation **[28]**:

$$Sh_e = 2 + C_1 \mathrm{Re}_p^{C_2} Sc^{C_3} \qquad \text{--- (4.6)}$$

Where C_1 ,C_2 and C_3 are constants and to be determined from the experimental results in empty bed.

Chapter Five

Results and Discussion

Twenty four experiments at different air velocities and different temperatures were performed at first for mass transfer in empty bed (air stream only), to check the results with previous works. Samples of data sheets for experiment listed in Tables (5.2, 5.3 and 5.4) respectively. List of experimental conditions are illustrated in Table (A-1).

To correlate the experimental results for mass transfer in gas stream only, it was assumed that the limiting value of Sherwood number, at zero Reynolds number, is equal to 2 because it agrees with the theoretical approach. The experimental results were correlated by the following equation (see Appendix B):

$$Sh_e = 2 + 0.657\,\mathrm{Re}_p^{1/2}\,Sc^{1/3} \qquad \text{--- (5.1)}$$

The above equation shows good agreement with the previous work by Ranz and Marshal **[28]**. Fig.(5.2) illustrates the experimental results and correlation.

For mass transfer in fluidized bed, a set of experiments were performed to determine the value of mass transfer coefficient from the sphere to the fluidized bed. The variables in this investigation which effect mass transfer were: air velocity, sphere surface temperature, size of fluidizing particles and sphere size. The range of sphere surface temperature varied from ambient temperature to a temperature below the melting point of naphthalene. The air velocity is chosen to be within the range (1-1.4) U_{mf}, because this range of flow is usually used in industrial practice. The particle size of sand was selected to be as fine particles in order to get a smooth fluidization. The experimental conditions and results are listed in Tables (A-3 and A-4) respectively.

- Samples of data sheet for experiments are listed in Tables (5.5, 5.6 and 5.7) respectively.
- Samples of the experimental result showing effect of air temperature on Sherwood number are illustrated in Figs. (5.3 and 5.4).
- Samples of the experimental result showing effect of air flow rate on Sherwood number are illustrated in Figs. (5.5).
- The effects of both air temperature and particle size on Sherwood number are showed in Figs. (5.8, 5.9 and 5.10).
- The effects of both air flow rate and particle size on Sherwood number are showed in Figs. (5.11, 5.12, 5.13).
- The effects of both particle size and different temperature on Sherwood number are showed in Figs. (5.14, 5.15, 5.16, and 5.17).

5.1 Correlation of Results

Results of experiments are illustrated in Table (A-4). For experiments that carried out at minimum fluidizing velocity, the value of the dimensionless group, Y, is inconsistent with other experiments due to the term (G-G_{mf}) which is equal to zero at minimum fluidizing velocity, so results obtained at minimum fluidization are neglected. The term $f(y)$ in the equation:

$$Sh = Sh_e + f(y) \qquad \text{--- (5.2)}$$

is chosen as a power function of Y ,that is

$$f(y) = C_1 y^{C_2} \qquad \text{--- (5.3)}$$

Two attempts have been made to correlate the experimental results:

1- The first attempt was made by choosing the dimensionless function ,*Y*, to be equal to $[(G - G_{mf})\mu / [(\psi d_p)^2 (\rho_p - \rho)\rho g]]$ as given by Ziegler and Holmes**[40]**,i.e.:

$$Sh = Sh_e + C_1 [\frac{(G - G_{mf})\mu}{(\psi d_p)^2 (\rho_p - \rho)\rho g}]^{C_2} \qquad \text{--- (5.4)}$$

The results of this attempt are illustrated in (Appendix C), in which C_1 and C_2 are equal to 9.571 and 0.01389 respectively. Fig.(5.6) shows a comparison of equation (5.4) with the experimental data. It can be seen from this figure, that the correlation suggested by Ziegler and Holmes don't fit the experimental results of this work.

2- The second attempt was made by taking the dimensionless group, Y , as obtained from the dimensionless analysis, i.e.:

$$Sh = Sh_e + C_1 [\frac{(\rho_p - \rho)\rho g (\psi d_p)}{(G - G_{mf})^2}]^{C_2} [\frac{(\psi d_p)(G - G_{mf})}{\mu}]^{C_3} \qquad \text{--- (5.5)}$$

The result of this attempt is illustrated in (Appendix C), in which C_1, C_2 and C_3 are equal to 16.8574, 0.07497 and 0.1284 respectively. Fig.(5.7) shows comparison of equation (5.5) with the experimental data. From this figure we notice that this correlation shows a better agreement with experiments, in which 97% of the points have an error less than 25%, consequently this correlation obtained from the present work.

5.2 Comparison of Experimental Results With previous Works and Model

Solid mass capacity has an essential effect on surface-to-fluidized bed mass transfer. For (C_{ms}=0) low mass transfer coefficients are attained and there is no similarity with surface-to-bed heat exchange. In the case of non-zero solid mass capacity, mass transfer coefficients are greater and for small values of (C_{ms}) they may be predicted from the theory proposed by Yokota **[55]**.

For relatively large values of ($C_{ms}d_s^2/\bar{\tau}_p$) (greater than 10^{-5} m^2 /s) the contact resistance is dominant and (k_{xp}) is inversely proportional to (d_s). For small values of ($C_{ms}d_s^2/\bar{\tau}_p$) (less than 10^{-10} m^2 /s) the packet resistance predominates and the surface-to-packet mass transfer coefficient (k_{xp}) is independent of particle size (see Table (5.1)).

Some values of the mass transfer data (k_{yp}) obtained from previous works and this work are presented in Fig. (5.1) in Yokota's coordinate.

Table (5.1) Comparison of the Orders of Magnitude of The Experimental Parameters

	Silica gel-air-water R_{mp}/R_{mw}	Sand-air-water $R_{mp}/R_{mw} \sim \infty$	Sand-air-naphthalene $R_{mp}/R_{mw} \sim \infty$
$\frac{D_f \rho_s}{\beta \rho_f}$ (m^2/s)	10^{-3}	10^{-2}	10^{-2}
$\bar{C}_m / C_{ms}$	10^{2}	10^{2}	10^{3}
$\frac{d_s^2 C_{ms}}{\bar{\tau}_p}$ (m^2/s)	$10^{-5}, 10^{-6}$	$10^{-11}, 10^{-12}$	$10^{-9}, 10^{-10}$

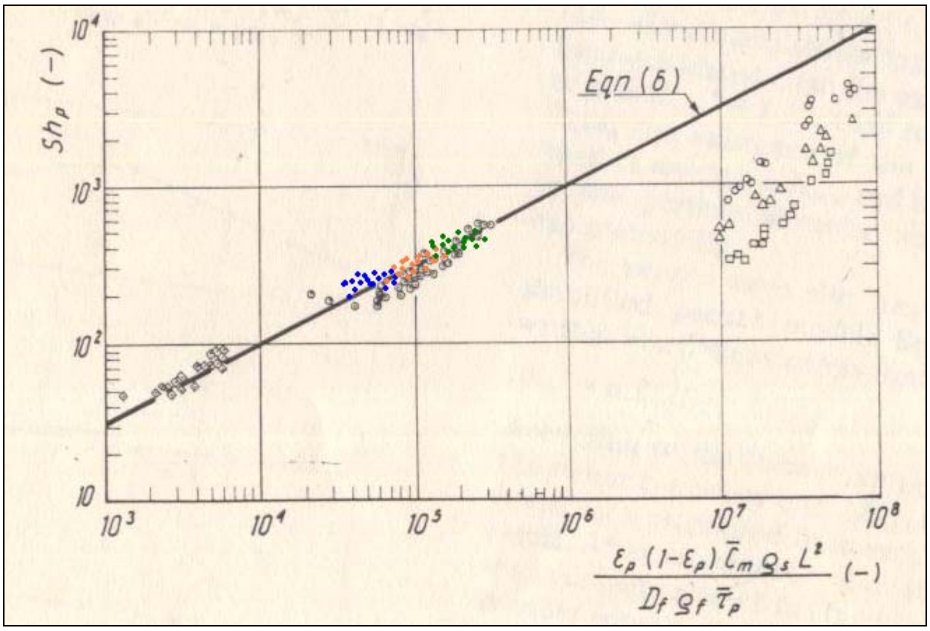


Figure (5.1): Comparison of Experimental Data with The Packet Theory Systems is follows: Silica gel-air-water [(o) d_s=0.548mm, (Δ) d_s=0.875 mm, (□) d_s=1.342mm]. Sand-air-water [(◊) d_s=0.496mm].Sand-air-naphthalene [(□)d_s=0.351mm]. **[55]**. Sand-air-naphthalene (present work) [(🟦) ds=215 micron, (🟧) ds=165 micron, (🟩) d_s= 112.5 micron]

]Table (5.2) Selected Data Sheet for Experiment of Mass Transfer in Empty Bed

Experiment No: (1)

Air flow rate (m^3/hr) = 2.8

Ambient temperature(C̊) = 39.0

Pressure drop (Cm H_2O) = 0.9

(Without Naphthalene)

Weight of sphere (gm) = 8.0088

Diameter of sphere (Cm) = 2.9

Time(*min.*)	**Wt.(*gm*)**	**T_1(C̊)**	**T_2 (C̊)**	**Wt. Loss (*gm*)**
0	10.2073	39.0	39.0	-
5	10.1194	39.1	39.1	0.0879
10	10.0402	39.0	39.1	0.0792
15	9.9680	39.0	39.0	0.0722
20	9.8847	39.1	39.0	0.0833
25	9.8051	39.0	39.0	0.0796

Note:

T_1 = Temperature below the sphere.

T_2 = Temperature above the sphere.

Table (5.3) Selected Data Sheet for Experiment of Mass Transfer in Empty Bed

Experiment No: (7)

Air flow rate (m^3/hr) = 2.8

Ambient temperature (°C) = 39.1

Pressure drop (Cm H_2O) = 0.9

(Without Naphthalene)

Weight of sphere (gm) = 8.0088

Diameter of sphere (Cm) = 2.9

Time(*min.*)	**Wt.(*gm*)**	**T_1(°C)**	**T_2 (°C)**	**Wt. Loss (*gm*)**
0	10.2210	51.3	51.3	-
5	9.9328	51.2	51.1	0.2882
10	9.6812	51.4	51.3	0.2516
15	9.3710	51.4	51.3	0.3102
20	9.1797	51.2	51.2	0.1913
25	9.0558	51.3	51.1	0.1239

Note:

T_1 = Temperature below the sphere.

T_2 = Temperature above the sphere.

Table (5.4) Selected Data Sheet for Experiment of Mass Transfer in Empty Bed

Experiment No: (15)	
Air flow rate (m^3/hr)	=3.7
Ambient temperature (C̊)	=39.1
Pressure drop (Cm H_2O)	= 1.4
(Without Naphthalene)	
Weight of sphere (gm)	= 8.0088
Diameter of sphere (Cm)	= 2.9

Time(*min.*)	**Wt.(*gm*)**	**T_1(*C̊*)**	**T_2 (*C̊*)**	**Wt. Loss (*gm*)**
0	12.4485	66.0	66.0	-
5	11.3768	66.2	66.1	1.0717
10	10.3553	66.2	66.1	1.0215
15	9.6381	66.0	66.1	0.7172
20	9.0130	66.1	66.0	0.6251
25	8.3365	66.0	66.1	0.6765

Note:

T_1 = Temperature below the sphere.

T_2 = Temperature above the sphere.

Table (5.5) Selected Data Sheet for Experiment of Mass Transfer in Fluidized Bed

Experiment No: (1)

Air flow rate (m^3/hr) =4.0

Ambient temperature (C̊) =39.0

Pressure drop (Cm H_2O) = 21.0

(Without Naphthalene)

Weight of sphere (gm) = 8.0088

Diameter of sphere (Cm) = 2.9

Time(*min.*)	**Wt.(*gm*)**	**T_1(*C̊*)**	**T_2 (*C̊*)**	**Wt. Loss (*gm*)**
0	12.7425	39.1	39.1	-
5	12.6278	39.1	39.0	0.1147
10	12.4324	39.0	39.0	0.1954
15	12.2603	39.2	39.1	0.1721
20	12.1274	39.0	39.1	0.1329
25	11.9746	39.2	39.2	0.1528

<u>Note</u>:

T_1 = Temperature below the sphere.

T_2 = Temperature above the sphere.

Table (5.6) Selected Data Sheet for Experiment of Mass Transfer in Fluidized Bed

Experiment No: (7)

Air flow rate (m^3/hr) = 4.8

Ambient temperature (C̊) = 39.3

Pressure drop (Cm H_2O) = 23.0

(Without Naphthalene)

Weight of sphere (gm) = 8.0088

Diameter of sphere (Cm) = 2.9

Time(*min.*)	**Wt.(*gm*)**	**T_1(C̊)**	**T_2 (C̊)**	**Wt. Loss (*gm*)**
0	12.3452	51.2	51.2	-
5	11.8920	51.0	51.0	0.4532
10	11.4041	51.2	51.1	0.4879
15	11.0048	51.1	51.1	0.3993
20	10.5330	51.1	51.0	0.4718
25	9.9769	51.0	51.1	0.5561

<u>Note</u>:

T_1 = Temperature below the sphere.

T_2 = Temperature above the sphere.

Table (5.7) Selected Data Sheet for Experiment of Mass Transfer in Fluidized Bed

Experiment No: (16)

Air flow rate (m^3/hr) = 5.2

Ambient temperature (C̊) = 39.4

Pressure drop (Cm H_2O) = 25.0

(Without Naphthalene)

Weight of sphere (gm) = 8.0088

Diameter of sphere (Cm) = 2.9

Time(*min.*)	**Wt.(*gm*)**	**T_1(*C̊*)**	**T_2 (*C̊*)**	**Wt. Loss (*gm*)**
0	15.5628	66.2	66.1	-
5	14.0431	66.1	66.1	1.5197
10	12.6253	66.1	66.0	1.4178
15	11.075	66.0	66.1	1.5503
20	9.7951	66.3	66.1	1.2799
25	8.0953	66.0	66.0	1.6998

Note:

T_1 = Temperature below the sphere.

T_2 = Temperature above the sphere.

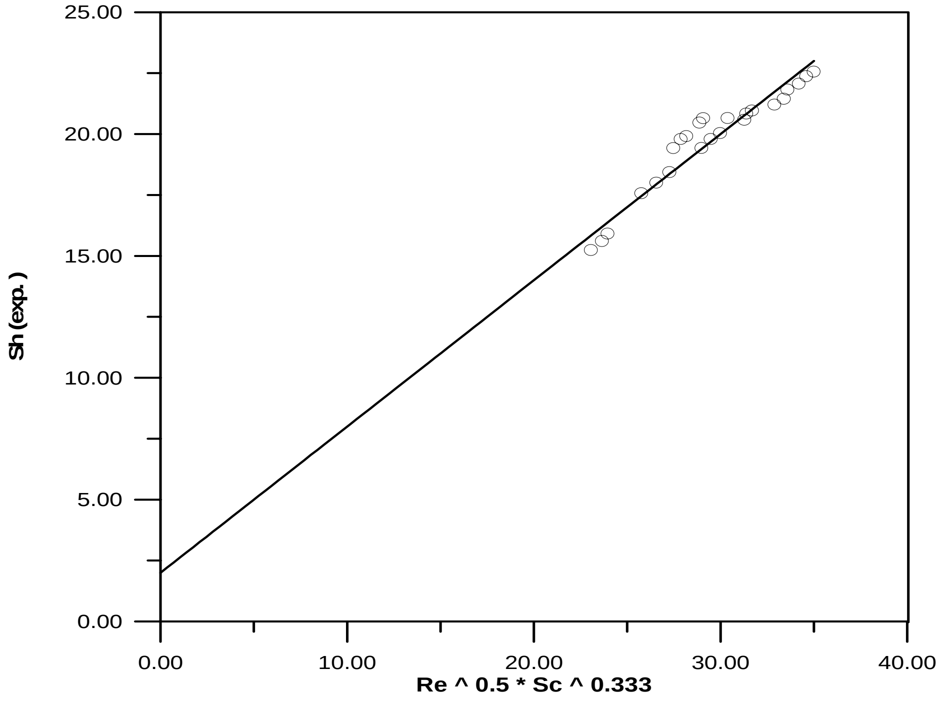


Figure (5.2): Correlation for Mass Transfer in Empty Bed

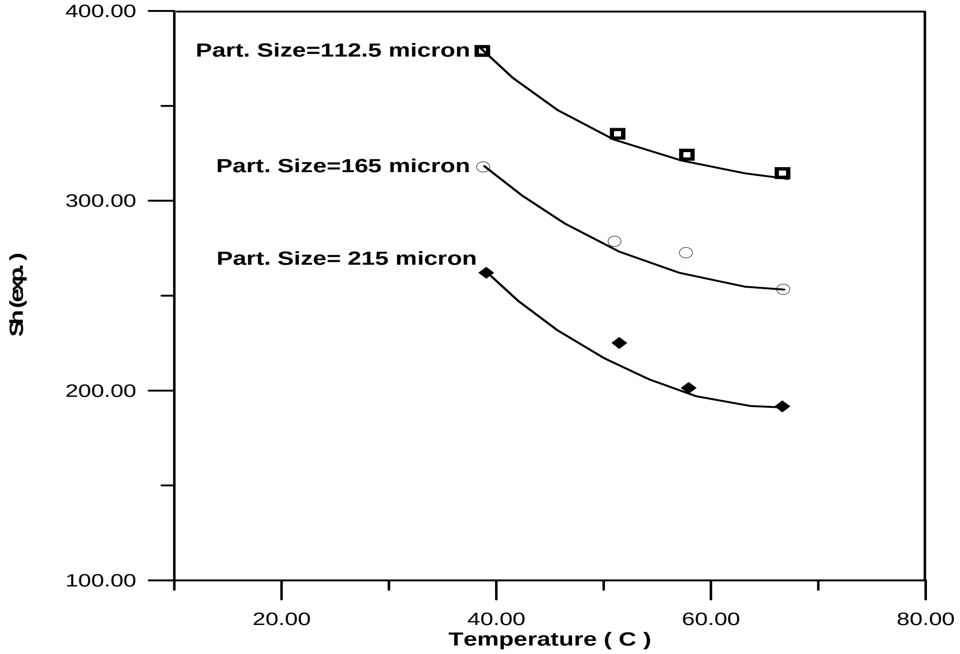


Figure (5.3): Experimental Sh. vs. Temperature at Air Flow Rate = 1.2 Umf.

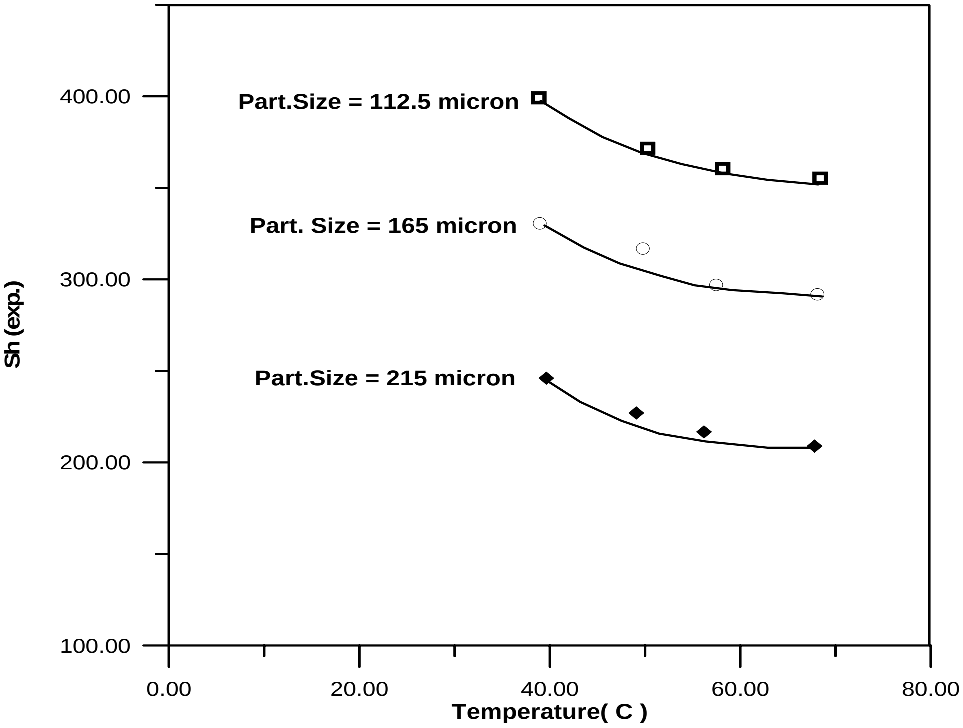


Figure (5.4): Experimental Sh. vs. Temperature at Air Flow Rate = 1.4 Umf

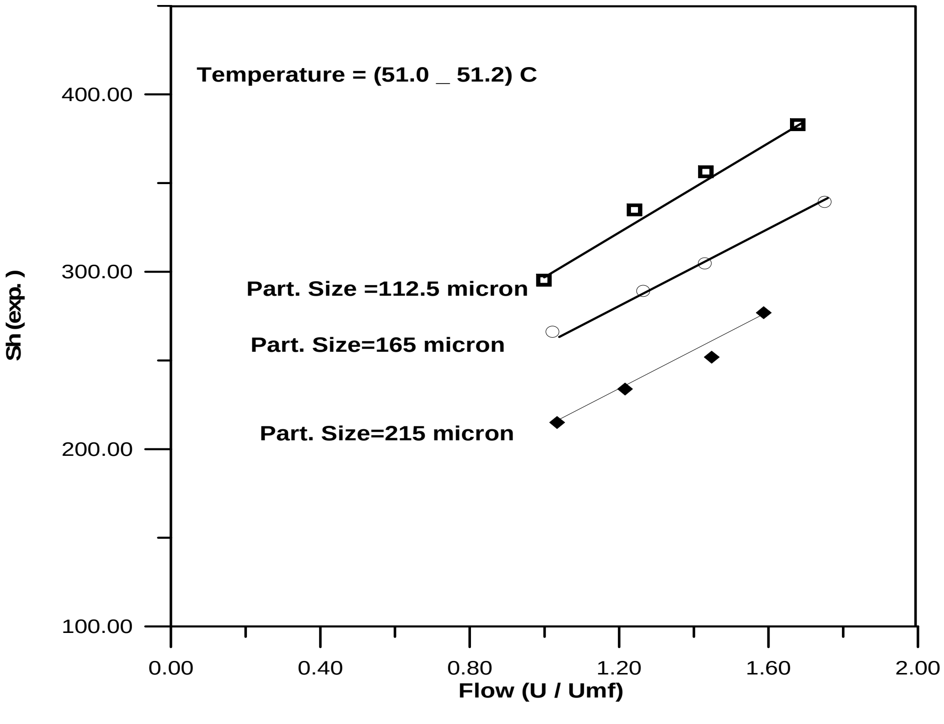


Figure (5.5): Experimental Sh. vs. Air Flow Rate

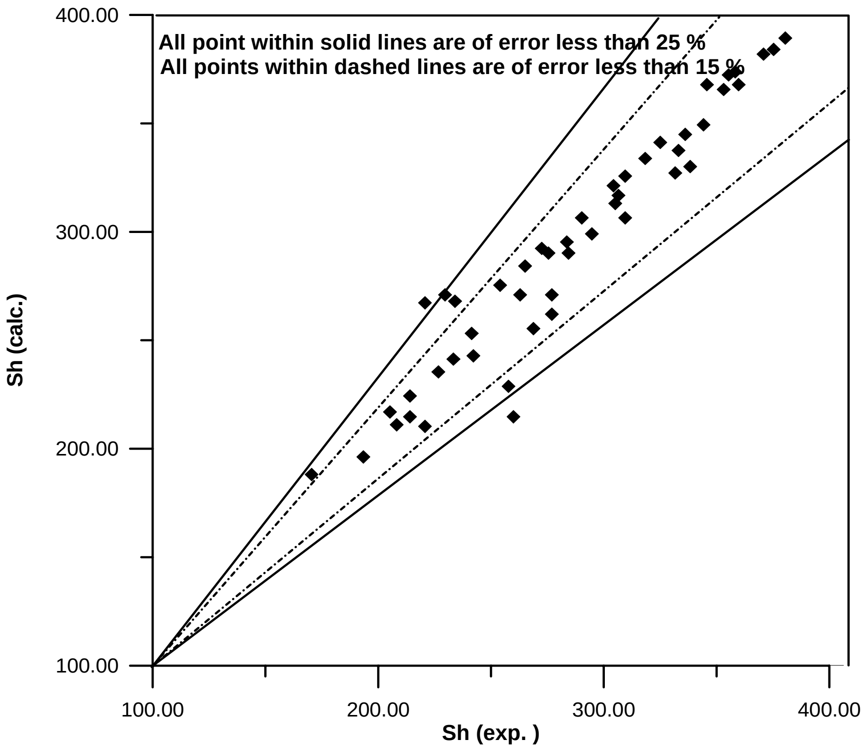


Figure (5.6): A Comparison of Equation 4.9 With the Experimental Data

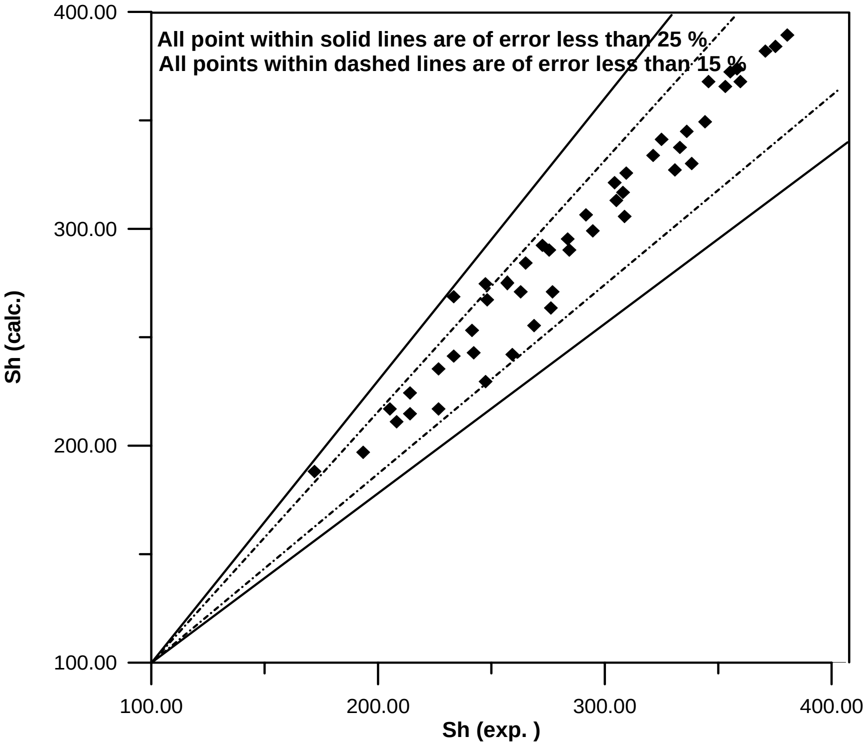


Figure (5.7): A Comparison of Equation 4.10 With the Experimental Data

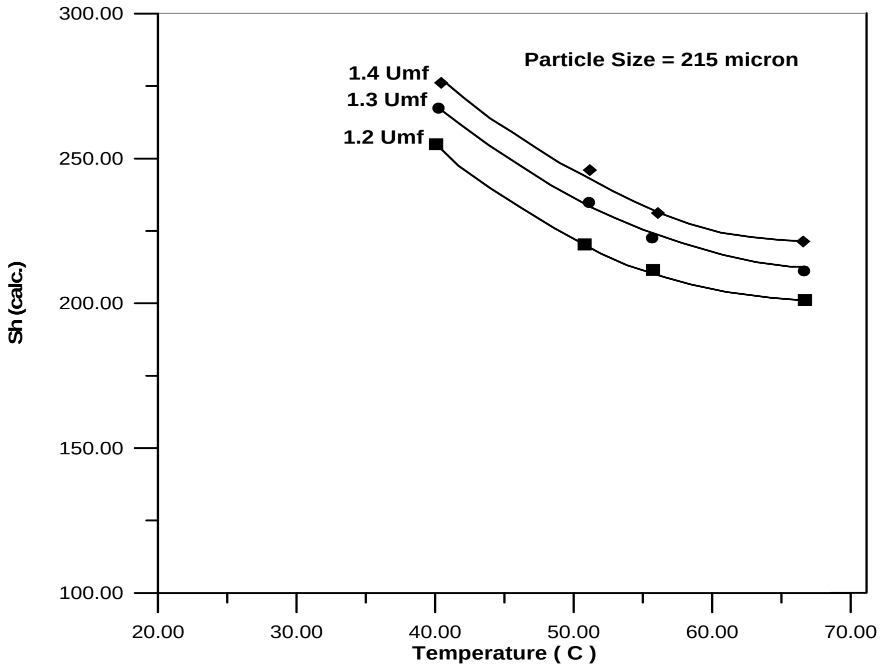


Figure (5.8): Effect of Temperature on Calculated Sh. No.

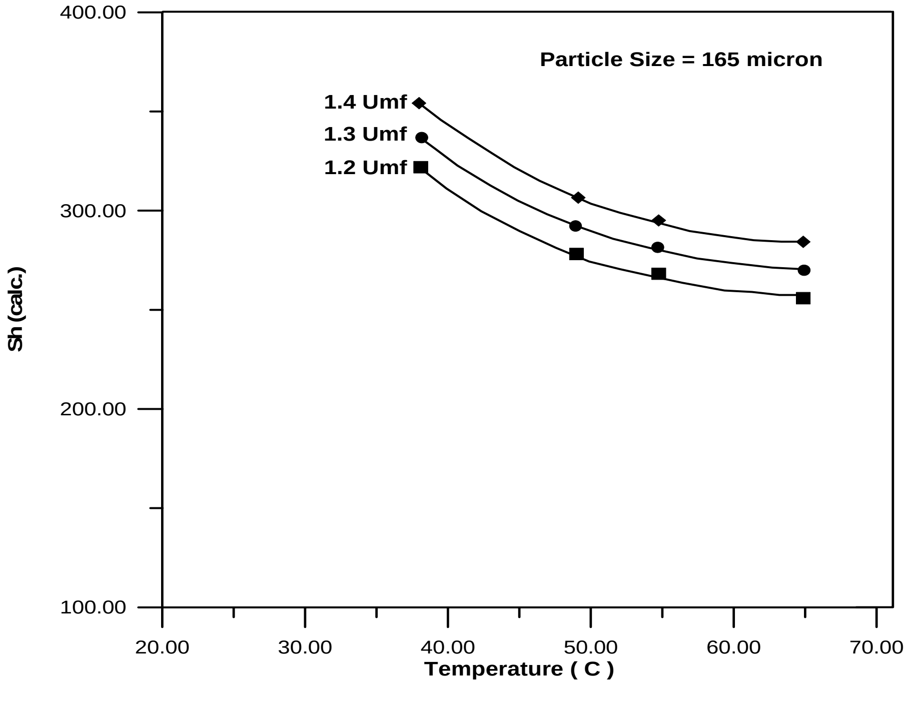


Figure (5.9): Effect of Temperature on Calculated Sh No.

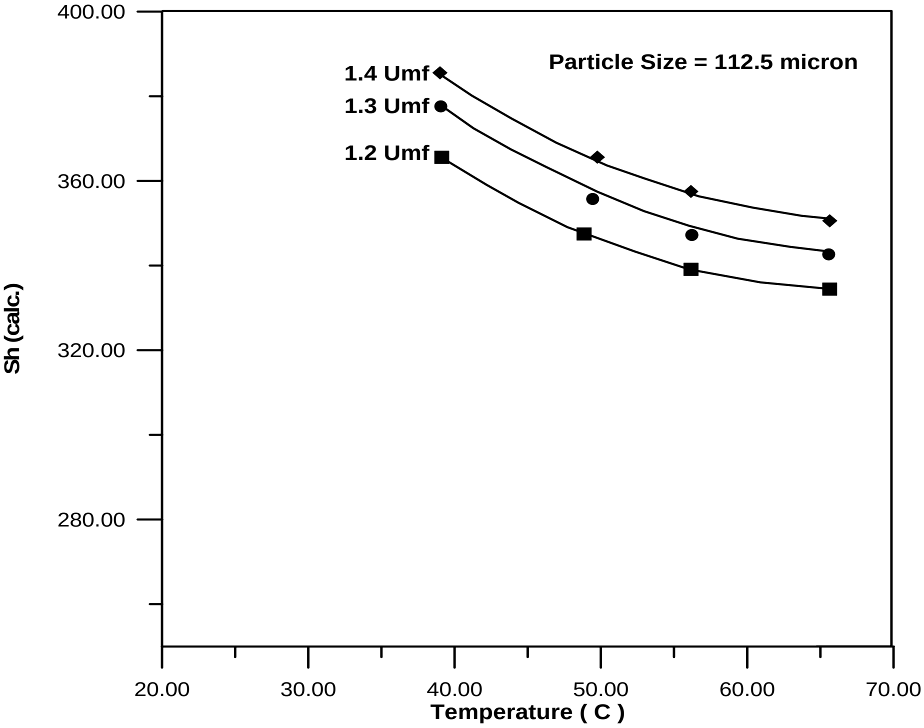


Figure (5.10): Effect of Temperature on Calculated Sh. No.

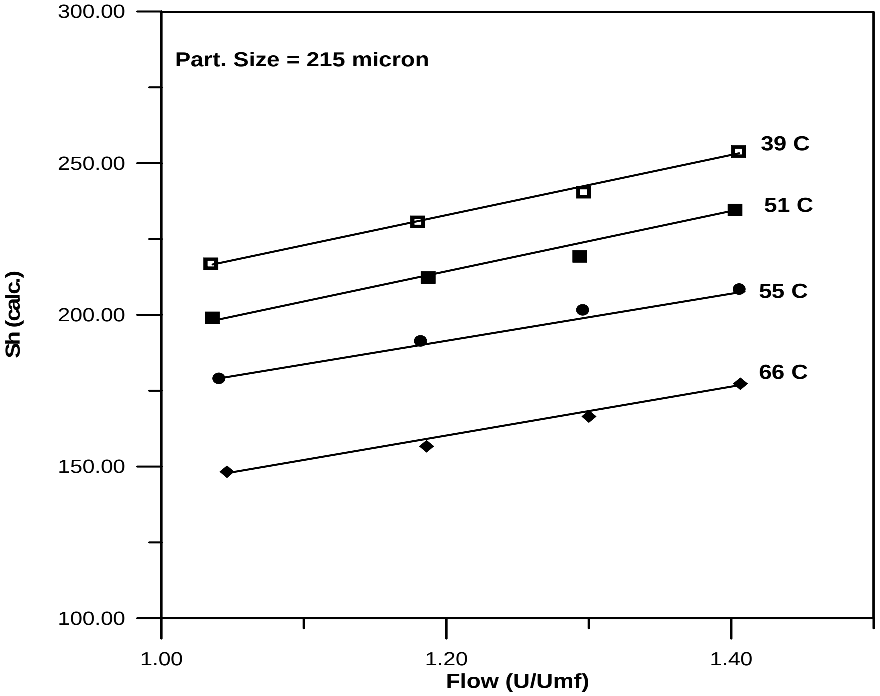


Figure (5.11): Effect of Air Flow Rate on Calculated Sh. No.

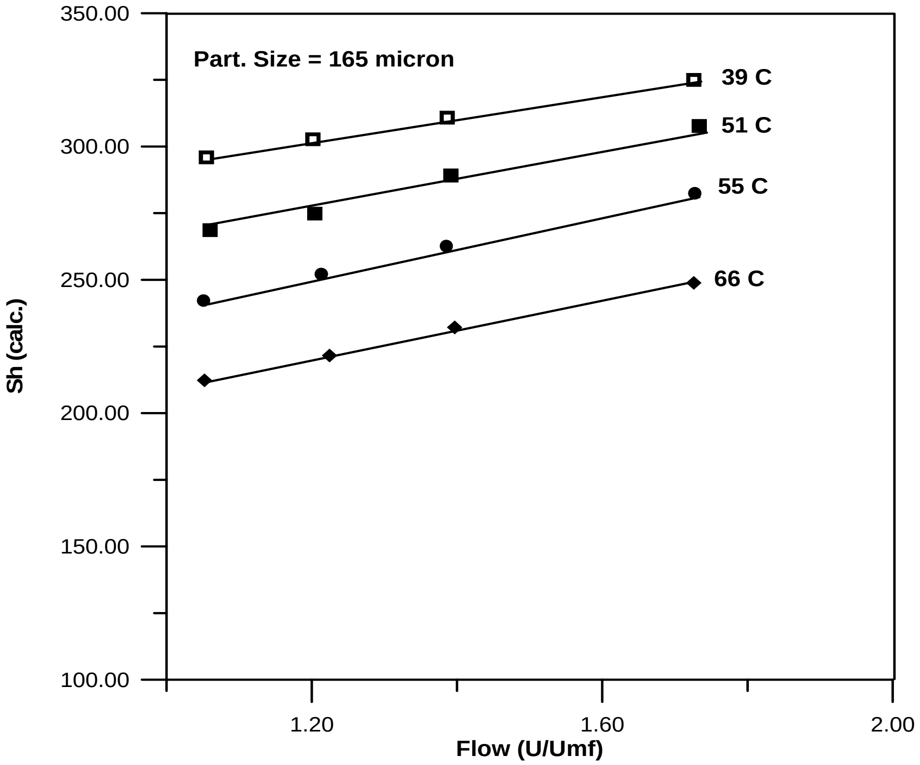


Figure (5.12): Effect of Air Flow Rate on Calculated Sh. No.

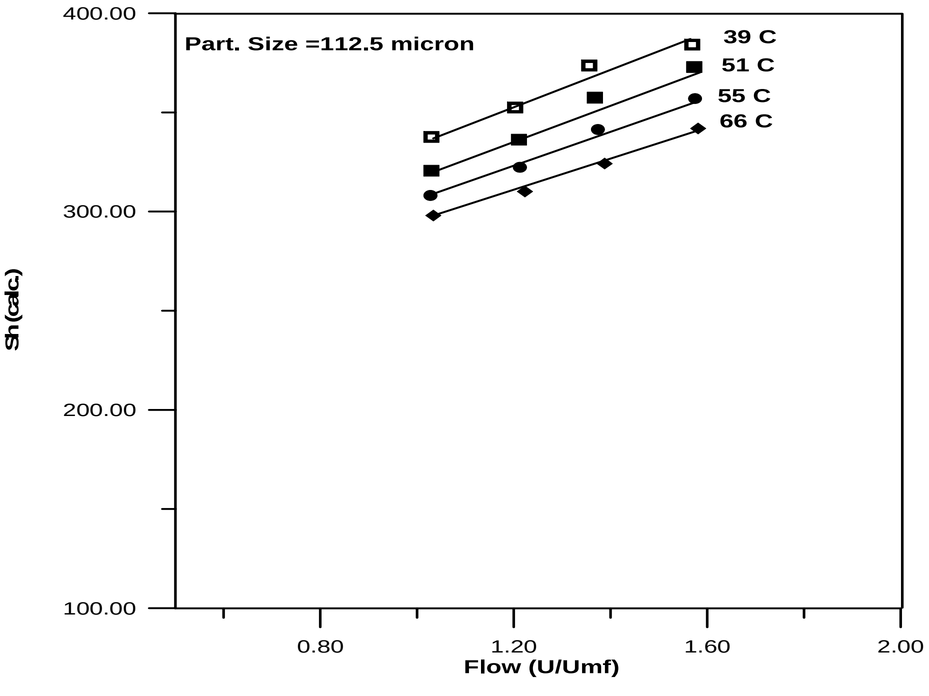


Figure (5.13): Effect of Air Flow Rate on Calculated Sh. No.

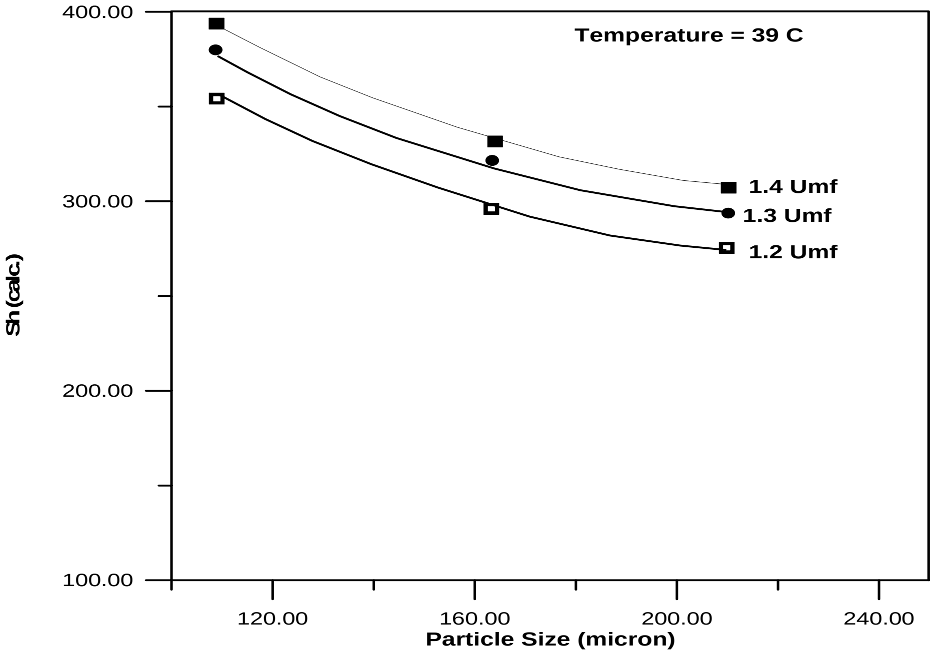


Figure (5.14): Effect of Sand Particle Size on Calculated Sh. No.

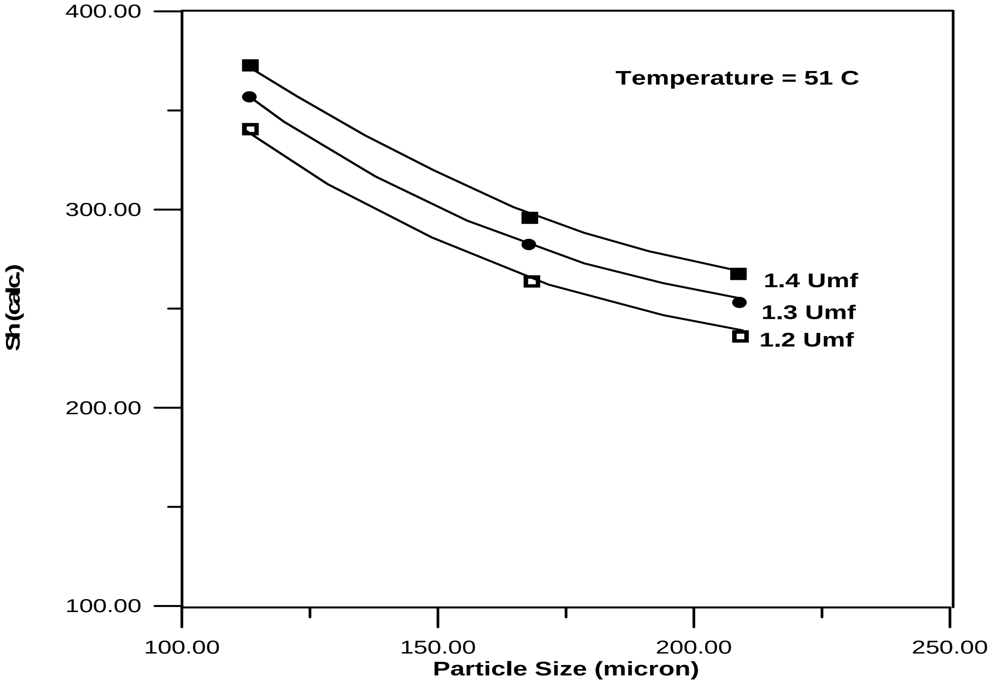


Figure (5.15): Effect of Sand Particle Size on Calculated Sh. No.

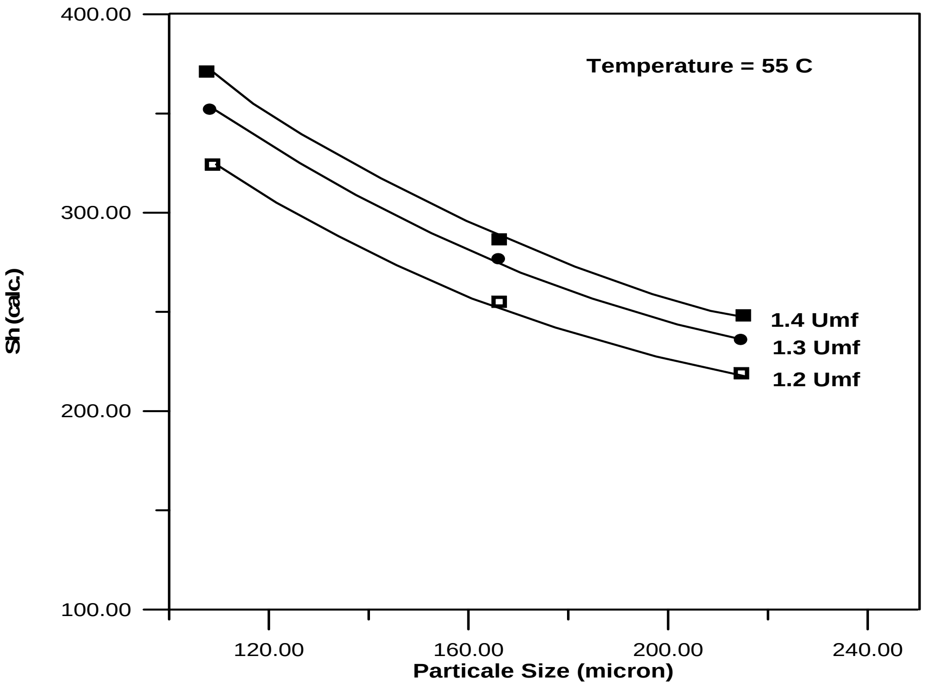


Figure (5.16): Effect of Sand Particle Size on Calculated Sh. No.

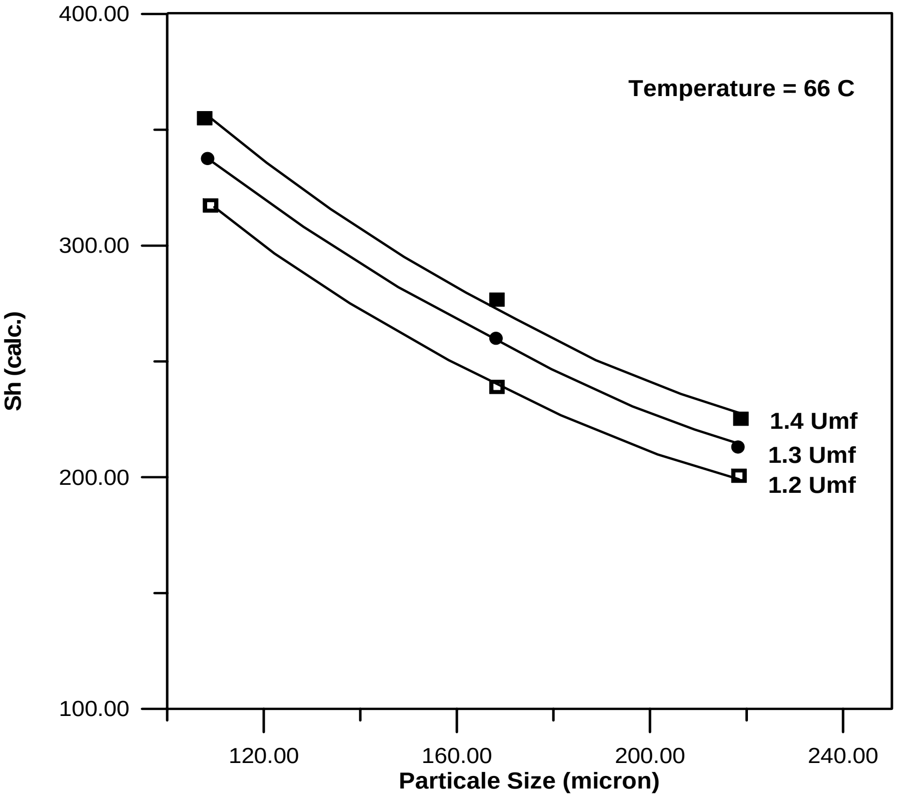


Figure (5.17): Effect of Sand Particle Size on Calculated Sh. No.

Chapter Six

Conclusion and Recommendations for Future work

6.1 Conclusions

From the investigation of mass transfer from an immersed surface to a gas fluidized bed, these conclusions are obtained:

1- Sherwood number has slow increasing with increasing gas velocity (at constant surface temperature and particle size).

2- Sherwood number increase with decreasing particle size (at constant U/U_{mf} and temperature).

3- Sherwood number increase with decreasing surface temperature of the sphere (at constant U/U_{mf} and particle size).

4- The ratio of Sherwood number for mass transfer in the presence of solid particles (fluidized bed) to that in absence of solid particles (empty bed) various up to (30).

5- Sherwood number depends on Reynolds number and Froude number (dimensionless groups), as noticed in equation (5.10).

6- There is a good improvement and accuracy in equation (5.10) (within a very good agreement of (1.54%) average error) than the equation presented by earlier work due to Ziegler and Holmes **[40]** which depends on one dimensionless group that result from the merge of the two dimensionless groups.

6.2 Recommendations for Future work

From the present study it was noticed that further studies in the following areas would be desired:

1- Studying the effect of geometry on transfer phenomenon by using different shapes of immersed objects which have important applications in industry.

2- Using different systems to study the common uses for mass transfer in fluidized bed.

3- Study of the mass transfer of the solid materials applied in Iraqi industries, by fluidized mass transfer to improve products quality.

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Appendix - A

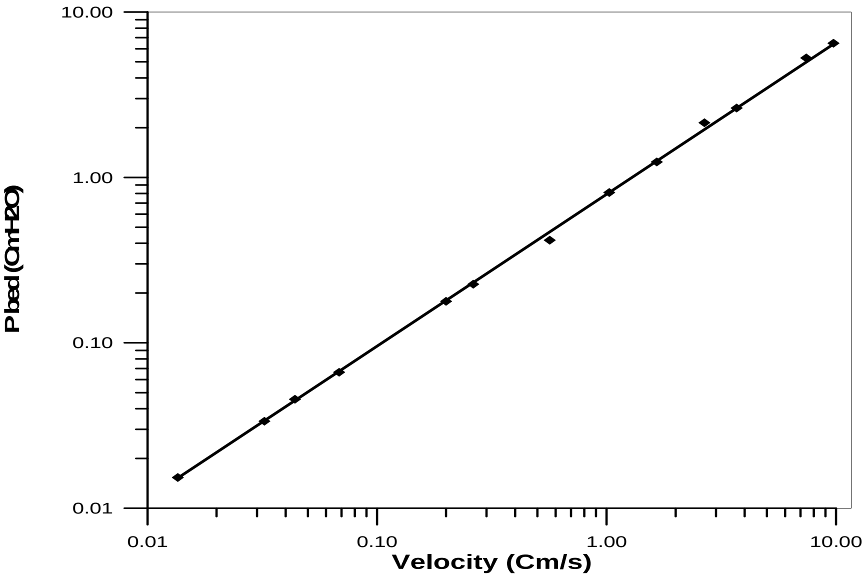

Figure (A.1): Distributor Pressure Drop

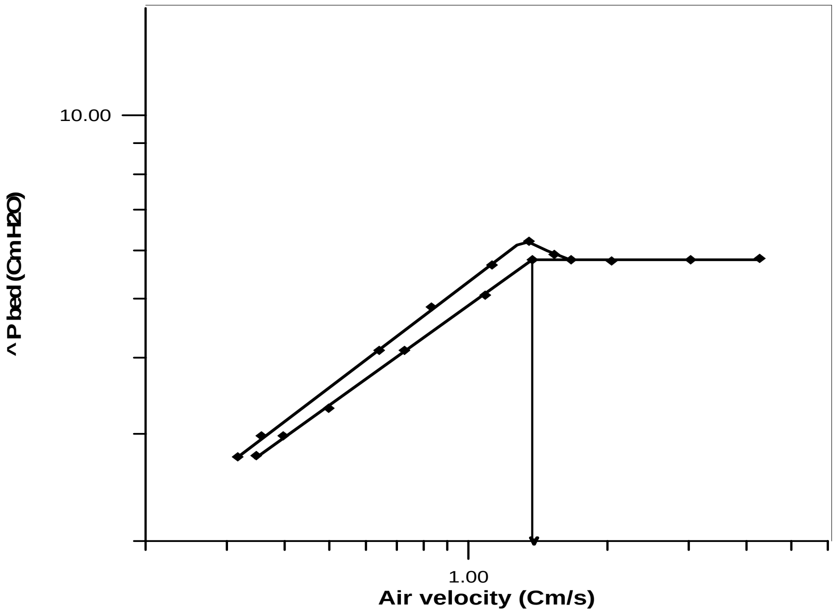

Figure (A.2): Bed Pressure Drop vs. Air Velocity

(Sand Particle Size = 215 Micron)

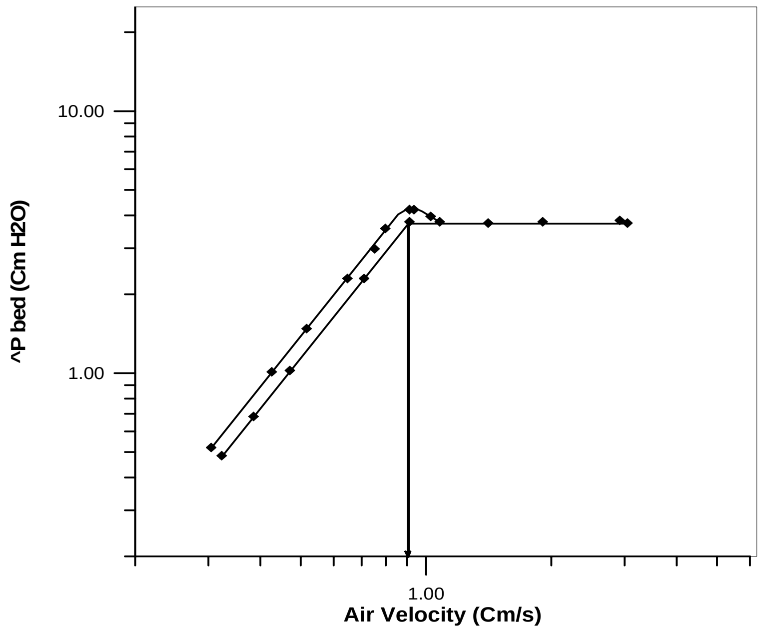

Figure (A.3): Bed Pressure Drop vs. Air Velocity
(Sand Particle Size = 165 Micron)

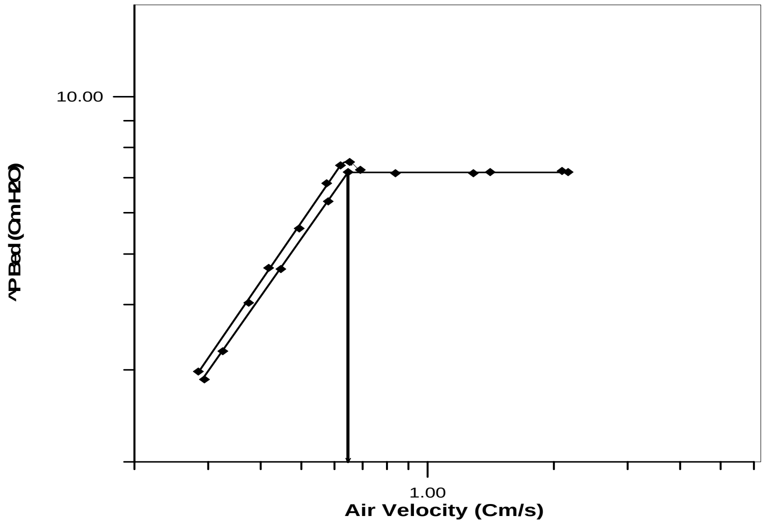

Figure (A.4): Bed Pressure Drop vs. Air Velocity
(Sand Particle Size= 112.5 Micron)

***Calibration of Air Flow meter:

The variable area rotameter has been calibrated for air at 101.3 KN /m^2 and 39 ^{0}C (ρ= 0.6013 kg / m^3 and ν= 1.326 * 10^{-5} m^3. s^{-1}).

The flow rate at other conditions may be calculated from:

$$\text{Actual flow rate} = \text{indicated flow rate} \times \sqrt{\frac{1.204}{\rho_m}}$$

Where ρ_m is the density of the air in the meter in kg/m^3 .

Note: from the ideal gas equation, $\rho = \dfrac{P}{RT}$

The following expression is the rotameter reading correction.

$$\sqrt{\frac{0.6013}{\rho_m}}$$

Dimensional Analysis:

The dimensionless group, *Y*, is a function of all the variables and dimensionless constant which influence particle motions. These factors may be arranged in a suitable form for dimensional analysis, such as

$$Y = f\,(\psi, d_p, (\rho_p - \rho), \rho, \mu, (G - G_{mf}), g) \quad \text{---- (A-1)}$$

$$Y = [\psi d_p]^a [(\rho_p - \rho)\rho]^b [G - G_{mf}]^c g^d \mu^e \quad \text{---- (A-2)}$$

The common groups for mass transfer are the Sherwood number, the Schmidt number, and the Reynolds number. The Sherwood number contains the mass transfer coefficient and the diffusivity. The Schmidt number contains only the physical properties of the fluid and its active component. The Reynolds number is, of course, a measure of flow rate.

The number of dimensionless group obtained from the dimensional analysis is equal to the number of variables, n=5, minus the number of fundamental dimensions, r=3, and hence two dimensionless groups will be obtained.

In term of fundamental dimensions:

$$1 = [L]^a [\frac{m^2}{L^6}]^b [\frac{m}{L^2\theta}]^c [\frac{L}{\theta^2}]^d [\frac{m}{L\theta}]^e \quad \text{--- (A-3)}$$

The exponents on each variable must be such that the group is dimensionless, so this requires that the following equations be satisfied:

For m $0 = 2b + c + e$ ---(1)

For L $0 = a - 6b - 2c + d - e$ ---(2)

For θ $0 = - c - 2d - e$ --- (3)

These three equations can be solved to give all the exponents in terms of two:

From (1) $c = - 2b - e$

From (3) $d = - 0.5c - 0.5e$

$= -0.5 (-2b-e) - 0.5 e$

$= b$

From (2) $0 = a - 6b - 2 (-2b-e) + b- e$

$= a + e - b$

$a =- e + b$

From these results we obtain

$$Y = [\psi d_p]^{-e-b}[(\rho_p - \rho)\rho]^b [G - G_{mf}]^{-2b-e} [g]^b [\mu]^e \qquad \text{--- (A-4)}$$

$$Y = [\frac{(\rho_p - \rho)\rho g (\psi d_p)}{(G - G_{mf})^2}]^b [\frac{(\psi d_p)(G - G_{mf})}{\mu}]^{-e} \qquad \text{--- (A-5)}$$

From equation above, one can notice that the first term is the invert of Froude number (*Fr*) and the second is the modified Reynolds number (*Re*).

Table (A-1) Experimental Conditions for mass Transfer in Empty Bed

Exp. No.	**Air Flow Rate** (m^3/hr)	**Temp.(C̊)**
1	2.8	39.0
2	3.4	39.1
3	3.7	39.2
4	4.0	39.0
5	4.5	39.1
6	5.2	39.0
7	2.8	51.3
8	2.8	55.3
9	2.8	66.1
10	3.4	51.2
11	3.4	55.2
12	3.4	66.1
13	3.7	51.2
14	3.7	55.3
15	3.7	66.0
16	4.0	51.1
17	4.0	55.3
18	4.0	66.2
19	4.5	51.3
20	4.5	55.3
21	4.5	66.2
22	5.2	51.2
23	5.2	55.3
24	5.2	66.2

*Air flow rate measured at ambient temperature.

Table (A-2) Experimental Results for Mass Transfer in Empty Bed

Exp. No.	Weight Loss (*gm/hr.m*^{2})	Re$_p$	Sh$_e$
1	16.0007	285.1003	17.3135
2	17.4398	345.8269	18.7098
3	18.3160	377.7627	19.3947
4	18.4545	406.7856	19.9931
5	19.6208	458.6925	21.0113
6	20.6403	529.8438	22.3181
7	46.3716	277.7204	17.1085
8	65.2329	272.8329	16.9799
9	148.507	266.4974	16.8026
10	49.6271	336.7652	18.4820
11	68.0483	330.7357	18.3388
12	160.042	323.0021	18.1415
13	51.7253	368.4121	19.1685
14	71.6804	362.3935	19.0308
15	163.612	350.6095	18.7530
16	52.5007	396.0089	19.7436
17	73.0415	388.2574	19.5742
18	171.623	378.8475	19.3535
19	56.4473	447.5064	20.7647
20	77.5178	440.0929	20.6113
21	179.185	424.9063	20.2874
22	59.3788	516.9524	22.0536
23	82.1496	507.9866	21.8811
24	189.438	489.2557	21.5112

Table (A-3) Experimental Conditions for mass Transfer in Fluidized Bed

Exp. No.	Air Flow Rate (m^3/hr)	Temp. (Ċ)	U/U_{mf}	Sand Mean Particle Size (*micron*)
1	4.0	39.1	1.081	215
2	4.4	39.1	1.189	
3	4.8	39.2	1.297	
4	5.2	39.2	1.405	
5	4.0	51.0	1.081	
6	4.4	51.0	1.189	
7	4.8	51.2	1.297	
8	5.2	51.3	1.405	
9	4.0	55.2	1.081	
10	4.4	55.1	1.189	
11	4.8	55.2	1.297	
12	5.2	55.0	1.405	
13	4.0	66.3	1.081	
14	4.4	66.1	1.189	
15	4.8	66.0	1.297	
16	5.2	66.3	1.405	

Exp. No.	Air Flow Rate(m^3/hr)	Temp.(Ċ)	U/U_{mf}	Sand Mean Particle Size (*micron*)
1	3.0	39.0	1.071	165
2	3.4	39.0	1.214	
3	3.8	39.2	1.714	
4	4.0	39.1	1.428	
5	3.0	51.1	1.071	
6	3.4	51.3	1.214	

7	3.8	51.0	1.714	
8	4.0	51.1	1.428	
9	3.0	55.0	1.071	
10	3.4	55.1	1.214	
11	3.8	55.3	1.714	
12	4.0	55.2	1.428	
13	3.0	66.0	1.071	
14	3.4	66.1	1.214	
15	3.8	66.2	1.714	
16	4.0	66.3	1.428	

Exp. No.	**Air Flow Rate (m^3/hr)**	**Temp. (℃)**	**U/U_{mf}**	**Sand Mean Particle Size (*micron*)**
1	2.4	39.0	1.091	112.5
2	2.8	39.1	1.272	
3	3.0	39.2	1.363	
4	3.2	39.0	1.454	
5	2.4	51.1	1.091	
6	2.8	51.2	1.272	
7	3.0	51.3	1.363	
8	3.2	51.0	1.454	
9	2.4	55.3	1.091	
10	2.8	55.2	1.272	
11	3.0	55.1	1.363	
12	3.2	55.0	1.454	
13	2.4	66.0	1.091	
14	2.8	66.1	1.272	
15	3.0	66.3	1.363	
16	3.2	66.2	1.454	

* Air flow measured at ambient temperature.

Table (A-4) Experimental Results for mass Transfer in Fluidized Bed

Exp. No.	Weight Loss ($gm/hr.m^2$)	Re_p	sh_e	Sand Mean Particle Size (*micron*)
1	30.550	943.711	254.612	215
2	31.754	964.753	258.791	
3	33.483	972.286	262.413	
4	34.911	982.705	269.751	
5	84.836	951.646	214.622	
6	89.614	967.903	225.704	
7	94.233	971.532	243.950	
8	98.408	979.463	257.568	
9	121.607	916.604	203.815	
10	124.457	945.031	217.780	
11	130.413	969.868	247.502	
12	132.781	972.372	227.235	
13	285.096	884.269	172.751	
14	295.131	989.403	194.224	
15	297.444	944.853	203.705	
16	313.011	961.185	214.443	

Exp. No.	Weight Loss ($gm/hr.m^2$)	Re_p	Sh_e	Sand Mean Particle Size (*micron*)
1	25.509	661.725	281.599	165
2	27.221	753.687	305.431	
3	29.075	839.761	334.890	
6	29.756	889.359	344.508	
5	73.244	648.804	264.114	
6	78.891	734.674	283.170	
7	81.180	822.972	295.800	

8	84.641	871.428	307.709	
9	100.240	640.279	247.105	
10	107.812	728.698	265.401	
11	115.400	814.634	281.715	
12	117.142	857.086	299.552	
13	240.074	629.500	227.746	
14	256.227	713.256	239.534	
15	270.089	782.938	252.753	
16	280.120	827.042	274.766	

Exp. No.	**Weight Loss (*gm /hr.m*^{2})**	**Re$_p$**	**Sh$_e$**	**Sand Mean Particle Size (*micron*)**
1	23.162	524.397	349.553	112.5
2	25.077	611.679	364.710	
3	25.833	650.192	377.455	
4	26.042	692.233	389.107	
5	64.938	505.762	327.114	
6	70.487	591.292	339.415	
7	74.378	639.272	359.770	
8	75.018	683.956	378.105	
9	91.653	500.889	314.211	
10	99.647	593.283	332.154	
11	100.780	630.218	351.005	
12	103.917	676.432	368.417	
13	212.427	489.986	305.215	
14	230.876	573.556	319.419	
15	682.796	930.620	335.498	
16	698.345	992.660	357.794	

الخلاصة

عنوان البحث : انتقال المادة من سطح النفثالين الغاطس الى طبقة (رمل_هواء) مميعة.

Mass Transfer of Naphthalene from an Immersed Surface to a Sand_air Fluidized Bed

ان عملية التميع واسعة الاستعمال في الصناعات العالمية المتعددة وبتكلفة تقدر ب(تريليون دولار). وتستعمل حاليا في عمليات :الفصل , التصنيف ,التجفيف وخلط الدقائق ,التفاعلات الكيمياوية وعمليات اعادة التصنيع. واحدة من هذه العمليات هي انتقال المادة من سطح غاطس الى طبقة غازية مميعة.

معامل انتقال المادة هو ثابت الانتشار والذي يتعلق بنسبة انتقال المادة , مساحة انتقال المادة ,ومحصلة التركيز والتي تمثل فاعلية الانتقال. ويمكن ان تحسب من عدة معادلات وعلاقات ونظريات مختلفة دالة على خصائص المادة.

تم استخدام عمود زجاجي بقطر داخلي (8 سم) وبطول (70 سم),وكان الغرض من هذا البحث هو ايجاد علاقة رياضية تربط معامل انتقال المادة الى طبقة غازية مميعة بعدة متغيرات تم دراستها وهي : درجة حرارة السطح(39-66C) معدل جريان الهواء($1.2-1.4U_{mf}$) , وحجم الدقائق(215,165,112.5 micron).

$$Sh = Sh_e + C_1\left[\frac{(\rho_p - \rho)\rho g(\psi d_p)}{(G - G_{mf})^2}\right]^{C_2}\left[\frac{(\psi d_p)(G - G_{mf})}{\mu}\right]^{C_3}$$

الثوابت هي: C_1, C_2, C_3 تساوي على التوالي 16.8574 , 0.07497 ,0.1284 وبنسبة خطأ تساوي 1.54%.

في هذا العمل وجد ان Sherwood: * له علاقة عكسية مع حجم الدقائق

* له علاقة طردية مع سرعة جريان المائع

* له علاقة عكسية مع درجة حرارة السطح

تم استعراض موديل رياضي لانتقال المادة من سطح غاطس الى طبقة مميعة لها القابلية على امتصاص المادة. وتم مقارنة نتائج الموديل مع النتائج العملية لهذا البحث و البحوث السابقة.

انتقال المادة من سطح النفثالين الغاطس الى طبقة (رمل - هواء) مميعة

رسالة
مقدمة إلى كلية الهندسة في جامعة نهرين
وهي جزء من متطلبات نيل درجة ماجستير علوم
في الهندسة الكيمياوية

من قبل

زهراء حميد محمد
(بكالوريوس علوم في الهندسة الكيمياوية 2004)

ذو الحجة 1428
كانون الثاني 2007