AN ANALYSIS FOR SIMULTANEOUS HEAT AND MASS TRANSFER IN A PNEUMATIC DRYER

by

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SUMMARY

An analysis is presented for simultaneous heat and mass transfer in a pneumatic dryer. The analysis can be used, either as a basis for determining the length of dryer necessary for a given drying load, or to determine the mass transfer coefficient for a particular set of conditions from appropriate experimental measurements. It was applied to the specific case for the pneumatic drying of granulated sugar and correlations are presented for both the heat and mass transfer coefficients, these being determined from measurements on an experimental dryer.

INTRODUCTION

The pneumatic dryer consists basically of a tube along which hot air is passed at such a rate that it can carry along with it the particles requiring to be dried. During the passage of the particles along the dryer, heat is transferred from the air to the particles and evaporation of moisture takes place from the particle surface. The residence time in this type of dryer is small, generally of the order of seconds, thus making it more suitable for drying materials in which the moisture is mainly on the surface of the particles. A full description of the practical aspects of this type of dryer has been given by Nonhebel and Moss[1].

In spite of its wide industrial use, there is little published work on the fundamental transfer processes which take place during the drying process. The work described in this paper is concerned with the

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development of a suitable analysis for the heat and mass transfer characteristics of the system and the applications of this analysis.

A study of this type of dryer was made because it is ideally suited to the drying of sugar crystals; these being small particles with little free internal moisture. The type of dryer normally used in the Caribbean region for sugar drying is the rotary dryer, which, apart from being large and expensive, requires that the sugar be conveyed using suitable extra equipment, from the centrifuges to the dryer, and from the dryer to the storage hopper. The pneumatic dryer has the following advantages over the rotary dryer:

(i) It conveys and dries simultaneously

(ii) It contains no moving parts, thus giving low maintenance costs

(iii) It takes up little space

(iv) It has a low capital cost

Despite these advantages over the more traditional equipment, a literature survey revealed only one reference[2] to the pneumatic drying of sugar, and the work described there was empirical in nature.

THEORETICAL APPROACH

The theoretical analysis is based on a series of numerical differential heat and mass balances, made successively along the dryer, starting at the inlet end and finishing at the exit.

The basic assumptions made in the analysis are as follows:

(i) The temperatures in the system are such that radiation effects can be neglected.

(ii) The particles are of sufficiently small size and/or large thermal conductivity to eliminate internal particle temperature gradients.

(iii) The particles are of uniform size and are uniformly distributed throughout the field of flow.

(iv) The moisture is present on the surface of the particles only and exerts its vapour pressure at all times.

(v) The particles are uniformly wet.

In the analysis, the mass transfer coefficient used is based on the humidity difference driving force. Thus, an expression for the saturation humidity as a function of temperature is necessary in order to determine the humidity difference driving force. The basic definitions of the saturation hunidity in terms of the vapour pressure of water vapour is given by:-

$$Y_s = \left[\frac{M_w}{M_a}\right] \cdot \left[\frac{p}{p_t - p}\right]$$

The vapour pressure is a function of temperature and may be obtained from the general relationship suggested by Calingaert and Davis[3].

$$\mathrm{In}\ p_s = A - (B/\{t_s + 230.3\})$$

where A and B are constants

Thus the general relationship for the saturation humidity in terms of temperature and pressure is given by:-

$$Y_s = \frac{(M_w/M_a)\ \exp\ (A - \{B/(t_s + 230.3)\})}{p_T - \exp\ (A - \{B/(t_s + 230.0)\})} \quad \ldots\ldots \quad (1)$$

The physical situation assumed to exist in a pneumatic dryer, is that where the hot air carrying the particles along with it, transfers heat to the colder particles. This heat simultaneously heats up the particles and evaporates moisture from the particle surface. If the dryer is long enough, then the particle heating portion may become negligible. Since the dryer lagging will never provide perfect insulation, heat loss through the walls must be included in the analysis.

Reference to Fig. 1 shows that the following heat and mass balances may be made on a typical differential section, dl, in the dryer:-

Overall moisture balance on dl

$$G_s dY = L_s dX \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (2)$$

Moisture balance on gas side

$$G_s dY = k_y a_p (Y_s - Y)\ dl \quad \ldots\ldots\ldots\ldots\ldots\ldots \quad (3)$$

Heat balance on the gas side

$$G_sC_gdt_g = h_ya_p\ (t_g - t_s)\ dl + Ua_t\ (t_g - t_a)\ dl \quad \ldots\ldots\ldots (4)$$

Heat balance on the particles

$$L_sC_sdt_s = h_ya_p\ (t_g - t_s)\ dl - k_ya_p\lambda\ (Y_s - Y)\ dl \quad \ldots\ldots\ldots (5)$$

It is not possible to carry out direct integrations over the full length of the dryer, because of the interactions of the variables involved. Thus, it is necessary to use these balances as a basis for a numerical methods of analysis.

In the special case where no drying takes place in the system, the particles are simply heated up during passage along the dryer. Thus, equations (2) and (3) are not necessary in the analysis. Also, the mass transfer term on the right hand side of equation (5) disappears to give:-

$$L_sC_sdt_s = h_ya_p\ (t_g - t_s)\ dl \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots (6)$$

The two useful differential balances for the use of heat transfer only, are thus equations (5) and (6).

GENERAL APPLICATION OF THEORY

The differential balances given by equations (2) to (5) in conjunction with equation (1) can be used as the basis of numerical methods using the computer to either:-

(i) calculate the mass transfer coefficient for a particular situation from appropriate experimental measurements.

or (ii) calculate the length of dryer necessary for a particular duty.

In each case, the basic method is to split up the whole length of the dryer up into a large number of small differential sections, dl, of known length. The balances given by equations (2) to (5) in conjunction with equation (1) are carried out step by step, from the inlet end of the tube to the outlet.

In the case where it is necessary to calculate a mass transfer coefficient from experimental measurements, a trial and error technique is necessary. The simplest technique is to start by assuming a

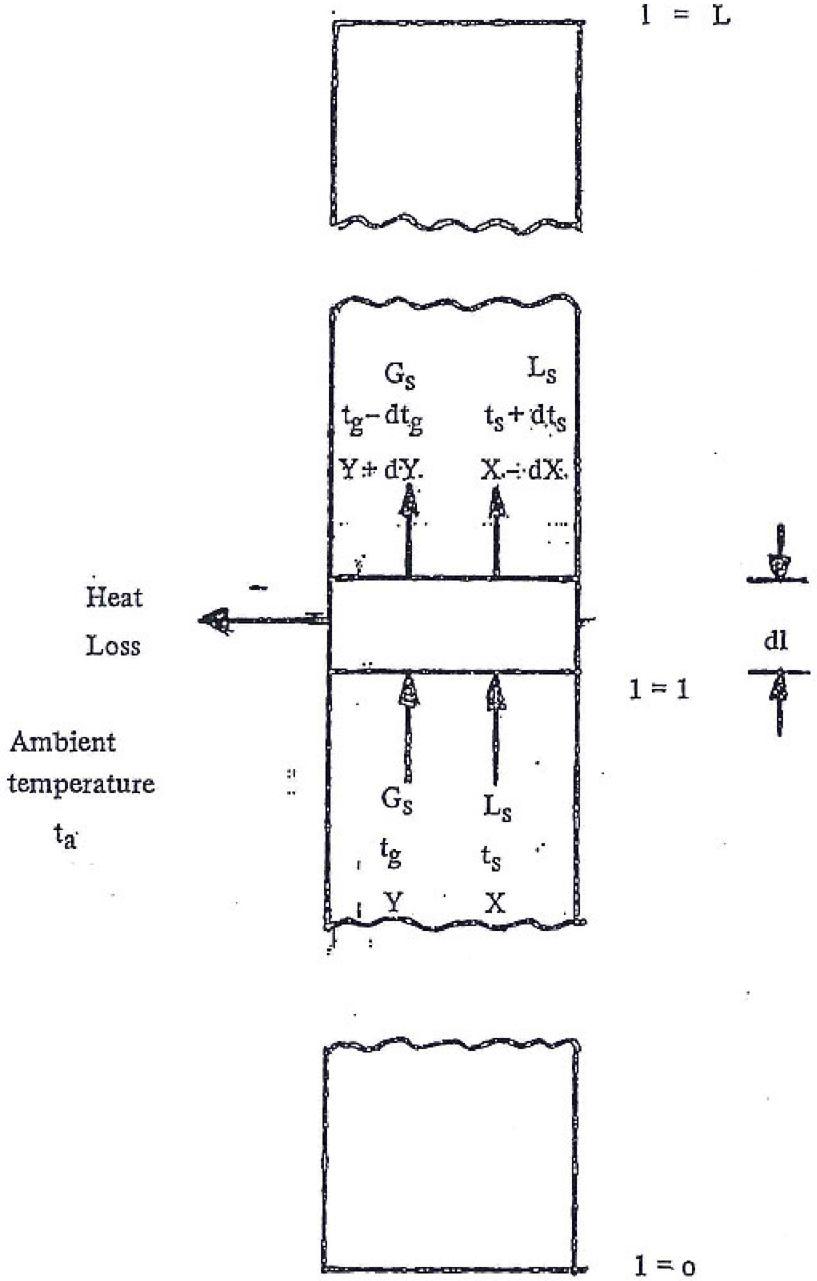

Fig. 1: Differential Heat and Mass Balances.

value for the mass transfer coefficient which is obviously too low. The step by step numerical balances are carried out, using this coefficient together with the known heat transfer coefficient and the appropriate experimental readings, in order to calculate the outlet moisture content of the particles. This will be found to be lower than the measured outlet particle moisture content. The assumed mass transfer coefficient is then successively increased until the calculated outlet moisture content is equal to the measured value. A general technique for carrying this out is illustrated in a typical computer programme flow diagram shown in Fig. 2. More sophisticated trial and error techniques than that illustrated may be used, if required.

A similar technique may be used for the determination of the heat transfer coefficient from experiments where no drying takes place. In this special case however, only differential heat balances will be carried out, so only equations (4) and (6) will be used. Either of the calculated outlet air or particle temperatures may be balanced against the experimental measurements, as the criterion for determination of the correct heat transfer coefficient.

The difference in the design case is that no trial and error would normally be necessary, since the mass transfer coefficient must be known. The programme will carry out a series of differential balances until the required outlet moisture content is reached. The length of dryer required can then be determined from the number of balances carried out and the chosen differential length. If it is known that the mass transfer coefficient varies along the dryer in some way; for example to take into account different conditions in the solids entry region, then it can be incorporated appropriately into the programme.

SPECIAL APPLICATION - PNEUMATIC DRYING OF GRANULATED SUGAR

The work was carried out specifically to investigate the pneumatic drying characteristics of sugar. Unfortunately, no data was found for either the heat or mass transfer coefficients for this system. It was thus necessary to use the theory to calculate and correlate the coefficients

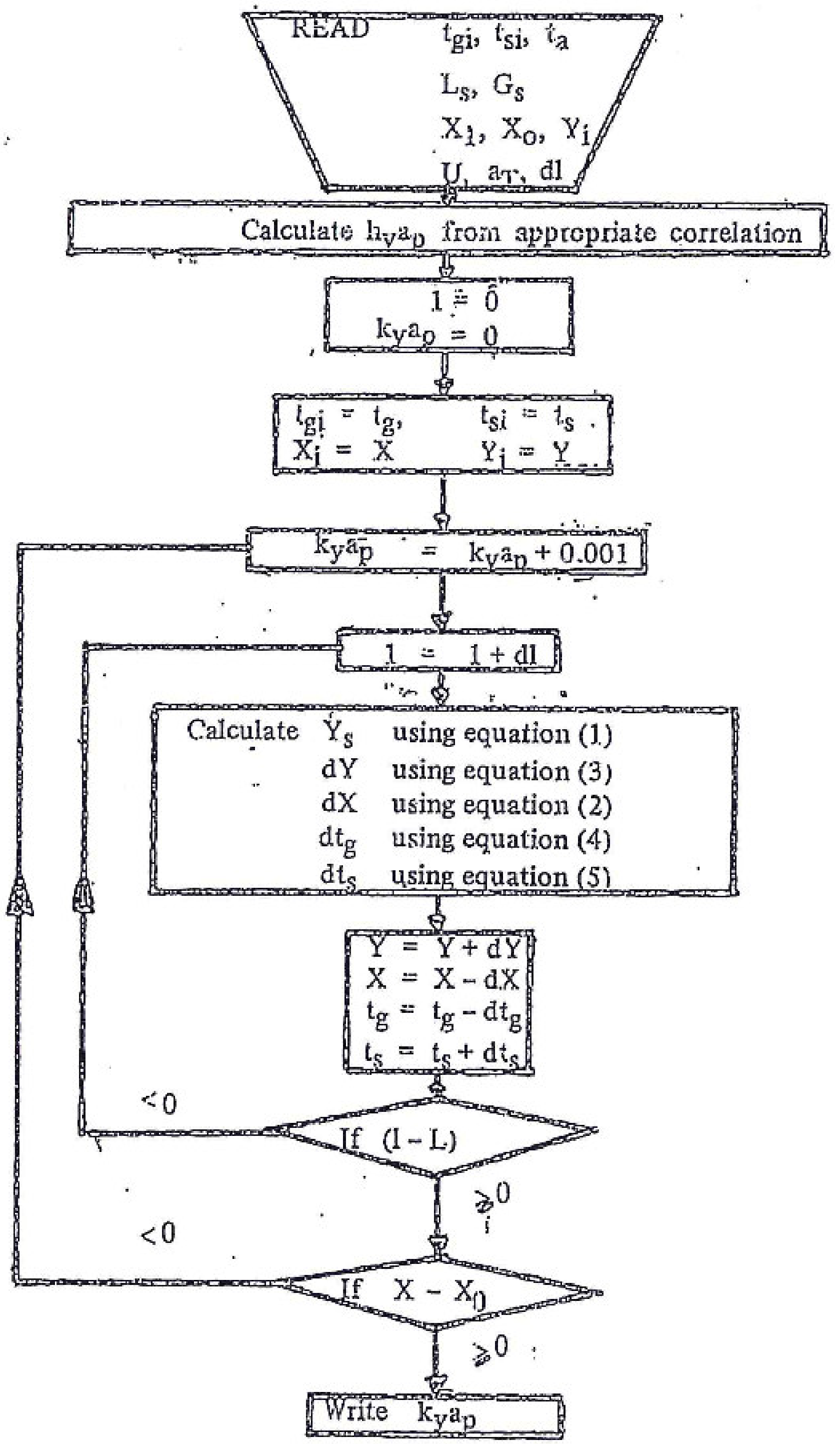

Fig. 2: Flow Diagram for Computer Programme to Determine Mass Transfer Coefficient.

for the system from suitable experimental data. A pneumatic dryer was built and operated for this purpose. A full description of this dryer together with the sampling and measurement techniques as well as the experimental conditions and readings for the drying experiments have been given elsewhere[4,5]. The material used in the experimental programme was granulated sugar, manufactured and donated by Caroni Ltd., Trinidad.

Two separate series of experiments were carried out, one for heat transfer only, and one for simultaneous heat and mass transfer. In each series of experiments, the variables investigated were the flow rates of air and sugar.

(a) Heat Transfer Experiments

In this series of experiments, the particles were dried before use, and the measurements were used to calculate the volumetric heat transfer coefficient $h_y a_p$. The following correlation was derived:-

$$\frac{h_y a_p L}{k d_p} = 232.6 \left[\frac{\rho u d_p}{\mu}\right]^{1.22} \left[\frac{L_s}{G_s}\right]^{0.38} \quad \ldots\ldots\ (7)$$

The range of variables investigated was:-

$\frac{L_s}{G_s}$:- 0.65 to 1.98

$\frac{\rho u d_p}{\mu}$:- 265 to 568

(b) Drying Experiments

In this series of experiments, the sugar crystals were wetted before use and the measurements were used in conjunction with the above heat transfer coefficient correlation equation (7) to calculate the volumetric mass transfer coefficient, $k_y a_p$. The following correlation was derived:-

$$\frac{k_y a_p L}{D_m \bar{m} d_p} = 2.02 \times 10^{11} \left[\frac{\rho u d_p}{\mu}\right]^{-2.52} \left[\frac{L_s}{G_s}\right]^{2.47} \quad \ldots\ldots\ (8)$$

The range of variables investigated was:-

$\frac{L_s}{G_s}$:- 0.51 to 1.42

$\frac{\rho u d_p}{\mu}$:- 308 to 495

The series of drying experiments showed that granulated sugar was well suited to pneumatic drying. The design of a dryer for granulated sugar could be made using the theory presented together with the derived equations (7) and (8) for the transfer coefficients provided the required conditions more similar to those used in the experimental programme.

NOMENCLATURE

Notation	Description	Units	Dimension
G_s	Gas flow rate (dry basis)	lbs/hr	MT^{-1}
L_s	Solids flow rate (dry basis)	lbs/hr	MT^{-1}
U	Overall heat loss transfer coefficient	CHU/hr.ft^2 °C	$HL^{-2}T^{-1}\theta^{-1}$
a_t	Cross sectional area of tube	ft^2	L^2
t	Temperature	°C	θ
L	Length of column	ft.	L
C_s	Specific heat of solids	CHU/lb°C	$HM^{-1}\theta^{-1}$
C_g	Specific heat of gas at constant pressure	CHU/lb°C	$HM^{-1}\theta^{-1}$

NOMENCLATURE CONT'D

h	Heat transfer coefficient: h_y - gas to particle heat transfer coefficient	CHU/hr.ft^2 °C	$HL^{-2}T^{-1}\theta^{-1}$
a_p	Particle surface area/unit length of column	ft^2/ft	L
$h_y a_p$	Gas to particle volumetric heat transfer coefficient	CHU/hr.ft°C	$HL^{-1}T^{-1}$
k_y	Mass transfer coefficient	lbs/hr.ft^2	$ML^{-2}T^{-1}$
$k_y a_p$	Volumetric mass transfer coefficient	lbs/hr.ft	$MT^{-1}L^{-1}$
Y	Humidity of gas	lbs water/lbs dry air	–
dY	Differential change in humidity	lbs water/lbs dry air	–
Y_s	Saturation humidity of gas	lbs water/lbs dry air	–
X	Solids moisture content	lbs water/lbs dry solids	–
dX	Differential change in solids moisture content	lbs water/lbs dry solids	–
M_w	Molecular weight of water	–	–
M_a	Molecular weight of air	–	–
A	Constant		
B	Constant		
p_s	Partial pressure of water vapour	lbs$_f$/in^2	$ML^{-1}T^{-2}$
p_T	Total pressure in system	lbs$_f$/in^2	$ML^{-1}T^{-2}$
$N_{Nu} - \frac{h_y d_p}{K}$	Nusselt number	–	–

NOMENCLATURE CONT'D

$N_{Re} - \frac{\rho u d_p}{\mu}$	Reynolds number	–	–
μ	Gas viscosity	lbs/ft.hr.	$ML^{-1}T^{-1}$
K	Thermal conductivity	CHU/hr.ft^2 °C/ft	$HL^{-1}T^{-1}\theta^{-1}$
ρ	Density of gas	lbs/ft^3	ML^{-3}
u	Gas velocity	ft/hr	LT^{-1}
d_p	Particle diameter	ft.	L
d	Tube diameter	ft.	L
$N_{sh} - \frac{k_y d_p}{D_m}$	Sherwood number	–	–
D_m	Molal Gas Diffusivity	lb. moles ft. hr.	$ML^{-1}T^{-1}$
$N_{ph} - \frac{h_y a_p L}{k d_p}$	Modified Nusselt number	–	–
$N_{pm} - \frac{k_y a_p L}{D_m \bar{m} d_p}$	Modified Sherwood number	–	–
$\bar{m}$	Molecular weight	–	–
$N\alpha - \frac{L_s}{G_s}$	Solids to gas ratio	–	–
λ	Latent heat of vaporisation	CHU/lb.	H/M
d_s	Mean particle based surface area	ft.	L
dl	Differential length of column	ft.	L

dA	Differential cross sectional area	ft^2	L^2

Subscripts

o	outlet conditions
a	air
i	inlet conditions
P	particle
S	solids
W	Water
t	tube
T	total
col.	column

REFERENCES

1. Nonhebel G. and Moss A.A.M. *Drying of Solids in the Chemical Industry* Butterworths, London 1971 p. 221 - 252.
2. Lyudboshites I.K. et al. *Inzh. Fiz. Zhurn* 14 (4) 726, 1968.
3. Calingaert G. and Davis D.S. *Ind. Eng. Chem.* 17 1287, 1925.
4. Shah L.A. and Mc Gaw D. R. *Paper presented to West Indian Sugar Association Conference,* Barbados, May 1973.
5. Shah L.A. *M.Sc. Thesis,* U.W.I., 1973