Fenômenos de Transportes 3 (ZEA0764)

Transferência de Massa por Convecção Forçada II

Profs. Responsáveis:
Paulo José do Amaral Sobral

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Tópicos

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I. Introdução

Continuamos interessados no cálculo de k_m para poder usar as equações da convecção mássica:

$$\vec{n}_{A,z} = k_m(\rho_{As} - \rho_{A\infty})$$

ou

$$\vec{N}_{A,z} = k_m(C_{As} - C_{A\infty})$$

Existem muitas correlações para o cálculo de Sh ou j_M.

INTERNATIONAL JOURNAL OF FOOD PROPERTIES, 4(3), 373–382 (2001)

II. Revisão de Krokida et al. - alimentos

MASS TRANSFER COEFFICIENT IN FOOD PROCESSING: COMPILATION OF LITERATURE DATA

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ABSTRACT

Mass transfer coefficient data in food processing were retrieved from literature and classified per process and material. Most of the data were available in the form of empirical equations using dimensionless numbers. All available empirical equations were transformed in the form of mass transfer factor versus Reynolds number ($j_M = a\mathrm{Re}^n$). Average equations for each process are also proposed.

INTRODUCTION

The interface mass transfer coefficient is important in designing of food processes and processing equipment, and in the control of food packaging and storage. Mass transfer coefficients are essential in designing drying, storage and separation processes. One basic feature of mass transfer coefficient is that it is affected strongly by the characteristics of the processing equipment.

The surface mass transfer coefficient can be defined using the following equation:

$$J = h_M A(X_A - X_{AS}) \tag{1}$$

where h_M (kg/m^{2}s) is the surface mass transfer coefficient at the material-air interface, J (kg/s) is the rate of mass transfer, A (m^2) is the effective surface

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Table 3. Retrieved Data for Each Food Material

Drying Convective Carrot
Cubes in basket $T = 30\,°C$, $G = 1000–9000\ kg/m^2\,h$
$Sh/(Re\ Sc^{1/3}) = 0.692\ Re^{-0.486}$
$500 < Re < 5000$
Mulet et al. [3]

Drying Convective Grapes
Seedless Sultana grapes in rectangular metal pan $T = 50–76\,°C$, $V = 2.0–2.5\ m/s$
$Sh = 0.74\ Re^{0.57}\ Sc^{0.33}$
$900 < Re < 3000$
Vagenas et al. [4]

Drying Convective Grapes
Air drying of grapes in trays with mixed air flow
$Ta = -50 \sim 70\,°C$, $Tp = 26 \sim 70\,°C$, $V = 0.03\ m/s$ (estimated)
$K_M = h/(H_L^{64.7})(m/s)$, $K_M = k_C\ (m/s)$
h is calculated from the relation: $Nu = 0.664\ Re^{0.5}\ Pr^{0.33}$,
HL = latent heat of water vaporization (J/kg), $10 < Re < 40$
Ghiaus et al. [5]

Drying Convective Maize
Fluidized bed of maize-sand mixture $T = 60–120\,°C$
$Sh = 34.565\ Sc^{0.33}$
$5 < Re < 15$
Mourad et al. [6]

Drying Convective Rice
Crossflow moving bed
$j_M = (2.06\ Re^{-0.575})/\varepsilon$
ε bed porosity, $20 < Re < 1000$
Torrez et al. [7]

Drying Convective Corn
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Krokida et al. (2001)

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Drying Spray Milk

Milk powder $Ta = 185\,°C$, $Tp = 25 \sim 75\,°C$

$$Sh = 2 + 0.58\ Re^{0.5}\ Sc^{0.33}$$

$1 < Re < 2$

Straatsma [8]

Freezing Forced Convection Meat

Beef hamburgers and meatballs refrigerated on a conveyor belt tunnel, with various types of air flow $Ta = -35\,°C \sim -25\,°C$, Tp (initial) $= -60\,°C$, $u = 1 \sim 7.5\ m/s$

$$Sh = Nu(Sc/Pr)^{1/3}, Nu = a\ Re^{b}\ Pr^{0.33}$$

Values of *a* and *b* constants are given in Heat Transfer Coefficient database for various geometries and flow types. Equations for the evaluation of air properties are given in the paper. $2500 < Re < 25000$

Tocci et al. [9]

Krokida et al. (2001)

Table 3. Continued

Storage Forced Convection Potatoes

Cooling and storage of a cylindrical bed of potatoes in aerated silos $Ta = 6.7\,°C$, $Tp = 6.7 \sim 15.5\,°C$, $V = 0.011$ m/s

$$Sh = (0.5\,Re^{1/2} + 0.2\,Re^{2/3})Sc^{1/3}$$

Bed, height = 2.4 m, diam. = 0.7 m, Dimensions of potato tube $(L \times D) = 95 \times 51\,mm^2$, d_p = equivalent potato diam. = 65 mm, ε = porosity = 0.61, L^* = characteristic length in Reynolds number = $d_p\,\varepsilon/(1-\varepsilon)$, $10 < Re < 104$

Xu et al. [10]

Sterilization Forced Convection Model Food

Heat transfer from fluid (tap water) to particles (sucrose coated polysterene spheres) flowing into a tube

$$Sh = -110.4\,t + 215.1$$

t = dimensionless residence time = 0.85 ~ 0.97, $Sh = h_M\,d/DAB$, tube diam. = 41 mm, d = particle diam. = 6 mm, DAB = diffusivity of sucrose/water = 5.24 10-10 m^2/s

Fu et al. (11)

Krokida et al. (2001)

Table 4. Parameters of the Equation $j_M = a\mathrm{Re}^n$ for Each Process and Each Material

Process/Product/Reference	a	n	min Re	max Re
Drying Convective				
Corn				
Torrez N. et al., 1998	5.15	– 0.575	20	1000
Grapes				
Ghiaus A.G. et al., 1997	0.004	– 0.462	10	40
Vagenas G. et al., 1990	0.741	– 0.430	900	3000
Maize				
Mourad M. et al., 1997	34.6	– 1.000	5	15
Rice				
Torrez N. et al., 1998	5.15	– 0.575	20	1000
Carrot				
Mulet A. et al., 1987	0.69	– 0.486	500	5000
Drying Spray				
Milk				
Straatsma J. et al., 1999	2.947	– 0.890	1	2
Freezing				
Meat				
Tocci A.M. et al., 1995	2.496	– 0.495	2500	70,000
Storage				
Potatoes				
Xu Y. et al., 1999	0.667	– 0.428	50	55
Sterilization				
Model Food				
Fu W.R. et al., 1998	11.220	– 1.039	6500	26,000

Krokida et al. (2001)

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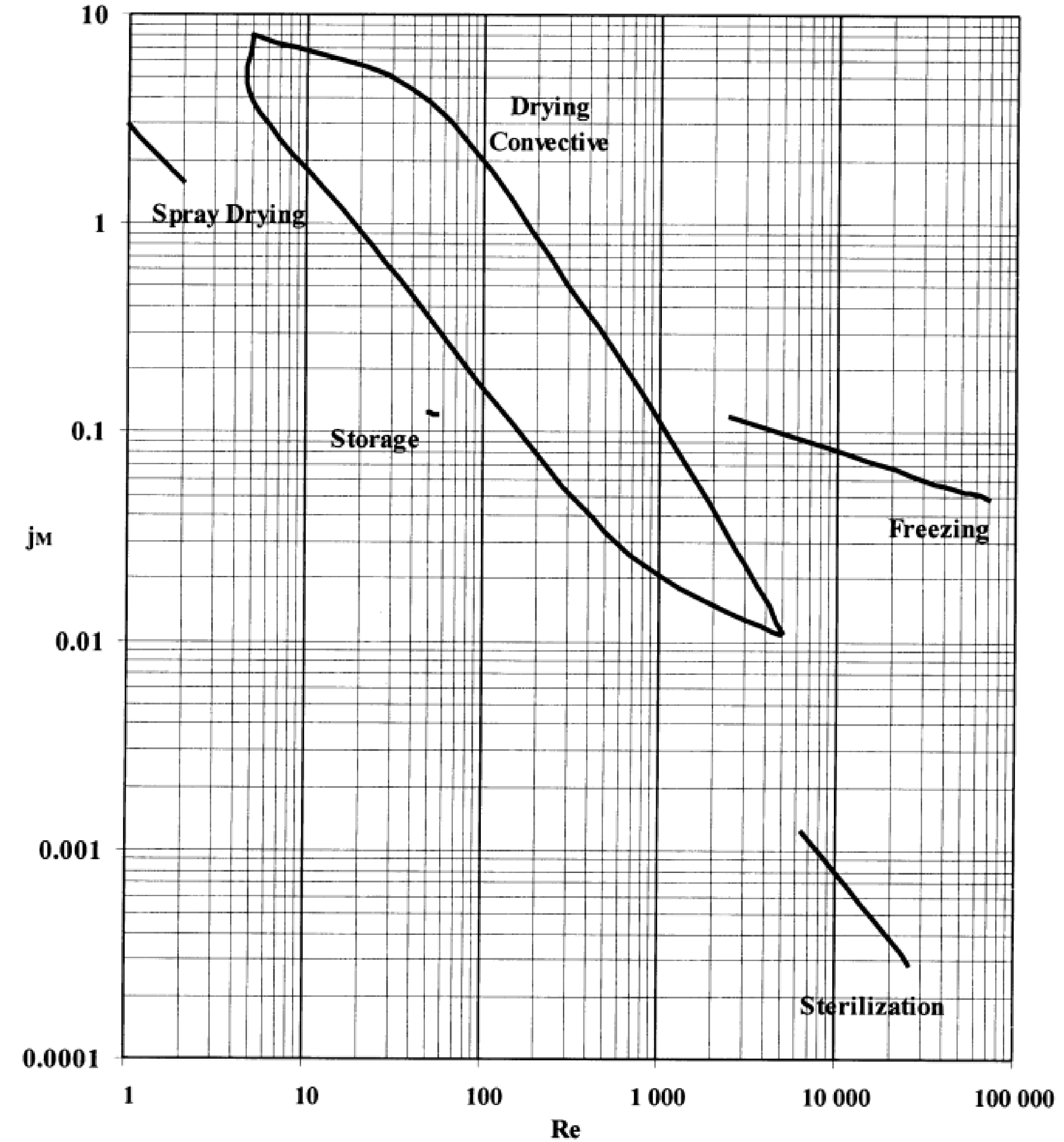

Figure 2. Ranges of variation of the mass transfer factor versus Reynolds Number for all the examined processes.

Krokida et al. (2001)

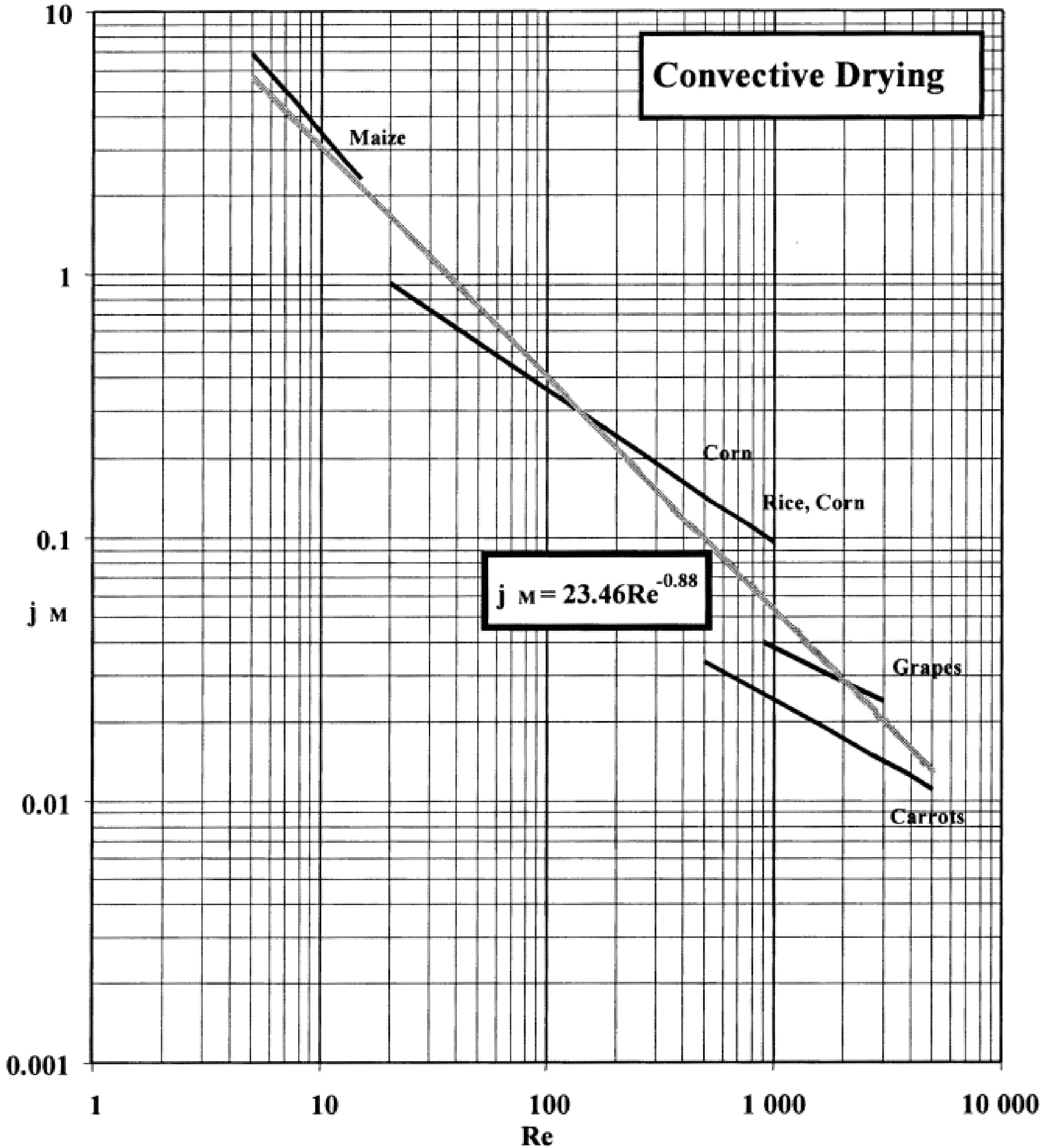

Figure 3. Mass transfer factor versus Reynolds Number for convective drying process and for various materials.

Krokida et al. (2001)

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Table 5. Parameters of the Equation $j_M = a\mathrm{Re}^n$ for Each Process

Process	a	n	min Re	max Re
Drying/convective	23.5	– 0.882	5	5000
Drying/spray	2.95	– 0.889	1	2
Freezing	0.10	– 0.268	2500	70,000
Storage	0.67	– 0.427	50	55
Sterilization	11.2	– 1.039	6500	26,000
Generalized equation	1.11	– 0.540	1	70,000

Krokida et al. (2001)

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III. Trabalho de Tocci & Mascheroni. - carne

Heat and Mass Transfer Coefficients During the Refrigeration, Freezing and Storage of Meats, Meat Products and Analogues

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ABSTRACT

The existing bibliographical data on heat and mass transfer coefficients during refrigeration, freezing and storage of meat and meat products were reviewed.

Heat transfer coefficients for meat balls and hamburgers were determined experimentally in a prototype belt freezer. Measurements were carried out at different air velocities and directions of air flow. In each case, the coefficients thus obtained were correlated with working conditions.

Mass transfer data for the preceding cases were calculated from the heat transfer coefficients.

NOTATION

C_p	Heat capacity (J/kg K)
d	Diameter (m)
D_a	Diffusion coefficient of water vapour in air (m^2/s)
h	Heat transfer coefficient (W/m^2 K)
k	Thermal conductivity (W/m K)
K	Mass transfer coefficient (kg/m^2 s)
L	Latent heat of sublimation of water (J/kg)
Nu	Nusselt number ($= hd/k_a$)

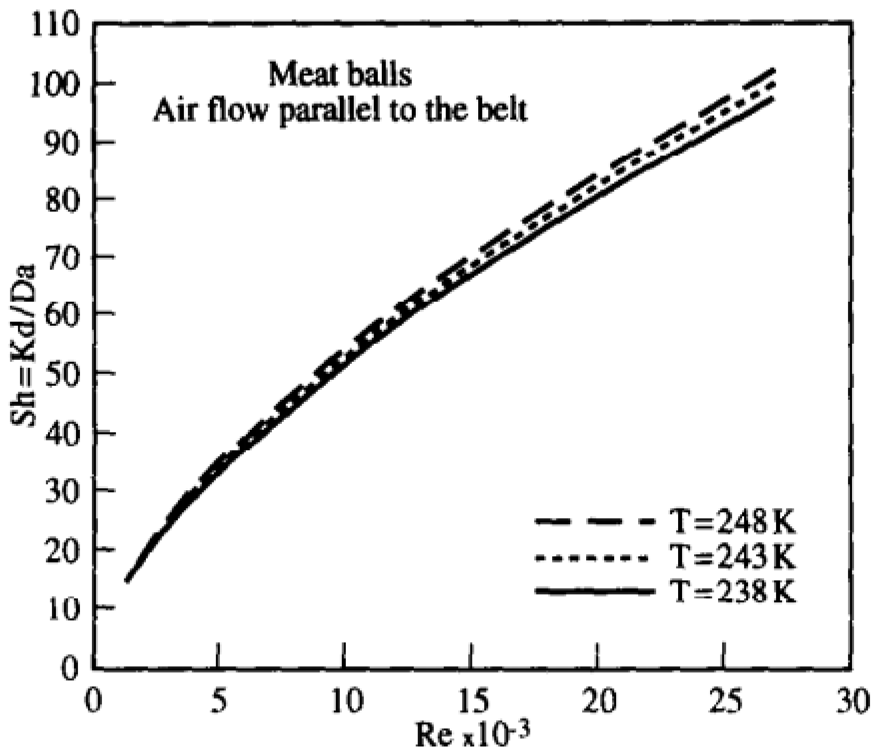


Fig. 3. Predicted variation of *Sh* with *Re* for meat balls at different air temperatures and for air flow parallel to the belt.

Tocci & Mascheroni (1995)

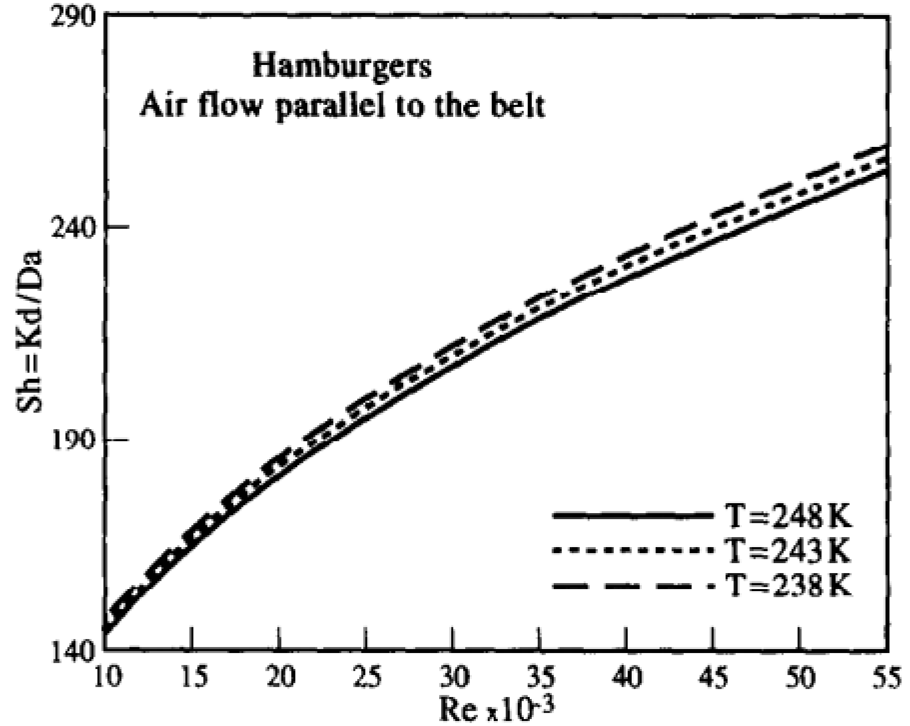


Fig. 4. Predicted variation of *Sh* with *Re* for hamburgers at different air temperatures and for air flow parallel to the belt.

Tocci & Mascheroni (1995)

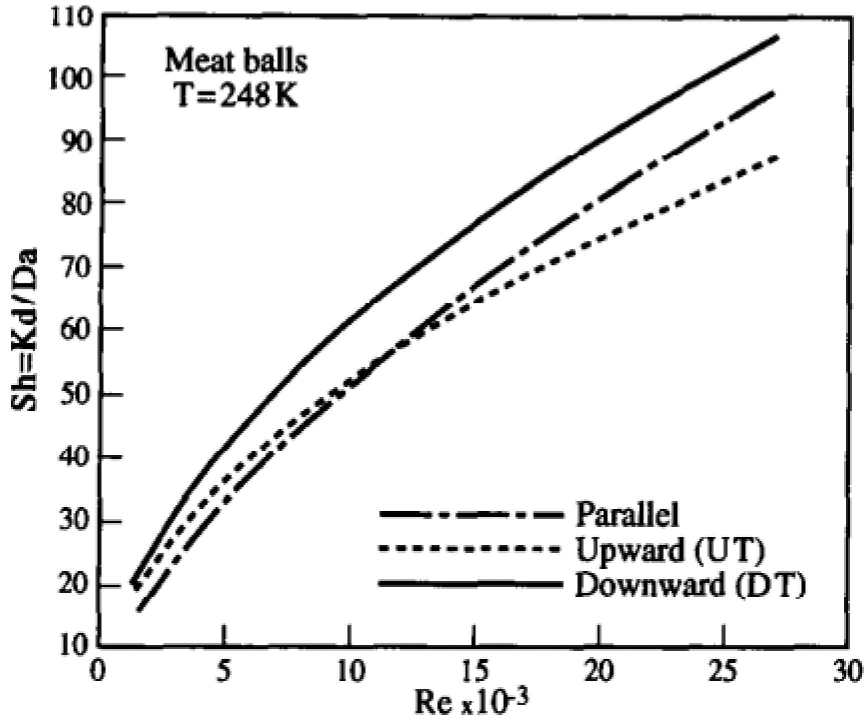
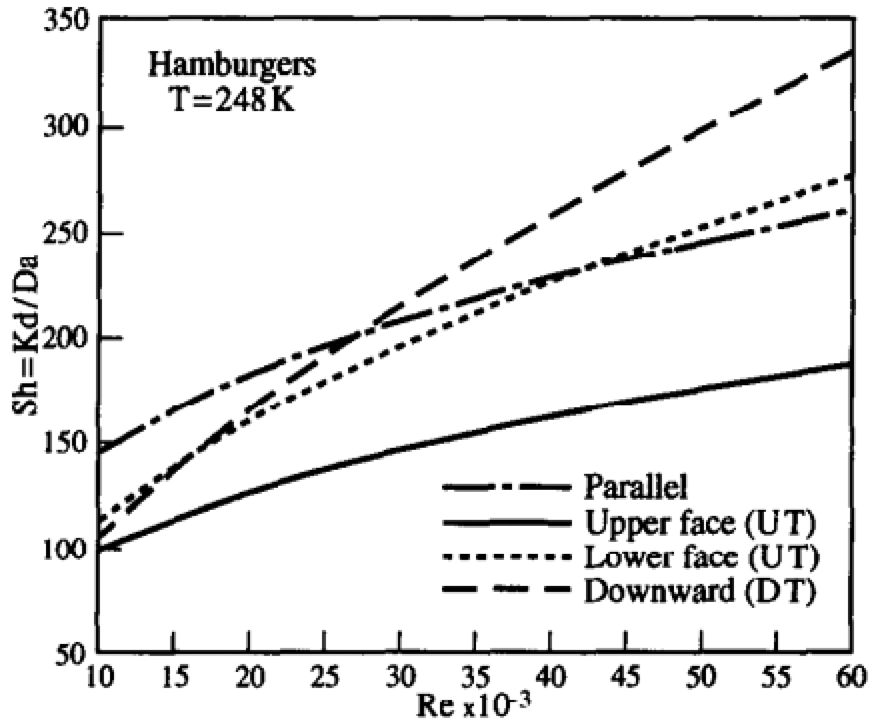


Fig. 5. Predicted variation of *Sh* with *Re* for meat balls at $T_a = 248$ K and for different types of air flow.

Tocci & Mascheroni (1995)



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IV. Trabalho de Shi et al. – extração CO2 SC

Correlation of mass transfer coefficient in the extraction of plant oil in a fixed bed for supercritical CO_2

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Abstract

The estimation of mass transfer coefficients in supercritical fluid extractions for fixed bed systems has been investigated. Two types of correlations are discussed: $Sh' = m(Re^{1/2}Sc^{1/3})^n$ and $Sh' = m(Re^{n_1}Sc^{n_2})$. Two statistical methods, the weighted least squares method and the error-in-variable-model (EVM), were used to estimate the model parameters based on published data, respectively. The linear least squares method is not suitable theoretically as the estimated error for $Re^{1/2}Sc^{1/3}$ is larger than the error of Sh'. The estimation result based on the data of Puiggene et al. was revised as $Sh' = 0.422(Re^{0.580}Sc^{0.3074})$ when $10 < Re < 100$. It shows consistent kinetic behavior in this region according to their data and it is also validated by the other literature data. The estimation also shows that it is acceptable to account the exponent of Sc as 1/3.
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Keywords: Mass transfer coefficient; Supercritical fluid extraction; Parameter estimation; Essential oil

1. Introduction

Supercritical carbon dioxide ($SC\text{-}CO_2$) extraction has been widely used to extract essential oils from various plants in food industry (Rozzi & Singh, 2002; Williams, 1981). Fixed-bed apparatus has been often used for this purpose (Brunner, 1984; Lee, Bulley, Fattori, & Meisen, 1986; Sovová, 1994).

One of the important parameters to describe the extraction kinetics is the mass transfer coefficient in the supercritical fluid phase. An accurate method for predicting the mass transfer coefficient is required to develop process modeling and quality control on line. Correlation equations using characteristic dimensionless numbers are often used for this purpose. It means that the pure chemical system can be used to simulate the oil mixture system and obtain more accurate physical properties. Ferreira (1996) proposed correlation equation in terms of Sherwood number (Sh), Rayleigh number ($Ra \equiv GrSc$), Reynolds number (Re) and Grashof number (Gr), and combined the effects of natural and forced convection together as follows:

$$Sh = 1.451Ra^{1/4}\left(\frac{Re}{Gr^{1/2}}\right)^{0.525} \quad (1)$$

where Sh is Sherwood number, Sc is Schmidt number, Re is Reynolds number, and Gr is Grashof number.

The parameters of Eq. (1) were estimated for pepper oil system in the presence of mixed convection. If Ra is expanded as $GrSc$, the exponent of Gr is much smaller than Re and Sc. That reveals a weak correlation between Sh and Gr.

Based on an assumption that the natural and forced convection are independent, their effects can be combined in a linear manner. Lee and Holder (1995) proposed the following correlation equation:

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Taking into account that the natural convection is very weak, the correlation can be simplified as the first item of the right side of Eq. (2):

$$Sh = m(Re^{1/2}Sc^{1/3})^{n} \quad (3)$$

or

$$Sh = m(Re^{n_1}Sc^{n_2}) \quad (4)$$

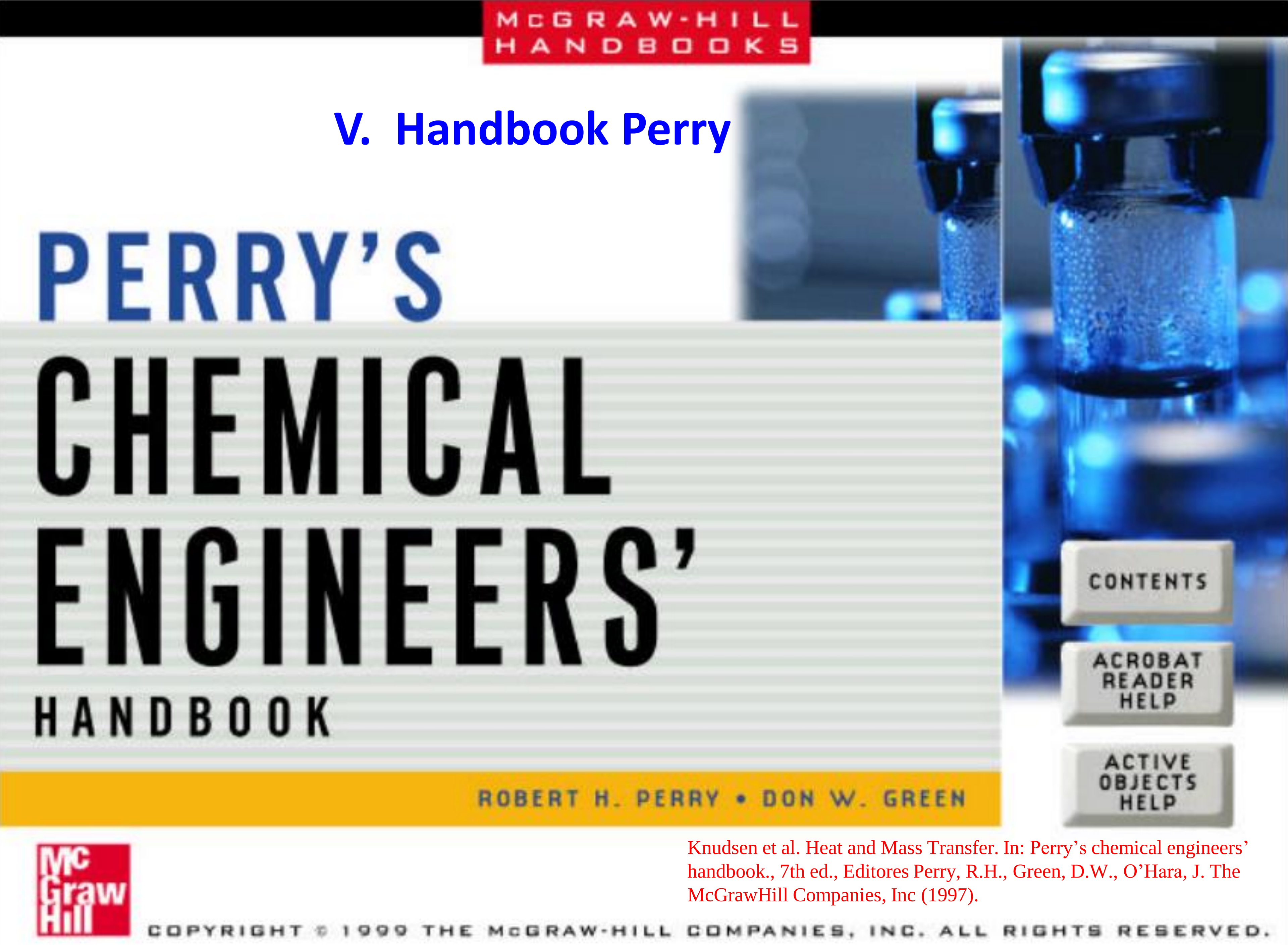
This assumption is reasonable for essential oil extraction.

Table 1
Mass transfer correlation parameters for essential oils at SC-CO_2 extraction

Sources	Parameters of mass transfer correlation			Range
	m	n_1	n_2	
Catchpole (1993)	0.82	0.66	1/3	$1 < Re < 70$, $3 < Sc < 11$
Tan et al. (1988)	0.380	0.83	1/3	$2 < Re < 40$, $2 < Sc < 20$
Puiggene et al. (1997)	0.206	0.80	1/3	$10 < Re < 100$, $3 < Sc < 20$
King et al. (1997)	0.2548	0.5	1/3	$Re < 1$, $70 < Sc < 100$

King et al. (1997), Puiggene et al. (1997) and Tan et al. (1988) used apparent Sherwood number, Sh'.

V. Handbook Perry



Knudsen et al. Heat and Mass Transfer. In: Perry's chemical engineers' handbook., 7th ed., Editores Perry, R.H., Green, D.W., O'Hara, J. The McGrawHill Companies, Inc (1997).

Perry's Chemical Engineers' Handbook

Knudsen et al. (1997).

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TABLE 5-21 Mass Transfer Correlations for a Single Flat Plate or Disk—Transfer to or from Plate to Fluid

TABLE 5-22 Mass Transfer Correlations for Falling Films with a Free Surface in Wetted Wall Columns—Transfer between Gas and Liquid

TABLE 5-23 Mass-Transfer Correlations for Flow in Pipes and Ducts—Transfer is from Wall to Fluid

TABLE 5-24 Mass Transfer Correlations for Flow Past Submerged Objects

TABLE 5-25 Mass-Transfer Correlations for Drops and Bubbles

TABLE 5-26 Mass-Transfer Correlations for Particles, Drops, and Bubbles in Agitated Systems

TABLE 5-27 Mass Transfer Correlations for Fixed and Fluidized Beds

TABLE 5-28 Mass Transfer Correlations for Packed Two-Phase Contactors—Absorption, Distillation, Cooling Towers, and Extractors (Packing Is Inert)

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TABLE 5-21 Mass Transfer Correlations for a Single Flat Plate or Disk—Transfer to or from Plate to Fluid

Situation	Correlation	Comments E = Empirical, S = Semiempirical, T = Theoretical
A. Laminar, local, flat plate, forced flow	$N_{Sh,x} = \frac{k'x}{D} = 0.323(N_{Re,x})^{1/2}(N_{Sc})^{1/3}$ Coefficient 0.332 is a better fit.	[T] Low M.T. rates. Low mass-flux, constant property systems. $N_{Sh,x}$ is local k. Use with arithmetic difference in concentration. Coefficient 0.323 is Blasius' approximate solution. $N_{Re,x} = \frac{xu_\infty \rho}{\mu}$, x = length along plate
Laminar, average, flat plate, forced flow	$N_{Sh,avg} = \frac{k'_m L}{D} = 0.646(N_{Re,L})^{1/2}(N_{Sc})^{1/3}$ k'_m is mean mass-transfer coefficient for dilute systems.	$N_{Re,L} = \frac{Lu_\infty \rho}{\mu}$, 0.664 (Polhausen) is a better fit for $N_{Sc} > 0.6$, $N_{Re,x} < 3 \times 10^5$.
j-factors	$j_D = j_H = \frac{f}{2} = 0.664(N_{Re,L})^{-1/2}$	[S] Analogy. $N_{sc} = 1.0$, f = drag coefficient. j_D is defined in terms of k'_m.

Knudsen et al. (1997).

F. Turbulent, local flat plate, forced flow	$N_{Sh,x} = \frac{k'x}{D} = 0.0292 N_{Re,x}^{0.8}$	[S] Low mass-flux with constant property system. Use with arithmetic concentration difference. $N_{Sc} = 1.0$, $N_{Re,x} > 10^5$
Turbulent, average, flat plate, forced flow	$N_{Sh,avg} = \frac{k'L}{D} = 0.0365 N_{Re,L}^{0.8}$, average coefficient	Based on Prandtl's 1/7-power velocity law, $\frac{u}{u_\infty} = \left(\frac{y}{\delta}\right)^{1/7}$
G. Laminar and turbulent, flat plate, forced flow	$j_D = j_H = \frac{f}{2} = 0.037 N_{Re,L}^{-0.2}$	Chilton-Colburn analogies, $N_{Sc} = 1.0$, (gases), f = drag coefficient. Corresponds to item 5-21-F and refers to same conditions. $8000 < N_{Re} < 300{,}000$. Can apply analogy, $j_D = f/2$, to entire plate (including laminar portion) if average values are used.

Knudsen et al. (1997).

K. Turbulent, spinning disk	$N_{Sh} = \frac{k' d_{disk}}{D} = 5.6 N_{Re}^{1.1} N_{Sc}^{1/3}$	[E] Use arithmetic concentration difference. $6 \times 10^5 < N_{Re} < 2 \times 10^6$ $120 < N_{Sc} < 1200$ $u = \omega d_{disk}/2$ where ω = rotational speed, radians/s. $N_{Re} = \rho \omega d^2 / 2\mu$.
L. Mass transfer to a flat plate membrane in a stirred vessel	$N_{Sh} = \frac{k' d_{tank}}{D} = a N_{Re}^{b} N_{Sc}^{c}$ a depends on system. $a = 0.0443$ [73, 165]; b is often 0.65–0.70 [110]. If $N_{Re} = \frac{\omega d_{tank}^2 \rho}{\mu}$ $b = 0.785$ [73]. c is often 0.33 but other values have been reported [110].	[E] Use arithmetic concentration difference. ω = stirrer speed, radians/s. Useful for laboratory dialysis, R.O., U.F., and microfiltration systems.

Knudsen et al. (1997).

TABLE 5-22 Mass Transfer Correlations for Falling Films with a Free Surface in Wetted Wall Columns—Transfer between Gas and Liquid

Situation	Correlation	Comments E = Empirical, S = Semiempirical, T = Theoretical
A. Laminar, vertical wetted wall column	$N_{Sh,avg} = \frac{k'_m x}{D} \approx 3.41 \frac{x}{\delta_{film}}$ (first term of infinite series) $\delta_{film} = \left(\frac{3\mu Q}{w\rho g}\right)^{1/3}$ = film thickness w = film width (circumference in column)	[T] Low rates M.T. Use with log mean concentration difference. Parabolic velocity distribution in films. $N_{Re,film} = \frac{4Q\rho}{w\mu} < 20$ Derived for flat plates, used for tubes if $r_{tube}\left(\frac{\rho g}{2\sigma}\right)^{1/2} > 3.0$. σ = surface tension If $N_{Re,film} > 20$, surface waves and rates increase. An approximate solution $D_{apparent}$ can be used. Ripples are suppressed with a wetting agent good to N_{Re} = 1200.
B. Turbulent, vertical wetted wall column	$N_{Sh,avg} = \frac{k'_m d_t}{D} = 0.023 N_{Re}^{0.83} N_{Sc}^{0.44}$ A coefficient 0.0163 has also been reported using $N_{Re'}$, where v = v of gas relative to liquid film.	[E] Use with log mean concentration difference for correlations in B and C. N_{Re} is for gas. N_{Sc} for vapor in gas. $2000 < N_{Re} \le 35{,}000$, $0.6 \le N_{Sc} \le 2.5$. Use for gases, d_t = tube diameter.

Knudsen et al. (1997).

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TABLE 5-23 Mass-Transfer Correlations for Flow in Pipes and Ducts—Transfer is from Wall to Fluid

Situation	Correlation	Comments E = Empirical, S = Semiempirical, T = Theoretical
A. Tubes, laminar, fully developed parabolic velocity profile, developing concentration profile, constant wall concentration	$N_{Sh} = \frac{k'd_t}{D} = 3.66 + \frac{0.0668(d_t/x)N_{Re}N_{Sc}}{1 + 0.04[(d_t/x)N_{Re}N_{Sc}]^{2/3}}$	[T] Use log mean concentration difference. For $\frac{x/d_t}{N_{Re}N_{Sc}} < 0.10$, $N_{Re} < 2100$. x = distance from tube entrance. Good agreement with experiment at values $10^4 > \frac{\pi}{4}\frac{d_t}{x}N_{Re}N_{Sc} > 10$
B. Tubes, fully developed concentration profile	$N_{Sh} = \frac{k'd_t}{D} = 3.66$	[T] Subset of 5-23-A for fully developed concentration profile. $\frac{x/d_t}{N_{Re}N_{Sc}} > 0.1$
C. Tubes, approximate solution	$N_{Sh,x} = \frac{k'd_t}{D} = 1.077\left(\frac{d_t}{x}\right)^{1/3}(N_{Re}N_{Sc})^{1/3}$ $N_{Sh,avg} = \frac{k'd_t}{D} = 1.615\left(\frac{d_t}{L}\right)^{1/3}(N_{Re}N_{Sc})^{1/3}$	[T] For arithmetic concentration difference. $\frac{W}{\rho D x} > 400$ Leveque's approximation: Concentration BL is thin. Assume velocity profile is linear. High mass velocity. Fits liquid data well.

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Knudsen et al. (1997).

R. Tubes, turbulent	$N_{Sh,avg} = \frac{k'_m d_t}{D} = 0.023 N_{Re}^{0.83} N_{Sc}^{0.44}$	[E] Evaporation of liquids. Use with log mean concentration difference. See item above. Better fit for gases. $2000 < N_{Re} < 35{,}000$ $0.6 < N_{Sc} < 2.5$.
S. Tubes, turbulent	$N_{Sh} = \frac{k' d_t}{D} = 0.0096 N_{Re}^{0.913} N_{Sc}^{0.346}$	[E] $430 < N_{Sc} < 100{,}000$. Dissolution data. Use for high N_{Sc}.

Knudsen et al. (1997).

TABLE 5-24 Mass Transfer Correlations for Flow Past Submerged Objects

Situation	Correlation	Comments E = Empirical, S = Semiempirical, T = Theoretical
C. Single spheres, molecular diffusion, and forced convection, low flow rates	$N_{Sh} = 2.0 + AN_{Re}^{1/2}N_{Sc}^{1/3}$ $A = 0.5$ to 0.62	[E] Use with log mean concentration difference. Average over sphere. Frössling Eq. ($A = 0.552$), $2 \le N_{Re} \le 800$, $0.6 \le N_{sc} \le 2.7$. N_{Sh} lower than experimental at high N_{Re}.
	$A = 0.60.$	[E] Ranz and Marshall $2 \le N_{Re} \le 200$, $0.6 \le N_{sc} \le 2.5$. See also Table 5-27-L
	$A = 0.95.$	[E] Liquids $2 \le N_{Re} \le 2{,}000$. Graph in Ref. 146, p. 217–218.
	$A = 0.95.$	[E] $100 \le N_{Re} \le 700$; $1{,}200 \le N_{Sc} \le 1525$.
	$A = 0.544.$	[E] Use with arithmetic concentration difference. $N_{Sc} = 1$; $50 \le N_{Re} \le 350$.

Knudsen et al. (1997).

H. Single cylinders, perpendicular flow	$N_{Sh} = \frac{k'd_s}{D} = AN_{Re}^{1/2}N_{Sc}^{1/3}, A = 0.82$	[E] $100 < N_{Re} \leq 3500, N_{Sc} = 1560.$
	$A = 0.74$	[E] $120 \leq N_{Re} \leq 6000, N_{Sc} = 2.44.$
	$A = 0.582$	[E] $300 \leq N_{Re} \leq 7600, N_{Sc} = 1200.$
	$j_D = 0.600(N_{Re})^{-0.487}$	[E] Use with arithmetic concentration difference.
	$N_{Sh} = \frac{k'd_{cyl}}{D}$	$50 \leq N_{Re} \leq 50{,}000$; gases, $0.6 \leq N_{Sc} \leq 2.6$; liquids; $1000 \leq N_{Sc} \leq 3000$. Data scatter ± 30%.

Knudsen et al. (1997).

TABLE 5-26 Mass-Transfer Correlations for Particles, Drops, and Bubbles in Agitated Systems

Situation	Correlation	Comments E = Empirical, S = Semiempirical, T = Theoretical
A. Solid particles suspended in agitated vessel containing vertical baffles, continuous phase coefficient	$\frac{k'_{LT} d_p}{D} = 2 + 0.6 N_{Re,T}^{1/2} N_{Sc}^{1/3}$ Replace v_{slip} with v_T = terminal velocity. Calculate Stokes' law terminal velocity $v_{Ts} = \frac{d_p^2 \lvert \rho_p - \rho_c \rvert g}{18 \mu_c}$ and correct: $N_{Re,Ts}$: 1, 10, 100, 1,000, 10,000, 100,000 v_T/v_{Ts}: 0.9, 0.65, 0.37, 0.17, 0.07, 0.023 Approximate: $k'_L = 2k'_{LT}$	[S] Use log mean concentration difference. Modified Frossling equation: $N_{Re,Ts} = \frac{v_{Ts} d_p \rho_c}{\mu_c}$ (Reynolds number based on Stokes' law.) $N_{Re,T} = \frac{v_T d_p \rho_c}{\mu_c}$ (terminal velocity Reynolds number.) k'_L almost independent of d_p. Harriott suggests different correction procedures. Range k'_L/k'_{LT} is 1.5 to 8.0.

$N_{Re,Ts}$	1	10	100	1,000	10,000	100,000
v_T/v_{Ts}	0.9	0.65	0.37	0.17	0.07	0.023

Knudsen et al. (1997).

VI. Exercícios

Ex. 1. Uma forma com água é colocada num túnel de vento, com ar fluindo a 7 m/s no sentido comprido da forma (4 m, considere que ela é muito larga).

A água está a 292 K e a pressão total é de 1 atm.

Nessas condições, a pressão de vapor da água é 2000 Pa, a difusividade da água no ar é $2{,}5 \times 10^{-5}$ m^2/s e a viscosidade cinemática do ar é $1{,}5 \times 10^{-5}$ m^2/s.

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a) Calcule a taxa mássica de evaporação da água.

R= 82,057.10^{-3} m^3.atm/mol.K = 8,314462 $m^3 Pa K^{-1} mol^{-1}$.

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R= 82,057.10^{-3} m^3.atm/mol.K = 8,314462 $m^3 Pa K^{-1} mol^{-1}$.

Então, temos que calcular Re:

$$Re = \frac{7\frac{m}{s}\, x\ 4\ m}{1{,}5x10^{-5}\ m^2/s} = 1{,}9x10^6$$

Então, podemos usar:

$$j_M = 0{,}0365/Re^{0,2} \quad p/\ Re > 3\ x10^5$$

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$$j_M = 0{,}0365/Re^{0,2} \quad p/\ Re > 3\ x10^5$$

Logo

$$j_M = 0{,}0365/(1{,}9x10^6)^{0,2} = 2{,}03x10^{-3}$$

Então, podemos usar:

$$j_M = 0{,}0365/Re^{0,2} \quad p/\ Re > 3\ x10^5$$

Logo

$$j_M = 0{,}0365/(1{,}9x10^6)^{0,2} = 2{,}03x10^{-3}$$

Sabendo que

$$St_M\ Sc^{2/3} = j_M$$

$$St_M = \frac{k_m}{V_\infty} = j_M\ Sc^{-2/3}$$

Portanto,

$$k_m = V_\infty j_M \, Sc^{-2/3}$$

$$k_m = 7\frac{m}{s} \quad 2{,}03\text{x}10^{-3} \left(\frac{1{,}5 \text{ x } 10^{-5} \text{ m}^2/\text{s}}{2{,}5 \text{ x}10^{-5} \text{ m}^2/\text{s}}\right)^{-2/3}$$

k_m = 0,02 m/s

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Como não temos a área da forma, vamos calcular o fluxo mássico ($\rho_{A\infty} = 0$)

$$n_A = k_m\,(\rho_{As})$$

Mas, antes precisamos calcular a concentração de água na interface. Usando a Lei dos gases ideais:

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Mas, antes precisamos calcular a concentração de água na interface. Usando a Lei dos gases ideais:

$$\rho_{AS} = \frac{M_A\, p_{As}}{RT}$$

Logo

$$\rho_{AS} = \frac{18x10^{-3}\,\frac{kg}{mol}\quad 2000\ Pa}{(8{,}314462\ \mathrm{m^3PaK^{-1}mol^{-1}})(292\ K)}$$

E,

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Logo

$$\rho_{AS} = \frac{18x10^{-3}\,\frac{kg}{mol}\;\;2000\;Pa}{(8{,}314462\ \mathrm{m^3 Pa K^{-1} mol^{-1}})(292\ K)}$$

E,

$$\rho_{AS} = 0{,}015\ \mathrm{kg/m^3}$$

Temos, então:

$$n_A = 0{,}02 \text{ m/s } (0{,}015 \text{ kg/m}^3)$$

$$n_A = 3 \text{ x } 10^{-4} \text{ kg/m}^2 s$$

b) Vamos considerar que a lâmina de água tem 1 cm de altura, e a bandeja tem 10 m de largura. Vamos calcular o tempo necessário para esvaziar a forma, ou seja, para que toda a água seja volatilizada.

Vou dar 3 minutos para vocês tentarem começar...

b) Vamos considerar que a lâmina de água tem 1 cm de altura, e a bandeja tem 10 m de largura. Vamos calcular o tempo necessário para esvaziar a forma, ou seja, para que toda a água seja volatilizada.

Vou dar 3 minutos para vocês tentarem começar...

Sabemos que a redução da lâmina d'água pode ser calculada:

$$W_A = -\rho \frac{dV}{dt}$$

E, que, considerando que o ar de entrada está seco.

$$W_A = A_s \, k_m (\rho_{As})$$

Logo

$$-\rho\frac{dV}{dt} = A_s\,k_m(\rho_{As})$$

Logo

$$-\rho \frac{dV}{dt} = A_s \, k_m (\rho_{As})$$

Sabendo que

$$\frac{dV}{dt} = X \, Y \frac{dz}{dt}$$

Logo

$$-\rho \frac{dV}{dt} = A_s \, k_m(\rho_{As})$$

Sabendo que

$$\frac{dV}{dt} = X \, Y \frac{dz}{dt}$$

Teremos

$$-\rho \, \cancel{X} \, Y \frac{dz}{dt} = \cancel{A_s} \, k_m(\rho_{As})$$

Então, podemos fazer:

$$-\int_{1}^{0} \rho dz = \int_{0}^{t} k_m(\rho_{As})dt$$

E,

$$-(0 - 1cm)\,\rho = k_m\,(\rho_{As})t$$

Então, podemos fazer:

$$-\int_{1}^{0} \rho dz = \int_{0}^{t} k_m(\rho_{As})dt$$

E,

$$-(0 \ - 1cm) \ \rho = k_m \ (\rho_{As})t$$

Então, teremos

$$t = (1cm) \ \rho / 0{,}02 \text{ m/s} (0{,}015 \text{ kg/m}^3)$$

A.3-3 Physical Properties of Air at 101.325 kPa (1 Atm Abs), SI Units

T (°C)	T (K)	ρ (kg/m^3)	c_p ($kJ/kg \cdot K$)	$\mu \times 10^5$ ($Pa \cdot s$, or $kg/m \cdot s$)	k ($W/m \cdot K$)	N_{Pr}	$\beta \times 10^3$ (1/K)	$g\beta\rho^2/\mu^2$ ($1/K \cdot m^3$)
−17.8	255.4	1.379	1.0048	1.62	0.02250	0.720	3.92	2.79×10^8
0	273.2	1.293	1.0048	1.72	0.02423	0.715	3.65	2.04×10^8
10.0	283.2	1.246	1.0048	1.78	0.02492	0.713	3.53	1.72×10^8
37.8	311.0	1.137	1.0048	1.90	0.02700	0.705	3.22	1.12×10^8
65.6	338.8	1.043	1.0090	2.03	0.02925	0.702	2.95	0.775×10^8
93.3	366.5	0.964	1.0090	2.15	0.03115	0.694	2.74	0.534×10^8
121.1	394.3	0.895	1.0132	2.27	0.03323	0.692	2.54	0.386×10^8
148.9	422.1	0.838	1.0174	2.37	0.03531	0.689	2.38	0.289×10^8
176.7	449.9	0.785	1.0216	2.50	0.03721	0.687	2.21	0.214×10^8
204.4	477.6	0.740	1.0258	2.60	0.03894	0.686	2.09	0.168×10^8
232.2	505.4	0.700	1.0300	2.71	0.04084	0.684	1.98	0.130×10^8
260.0	533.2	0.662	1.0341	2.80	0.04258	0.680	1.87	0.104×10^8

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−17.8	255.4	1.379	1.0048	1.62	0.02250	0.720	3.92	2.79×10^8
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93.3	366.5	0.964	1.0090	2.15	0.03115	0.694	2.74	0.534×10^8
121.1	394.3	0.895	1.0132	2.27	0.03323	0.692	2.54	0.386×10^8
148.9	422.1	0.838	1.0174	2.37	0.03531	0.689	2.38	0.289×10^8
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232.2	505.4	0.700	1.0300	2.71	0.04084	0.684	1.98	0.130×10^8
260.0	533.2	0.662	1.0341	2.80	0.04258	0.680	1.87	0.104×10^8

Interpolando: 37,8 -10 = 27,8 --- -0,109 = 1,137 – 1,246

37,8 -18,9 = 18,9 --- 1,137 – x

→ -0,074 = 1,137 – x

→ x = 1,211 kg/m^3

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Substituindo os respectivos valores, teremos:

$$t = (0{,}01m)1{,}211 \text{ kg/m}^3/0{,}02 \text{ m/s}(0{,}015\ kg/m^3)$$

Substituindo os respectivos valores, teremos:

$$t = (0{,}01\,m)1{,}211\ kg/m^3/0{,}02\ m/s(0{,}015\ kg/m^3)$$

$$\boxed{t = 40{,}4\ s}$$

Ex. 2. Uma bolha de gás de cloro puro de 0,5 cm de diâmetro está ascendendo a uma velocidade de 20 cm/s em água pura, a 1,0 atm e 16 °C.

Nessas condições, a difusividade do cloro na água é de $1,26 \times 10^{-9}$ m^2/s; a solubilidade do cloro em água é de 0,823 g de cloro/100 g de água; a densidade da água é de 1000 kg/m^3 e sua viscosidade é de $1,155 \times 10^{-3}$ kg/m.s.

O cloro, a partir da superfície da bolha, dissolve-se na água e difunde-se para longe, diminuindo o tamanho da bolha.

Qual a taxa mássica de absorção de cloro pela água quando a bolha for de 0,5 cm de diâmetro?

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Vamos calcular Re:

$$Re = \frac{1000\ kg/m^3\ 0{,}20\frac{m}{s}\ x\ 0{,}005\ m}{1{,}155x10^{-3}\ kg/m.s} = 8{,}66x10^2$$

Vamos calcular Re:

$$Re = \frac{1000\ kg/m^3\ 0{,}20\frac{m}{s}\ x\ 0{,}005\ m}{1{,}155x10^{-3}\ kg/m.s} = 8{,}66x10^2$$

Como 200 ≤ Re ≤ 4x10^4:

$$j_M = \frac{0{,}43}{Re^{0{,}44}} = \frac{0{,}43}{(8{,}66x10^2)^{0{,}44}}$$

$$j_M = 0{,}022$$

E, já vimos que

$$k_m = V_\infty j_M\ Sc^{-2/3}$$

Precisamos calcular Sc:

$$Sc = \frac{\upsilon}{D_{AB}} = \frac{\mu}{\rho \quad x D_{AB}}$$

Ou

$$Sc = \frac{1{,}155x10^{-3}\ \text{kg/m.s}}{1000\ \text{kg/m}^3\ x\ 1{,}26x10^{-9}\ m^2/s} = 916{,}7$$

Precisamos calcular Sc:

$$Sc = \frac{\upsilon}{D_{AB}} = \frac{\mu}{\rho \ \ x D_{AB}}$$

Ou

$$Sc = \frac{1{,}155x10^{-3} \text{ kg/m.s}}{1000 \text{ kg/m}^3 \, x \; 1{,}26x10^{-9} \; m^2/s} = 916{,}7$$

Agora, substituindo todos os valores, em

$$k_m = V_\infty j_M \, Sc^{-2/3}$$

teremos:

$$k_m = 20 \frac{cm}{s} \; 0{,}022 \; (916{,}7)^{-2/3} = 4{,}66x10^{-3} \text{ cm/s}$$

Portanto, considerando $\rho_{A\infty} = 0$:

$$n_A = k_m\,(\rho_{As})$$

Ou

$$n_A = 4{,}66\text{x}10^{-3}\ \text{cm/s}\ (0{,}00823\ \frac{g\ cloro}{g\ água} x\ 1{,}000\ \frac{g\ \ água}{cm^3})$$

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Portanto, considerando $\rho_{A\infty} = 0$:

$$n_A = k_m\,(\rho_{As})$$

Ou

$$n_A = 4{,}66\text{x}10^{-3}\ \text{cm/s}\ (0{,}00823\ \frac{g\ cloro}{\cancel{g\ água}}\ x\ 1{,}000\ \frac{\cancel{g\ água}}{cm^3})$$

Ou seja

$$n_A = 3{,}84\text{x}10^{-5}\ \text{g cloro/cm}^2\text{.s}$$

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Logo, como:

$$W_A = A_s\, n_A$$

$$W_A = 4\pi R^2 n_A = 4\pi(0{,}5\ cm)^2\ 3{,}84\text{x}10^{-5}\ \text{g cloro/cm}^2\text{.s}$$

Logo, como:

$$W_A = A_s\, n_A$$

$$W_A = 4\pi R^2 n_A = 4\pi(0{,}5\ \text{cm})^2\ 3{,}84\text{x}10^{-5}\ \text{g cloro/cm}^2\text{.s}$$

$$\boxed{W_A = 1{,}21\text{x}10^{-4}\ \text{g cloro/s}}$$

Último exercício: Vamos secar um tubo de diâmetro interno = 0,015 m cuja superfície interna está coberta com uma camada de água, com ar a 300 K e 1 atm escoando a 1,2 m/s.

Calcule o fluxo de água nesse processo.

Por causa das condições de baixo fluxo de massa, podemos usar as propriedades do ar seco:

$\nu = 1,58 \times 10^{-5}\ m^2/s$

$D_{AB} = 2,54 \times 10^{-5}\ m^2/s$

Até a próxima semana