

EXTENDED CHANNEL MODEL FOR PREDICTION OF HYDRAULICS AND MASS TRANSFER OF RANDOM PACKED COLUMNS FOR GAS-LIQUID SYSTEMS

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Abstract

The following work presents a new, generally applicable model for description the mass transfer in the liquid phase for packed columns filled with random non-perforated and lattice-type packing with size between 12 and 90 mm for gas-liquid systems in operating range up to flooding point.

The new equation was derived on the basis of the assumption that liquid flows down in packed bed, mainly in the form of droplets, and that mass transfer is transient and based on the channel model with a partly open structure.

The experimentally derived values for the mass transfer area in different types of random packings a_e are sufficiently consistent with the calculation based on the new model. It is therefore possible to separate the product $\beta_L \cdot a_e$ into liquid phase mass transfer coefficient β_L and effective interfacial area a_e .

Keywords: random packing, lattice-type packing, separation performance, effective mass transfer area, extended channel model, liquid phase mass transfer coefficient

1. Introduction

In the field of separation technology, the use of random, lattice-type packings in addition to structured packings has been gradually increasing. The new generation of lattice packings, so called Nor-Pac were first presented by Billet und Maćkowiak¹ (1979). Contrary to expectations, initial experimental results¹ showed that the mass transfer behaviour of the 25 mm Nor-Pac was similar to that of 25 mm Pall rings made of plastic, which were analysed for comparison. The loading capacity of the new lattice packings was significantly higher than that of Pall rings, whilst the pressure drop $\Delta p/H$ and the specific pressure drop $\Delta p/NTU_{OV}$ of the lattice packings was considerably lower, yet they were found to have the same separation efficiency when applied under the same operating conditions. As a result, a number of new lattice-type packing elements of different types were produced by leading German packing manufacturers, see Fig. 1. These lattice packings were initially made of plastic, followed later by models made of ceramic and metal¹⁻⁴.

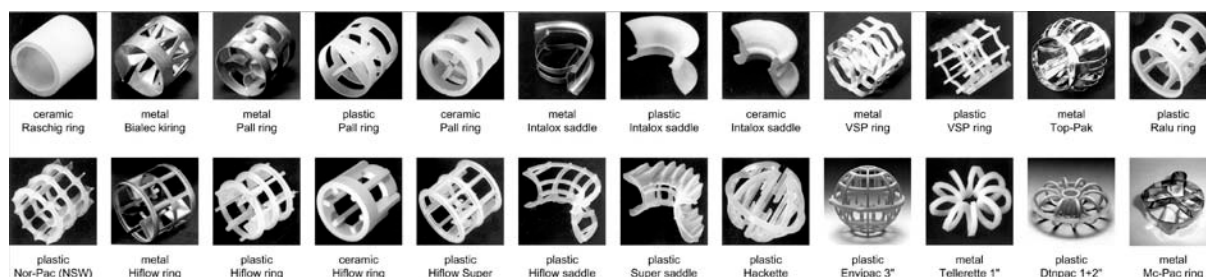


Figure 1. Overview of packing elements investigated

Following on from the correlations for liquid-liquid systems derived by Maćkowiak, Billet (1982/84)⁵, which are based on the model of non-stationary diffusion for short contact times, it was in the 1990s that dimensionless methods, developed by, were applied to gas-liquid systems.

The separation of the product $\beta_L \cdot a_e$ was first achieved by Zech, Mersmann⁶ and Billet and Schultes⁷ as well as Bornhütter and Mersmann⁸. Based on the assumption of rivulet formation, they derived new

correlations for determining the effective interfacial area per unit volume a_e for non-perforated ceramic packing elements such as spheres, Raschig rings and saddles⁶ and lattice-type packing⁸. A comprehensive overview of the methods used to describe the resistance of the mass transfer in the liquid phase is available in literature.

The aim of this study is to develop a generally applicable method for determining the volumetric mass transfer coefficient in the liquid phase $\beta_L \cdot a_e$ for gas-liquid systems, valid for different types of classic, non-perforated as well as for lattice-type packing elements and that can be used to predict the separation efficiency for any type of packing based on specific packing-related data.

2. Deriving a model for determining the volumetric mass transfer coefficient in the liquid phase $\beta_L \cdot a$ below the loading line

Visual observations and measurements of droplet proportions have shown that in random packings liquid primarily occurs in the form of droplets rather than rivulets². As the size of the packing element increases, the amount of the droplets also increases, an observation that was confirmed by studies carried out by Bornhütter and Mersmann⁸ in connection with lattice packings. For this reason, it can be expected that mass transfer occurs non-stationary and can be described by models that are valid for disperse systems. Droplets generated in the random packing fall in the gas phase, which constitutes the continuous phase.

The new method is derived on the basis of a model, whereby the liquid in a random packing flows down along the surface of the individual packing elements in the form of thin rivulets, whereas between the individual packing elements the liquid flows down mainly in the form of droplets. As a result, it is possible to determine the interfacial area per unit volume using the correlation of eqn. (1)

$$a_e = 6 \cdot \frac{h_L}{d_T} \quad [m^2/m^3] \quad (1)$$

which is valid for disperse systems⁵. The liquid flowing down the edges of the packing in the form of droplets has a composition that is not in equilibrium with the surrounding gas phase. This disequilibrium results in a mass transfer, which is highest at the beginning and decreases along the flow length l , which is referred to here as the contact path. During the formation of rivulets, mass transfer is interrupted and only recommences as new droplets are formed. The process is therefore non-stationary, as described by the well-known model of Higbie [10]

$$\beta_L = \frac{2}{\sqrt{\pi}} \sqrt{\frac{D_L}{\tau}} \quad [m/s] \quad (2)$$

As a result, the mass transfer coefficient β_L for mass transfer in the liquid phase can be determined acc. to eqn. (2) if the contact time τ is known. The contact time τ in eqn. (2) is described by the time that a droplet needs to cover the distance l between two contact points within the packing. Hence:

$$\tau = \frac{l}{\bar{u}_L} \Rightarrow \frac{l \cdot h_L}{u_L} \quad \text{for} \quad \bar{u}_L = \frac{u_L}{h_L} \quad [s] \quad (3)$$

Acc. to [2], the liquid hold-up h_L in random packings for turbulent liquid flow $Re_L \geq 2$ in the range below loading point can be described by eqn. (4):

$$h_L = C_p \cdot Fr_L^{1/3} = C_p \cdot \left(\frac{a \cdot u_L^2}{g} \right)^{1/3} \quad \text{for} \quad F_V \leq 0.65 \cdot F_{V,FI} \quad \text{and} \quad C_p = 0.57 \quad [m^3/m^3] \quad (4)$$

Based on the evaluation of more than 1000 experimental data points for the liquid hold-up using systems with different physical properties², the constant C_p in eqn. (4) was found to have a mean value of $C_p = 0.57$. The experimental values² are reproduced by eqn. (4) for the operating range below the loading line with a relative error of ± 20 –25% for different types of plastic packings with nominal sizes of $d = 0.015$ – 0.090 m.

The product of the mass transfer coefficient β_L and the interfacial area per unit volume a_e results from eqns. (1) and (2). Substituting eqns. (3) and (4) in eqns. (1) and (2) leads to the following equation (5):

$$\beta_L \cdot a_e = 12 \cdot \left(\frac{C_p}{\pi \cdot l} \right)^{1/2} \cdot \left(\frac{a}{g} \right)^{1/6} \cdot \left(\frac{D_L \cdot \Delta \rho \cdot g}{\sigma_L} \right)^{1/2} \cdot u_L^{5/6} \quad [1/s] \quad (5)$$

3. Experimental results

Eqn. (5) and the experimental results, shown as example in Fig. 2, reveal two parameters that have a main effect on mass transfer coefficient $\beta_L \cdot a_e$ in the liquid phase for given systems with known physical parameters: the geometric surface area of the packing per unit volume a and the contact path l as well as specific liquid load u_L .

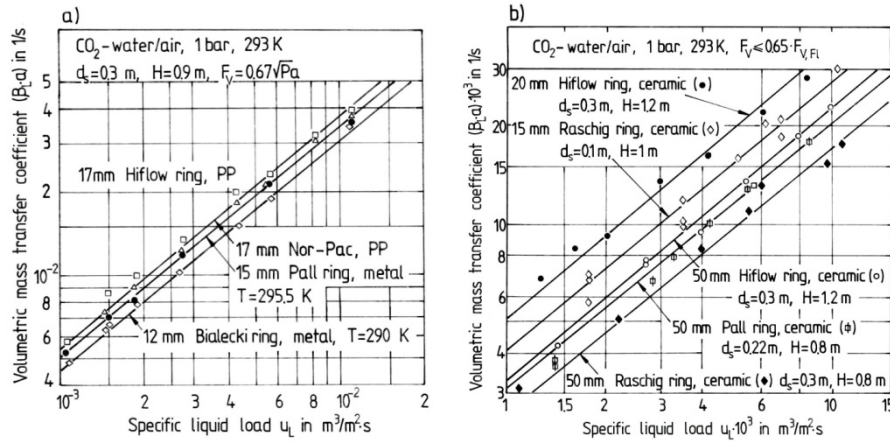


Figure 2. Volumetric mass transfer coefficient $\beta_L \cdot a_e$ as a function of the specific liquid load u_L , valid for: **a)** randomly filled 15 mm Pall rings, 12 mm Bialecki rings and 17 mm Hiflow rings as well as 17 mm Nor-Pac made of plastic. System: CO₂-water/air, 293 K, $d_s = 0.3$ m, $H = 0.9$ m; **b)** different packings with nominal dimensions of 15-50 mm made of ceramic. System: CO₂-water/air, 1 bar, 295 K

The effect of the individual packing elements on the fluid dynamics of random packings described by a new extended “**channel model with open structure**” was discussed in a previous study^{2,9}. Acc. to this model, a random packing is characterised not only by the geometric surface area of the packing a and the void fraction ε , but also by third parameter, which is: the form factor φ_P , defined as the ratio of the open area to the total surface area of the packing element^{2,9}. In the case of classic packing elements with non-perforated walls, as Raschig rings and saddles, form factor is given as $\varphi_P = 0$ acc. To^{2,9}. Fig. 3a shows that for the same value of hydraulic diameter d_h their contact paths l are almost twice as long as those of lattice packings with a very open structure acc. to Fig. 3d with form factors of $\varphi_P = 0.55-0.7^2$.

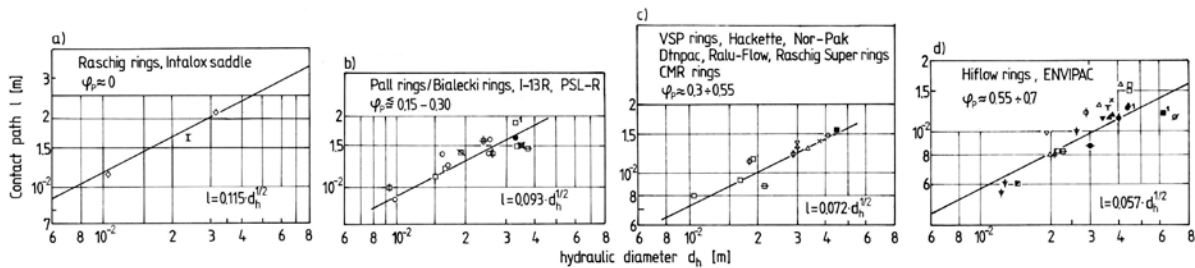


Figure 3. Effect of hydraulic diameter d_h on mean contact path l for types of packings investigated: **a)** for classic, non-perforated packing elements for $\varphi_P = 0$; **b)** for classic, perforated packing elements $\varphi_P = 0.15-0.30$; **c)** for lattice packings with perforated walls for $\varphi_P = 0.30-0.55$; **d)** for lattice packings with highly perforated walls for $\varphi_P \geq 0.55-0.70$

The numerical values for the contact paths l in eqn. (5) for the investigated packings were determined on the basis of the experimental data of this work shown for example in Figs. 2a/b. Plotting the contact path l on the hydraulic diameter d_h of the packing gives the following correlation for the packings investigated acc. to Figs. 3a-d:

$$l = 0.115 \cdot (1 - \varphi_P)^{2/3} \cdot d_h^{1/2} \quad [m] \quad (6)$$

where φ_P is a parameter relating to a different characteristic form of packing element^{2,9}. Figs. 3a-d show that not only the size and type of the packing element has a significant effect on the contact path. It can be noted that the more open the structure of the packing element, the shorter the contact paths l .

Substituting the relations of eqn. (6) in eqn. (5) leads to the new, generally valid eqn. (7) for the prediction of volumetric mass transfer coefficient $\beta_L \cdot a_e$ in columns with random packings:

$$\beta_L \cdot a_e = \frac{15.1}{(1 - \varphi_P)^{1/3} \cdot d_h^{1/4}} \cdot \left(\frac{D_L \cdot \Delta \rho \cdot g}{\sigma_L} \right)^{1/2} \cdot \left(\frac{a}{g} \right)^{1/6} \cdot u_L^{5/6} \quad [1/s] \quad (7)$$

This equation is valid below the loading line $F_V \leq 0.65 F_{V,Fl}$ and for turbulent liquid flow $Re_L \geq 2$, s. Table 1. Eqn. (7) allows to consolidate the information on mass transfer in the liquid phase in random packings containing packing elements of different types and sizes, enabling us to predict the $\beta_L \cdot a_e$ values for different types of modern and classic packings sufficiently enough for practical applications with a mean error of $\pm 13\%$ in the range below loading line for experimental data presented in^{1,4,7,8}, new data of author, see Table 1, and other literature data.

Table 1. Experimental conditions and limits of validity of models as per eqn. (7) and eqn. (9) for $Re_L \geq 2$ for investigated classic and lattice packing shown in Fig. 1

d	=	0.012 – 0.090	m	Re_L	=	2 – 275
a	=	54.2 – 403.0	m ² /m ³	$F_V/F_{V,Fl}$	≤	1
ε	=	0.696 – 0.987	m ³ /m ³	Fr_L	=	$5.5 \cdot 10^{-6} - 1.4 \cdot 10^{-2}$
φ_P	=	0 – 0.70		We/Fr_L	=	0.8 – 4.5
d_S	=	0.10 – 1.2	m	Sc_L	=	5100 – 10000
H	=	0.71 – 2.1	m			

4. The volumetric mass transfer coefficient in the liquid phase $(\beta_L \cdot a_e)_S$ above the loading line and below the flooding point

The quotient $(\beta_L \cdot a_e)_S/(\beta_L \cdot a_e)$ can be expressed according to experimental data of this work as a function of the relative column load $F_V/F_{V,Fl}$ and can be described using the following empirical correlation (8) for $u_L = \text{const.}$:

$$(\beta_L \cdot a_e)_S = (\beta_L \cdot a_e) \cdot \left[1 + \left(\frac{F_V}{F_{V,Fl}} - 0.65 \right) \right]_{u_L = \text{const}} = (\beta_L \cdot a_e) \cdot \left(0.35 + \frac{F_V}{F_{V,Fl}} \right)_{u_L = \text{const}} \quad [1/s] \quad (8)$$

By substituting eqn. (7) in eqn. (8), we obtain eqn. (9):

$$(\beta_L \cdot a_e)_S = \frac{15.1}{(1 - \varphi_P)^{1/3} \cdot d_h^{1/4}} \cdot \left(\frac{D_L \cdot \Delta \rho \cdot g}{\sigma_L} \right)^{1/2} \cdot \left(\frac{a}{g} \right)^{1/6} \cdot \left(0.35 + \frac{F_V}{F_{V,Fl}} \right)_{u_L = \text{const}} \cdot u_L^{5/6} \quad [1/s] \quad (9)$$

The evaluation of approx. 40 experimental points in the range below the flooding line reveals a congruence between the calculation based on eqn. (9) and the experiment, with a relative error $\delta(\beta_L \cdot a_e)_S$ of less than $\pm 15\%$ in the range acc. to Table 1.

5. Conclusions

5.1. On the basis of presented model, according to the assumption that droplet flow occurs in random packings containing packing elements with perforated walls, combined with the application of the model of non-stationary diffusion for short contact times [10] for mass transfer in the liquid phase in irrigated packings for gas-liquid systems, it is possible to calculate the volumetric mass transfer coefficient $\beta_L \cdot a_e$ according to eqn. (9) in the range below the flooding point derived from experiments.

To predict the effective mass transfer area a_e and mass transfer coefficient β_L , eqn. (1) and (2) are derived. It is the area formed by droplets that determines the interfacial area per unit volume a_e in the random packing for turbulent liquid flow in the range of $Re_L = 2-250$ acc. to eqn. (1). The comparison between the researched effective mass transfer area a_e and the calculation acc. to eqn. (1) is presented as example in Fig. 4.

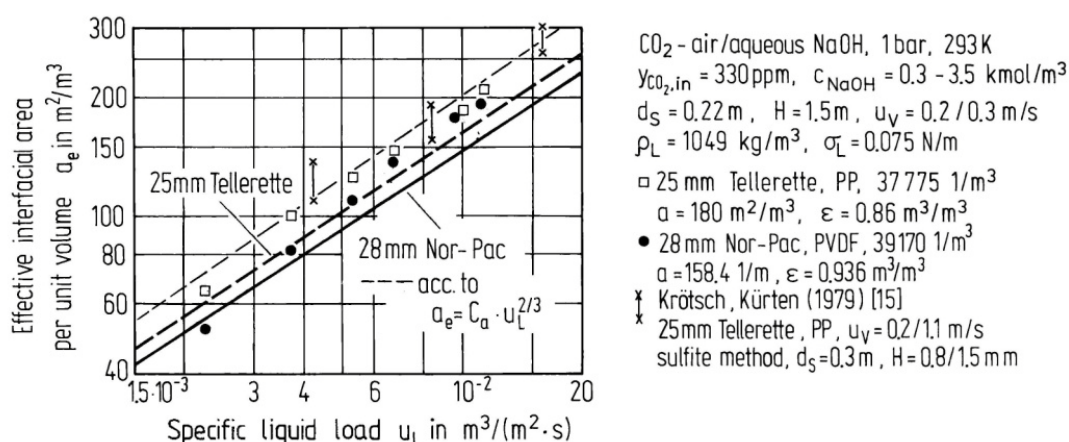


Figure 4. Effective interfacial area per unit volume a_e as a function of the specific liquid load u_L , valid for different packing elements made of plastic, 28 mm Nor-Pac and 25 mm Tellerette, derived from experiments using the system CO_2 -air/aqueous NaOH solution based on pseudo first order reaction, $F_V < 0.65 \cdot F_{V,F}$

5.2. The form factor ϕ_P influences not only the fluid dynamic behaviour^{2,9}, but also acc. to eqn. (7) and (9) the mass transfer of packed column. This additional parameter, characterising the packed bed, can be estimated for simple packing elements without experiments^{2,9} or for more complicated packing form from experiments of dry pressure drop^{2,9}. In this case the flooding gas velocity $u_{V,F}$, pressure drop of packed column $\Delta p/H$ up to flooding point can be calculated using the extended channel model^{2,9}, which shows a directly relation between the fluid dynamic behaviour and mass transfer of packed column.

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List of symbols

a	$[\text{m}^2/\text{m}^3]$	geometric surface area of packing per unit volume
a_e	$[\text{m}^2/\text{m}^3]$	interfacial area per unit volume
C_p	$[-]$	constant, eqn. (4)
d	$[\text{m}]$	packing diameter
d_h	$[\text{m}]$	hydraulic diameter; $d_h = 4 \cdot \varepsilon / a$
d_S	$[\text{m}]$	column diameter
d_T	$[\text{m}]$	mean droplet diameter acc. to Sauter, $d_T = \sqrt{\sigma_L / \Delta\rho \cdot g}$
D_L	$[\text{m}^2/\text{s}]$	diffusion coefficient in the liquid phase
F_V	$[\text{Pa}^{0.5}]$	gas load factor in relation to full column cross section, $F_V = u_V \cdot \sqrt{\rho_V}$
g	$[\text{m}/\text{s}^2]$	acceleration of gravity
h_L	$[\text{m}^2/\text{m}^3]$	liquid hold-up in relation to total packing volume V_S , $h_L = V_L / V_S$
H	$[\text{m}]$	packing height
l	$[\text{m}]$	mean contact path
$\Delta p/H$	$[\text{Pa}/\text{m}]$	pressure drop per 1 m packing height
$\Delta p/NTU_{OV}$	$[\text{Pa}]$	specific pressure drop in relation to number of transfer unit related to gas phase
u_V	$[\text{m}/\text{s}]$	linear gas velocity in relation to full column cross section
u_L	$[\text{m}/\text{s}]$	specific liquid load in relation to full column cross section
\bar{u}_L	$[\text{m}/\text{s}]$	effective liquid load in relation to full column cross section $\bar{u}_L = u_L / h_L$
V_L	$[\text{m}^3]$	liquid volume
V_S	$[\text{m}^3]$	packing volume, $V_S = H \cdot \pi \cdot d_S^2 / 4$

Greek symbols

β	$[\text{m}/\text{s}]$	mass transfer coefficient
ϕ_P	$[-]$	form factor
τ	$[\text{s}]$	contact time
$\rho, \Delta\rho$	$[\text{kg}/\text{m}^3]$	density, density difference $\Delta\rho = \rho_L - \rho_V$
σ_L	$[\text{N}/\text{m}]$	surface tension
ν	$[\text{m}^2/\text{s}]$	kinematic viscosity

Indices

L	relating to liquid
Fl	relating to operating point at flooding point
S	relating to operating point at loading point; $F_V = 0.65 \cdot F_{V,Fl}$
V	relating to gas

Dimensionless numbers

$Fr_L = \frac{u_L^2 \cdot a}{g}$	Froude number	$\frac{We}{Fr_L} = \frac{\rho_L \cdot g}{a^2 \cdot \sigma_L}$	Weber/Froude number
$Re_L = \frac{u_L}{a \cdot \nu_L}$	Reynolds number	$Sc_L = \frac{\nu_L}{D_L}$	Schmidt number

Abbreviations:

R	= ring
CMR	= Cascade Mini Rings (by Koch-Glitsch)
PP	= polypropylene