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Liquid-phase mass-transfer coefficients of Mellapak structured packings under desorption of oxygen from primary alcohols

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Highlights

- Liquid-side mass-transfer coefficient for organic systems on structured packings
- Common correlation for alcohols, water and several sizes of packing
- Dependence on the flow rate found similar for organic and aqueous systems
- Indirect dependence of k_L on the square root of the surface tension

Abstract

Volumetric mass-transfer coefficient in the liquid phase, $k_L a$, for Mellapak 250.Y, 350.Y and 500.Y structured packings has been measured by standard method of desorption of oxygen into nitrogen from pure methanol, ethanol and n-propanol in the column of i.d. of 150 mm. Although utilization of organic solvents for absorption experiments provides valuable behavior interlink between properties of aqueous absorption systems and ordinary (organic) systems under distillation conditions, studies of such kind are not published in the open literature.

The behavior of the k_La data on alcohol systems do not differ significantly from that found for aqueous system, the k_La values rise with the liquid superficial powered to $\sim 0.5 - 0.78$. For one solvent the k_La are almost identical for all three packings under the same superficial velocity. According to the expectations the k_La values under the same superficial liquid flow rate are highest for the solvent with the lowest viscosity and the highest diffusivity of oxygen, methanol, and decrease in the order ethanol, n-propanol and water.

The k_L data for all three packings, alcohols and also for water has been successfully correlated by dimensionless correlation $Sh = 0.207 \cdot Re^{0.798} Fr^{-0.563} We^{0.511} Sc^{0.528}$ with mean relative deviation of 12 %.

Keywords: absorption, mass transfer, mass-transfer coefficient, structured packing

1 Introduction

As an important step for more reliable design of packed absorption and distillation columns we see adoption of the rate-based approach for their modeling. Utilization of the rate-based modeling demands reliable and accurate knowledge of its parameters, volumetric mass-transfer coefficients. These coefficients can be predicted from various models which also describe their dependence on the process conditions. To make sure that these predictions are reliable and the models are generally true, they should be based or verified against experimental data obtained by standard procedures under as broad as possible conditions, namely also under distillation ones.

Evaluation of the volumetric mass-transfer coefficients, k_La , k_Ga , under distillation conditions by the novel “profile method” [1], [2], [3] provided very reasonable results in terms of gas-phase ones, but results in dependence of k_La values on liquid flow rate of $k_La \sim u_L^1$, which is steeper than that found during absorption experiments, where $k_La \sim u_L^{0.62-0.9}$. The significant difference in the properties of the distillation mixtures and mostly aqueous systems used for the absorption experiments was suspected as a possible cause of the difference.

This paper refers on k_La data obtained for Mellapak 250.Y, 350.Y and 500.Y structured packings by desorption of oxygen into nitrogen from oxygen saturated methanol, ethanol and n-propanol at 20°C. The alcohols have been chosen as the constituents of the distillation mixtures used for the profile method experiments. They also differ significantly in viscosity and oxygen diffusivity and therefore cover broad spectra of the physical properties. Although only Mellapak 250.Y has been used for the mirrored distillation experiments, usage of the other packings broadens the experimental portfolio by geometrically similar packings of different size.

2 Literature review

Volumetric mass-transfer coefficient in the liquid phase, $k_L a$, on structured packings has been measured under absorption / desorption conditions with water as a liquid phase by stripping of low soluble gases [4], [5], [6], [7] or toluene [8] into the inert gas. The results are shown in the Fig. 1.

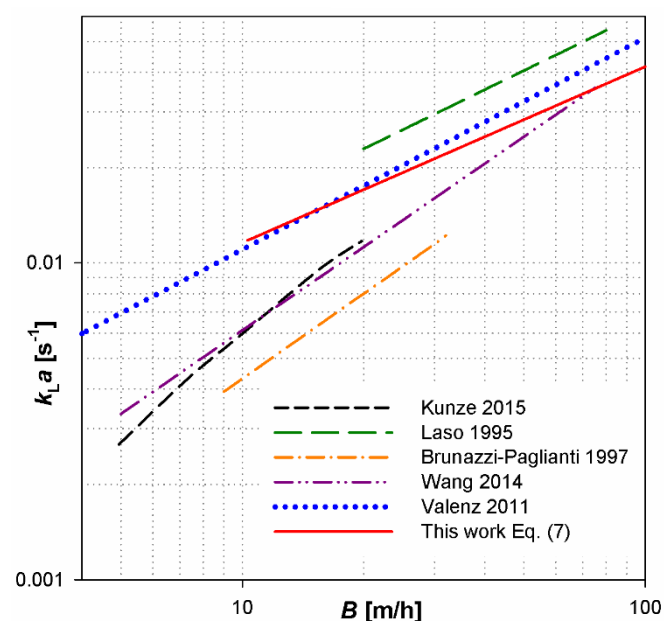


Fig. 1 – Liquid phase volumetric mass-transfer coefficient, $k_L a$. Experimental data for Mellapak 250.Y. Measured with, or recalculated to the O_2 - H_2O system.

For the mutual recalculation of the results obtained with different absorbed / desorbed solutes i, j equation based on the penetration theory (1) was used

$$\frac{k_L a(i)}{k_L a(j)} \propto \sqrt{\frac{D_i}{D_j}} \quad (1)$$

The $k_L a$ values reported by the investigators are given either only in the graphical form [6] or correlated [4], [5] as a dependence of $k_L a$ on the liquid velocity u_L in the power law form (2) with α in the range of 0.62 - 0.9 (see

Table 1).

$$k_L a \propto u_L^{C_0}$$

(2)

Table 1: $k_L a$ values and their trend with liquid velocity for metal Mellapak 250.Y and oxygen stripping from water

		C_0 in the equation (2)	$k_L a$ [1/s] for $B = 20$ m/h
Laso (1995)	oxygen stripping	0.62	0.0229
Brunazzi (1997)	CO ₂ stripping	0.90	0.0080 (bed height = 0.84m)
Valenz (2011)	oxygen stripping	0.668	0.0175
Wang (2014)	toluene stripping	0.875	0.0114*

* - recalculated to O₂-H₂O system according to Eq. (1)

Some authors performed $k_L a$ measurements using aqueous systems with modified physical properties [9] or with packings of different geometry [8] what enabled more elaborated dimensional or dimensionless correlation of the data. While dependence of $k_L a$ on the gas velocity up to the loading point was found negligible [7], significant dependence of the $k_L a$ on the total packed bed height has been reported for random [10] as well as structured [7] packings. Large difference in experimentally obtained $k_L a$ values can be observed and the recent urge for the measurement and evaluation method standardization [11], [12], [13] is therefore comprehensible.

Many mass-transfer models for packed columns have been developed on the basis of the experimental data and theoretical considerations; summary of the models developed until 2005 can be found in [14] and recent actualization on the limited topic of the liquid side mass-transfer coefficient models on structured packings in [15]. The $k_L a$ data calculated out of several models [16], [7], [17], [18], [19], [8] for several conditions treated in this work are depicted in the Figs 2-2a,b,c.

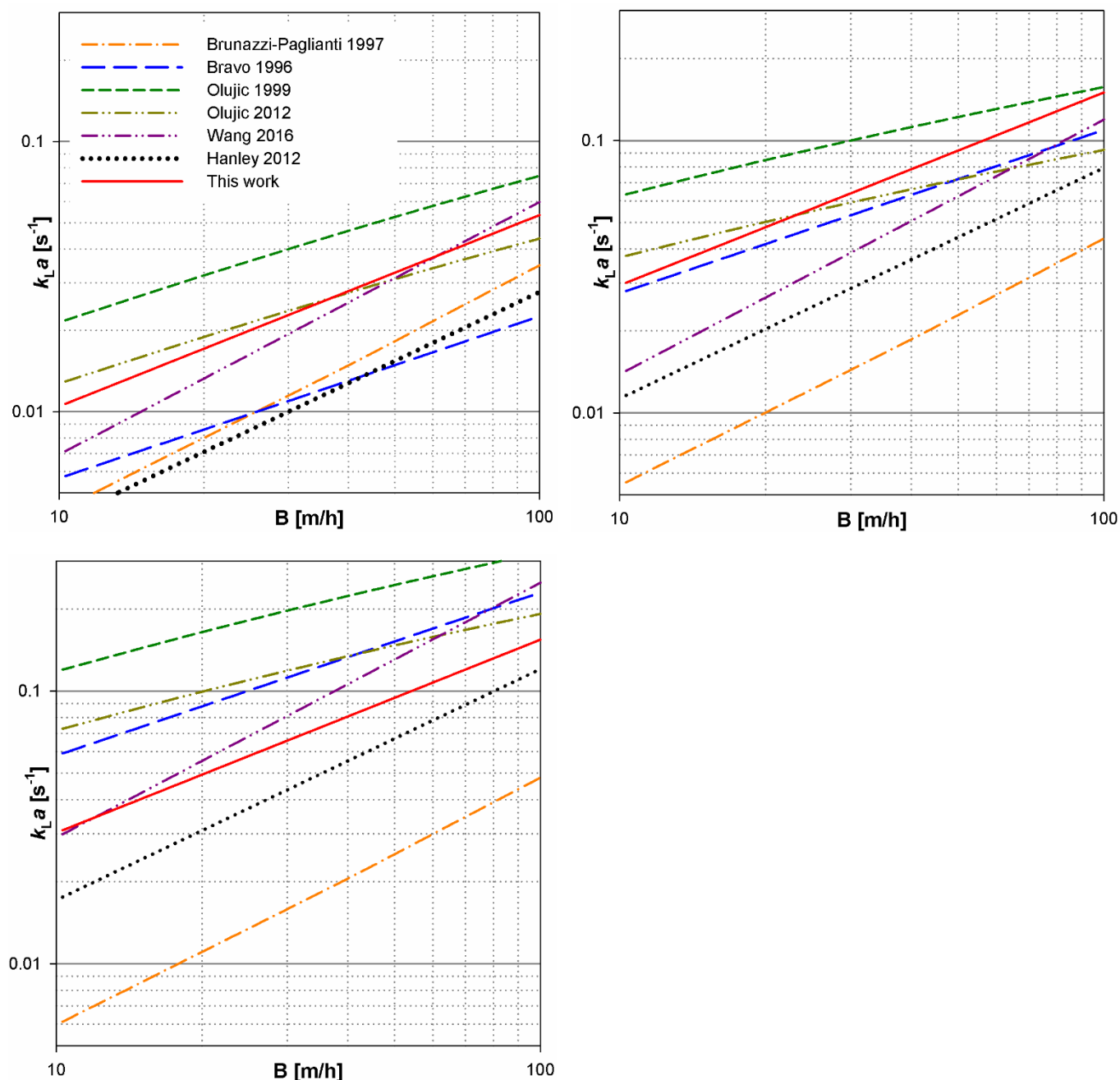


Fig. 2 – Liquid phase volumetric mass-transfer coefficient, $k_L a$. Prediction of the literature models. (a) Mellapak 250.Y ($\text{H}_2\text{O}-\text{O}_2$). (b) Mellapak 250.Y ($\text{MeOH}-\text{O}_2$). (c) Mellapak 500.Y ($\text{MeOH}-\text{O}_2$)

The $k_L a$ data predicted by the models for water (Fig. 2a) exhibit similar scatter as the experimental data (Fig. 1). The scatter increases when non-aqueous system (Fig. 2b) and different packing (Fig. 2c) is introduced, demonstrating unsatisfactory knowledge of the k_L dependence on the physical properties of the liquid and size of the packing.

The data obtained with the water-inert gas system serve often as an experimental base for development of mass-transfer models which are usually assumed to be valid also for distillation conditions. However, water is a substance with high surface tension, far from the properties of the organic systems treated by distillation. There can be anticipated that the organic liquid phase in distillation experiments behaves

significantly differently in terms of the liquid phase mass-transfer than aqueous systems under the desorption conditions.

Only few papers report on absorption/desorption from/into organic liquids in any packed device [20], [21]. Desorption experiments with organic liquid phase have been performed recently at our workplace using the wetted-wall column [22], however, although the device was often taken as a basis for development of mass-transfer correlations for structured packings, the effect of the liquid side hydrodynamics on the liquid-side mass-transfer coefficient is completely different.

There is currently no study dealing with absorption/desorption from/into organic liquids for structured packings published in the open literature. This article aims to fill this gap.

3 Theoretical background

The data for absorption of different gases into the liquid on the packing has been many times successfully correlated using penetration theory based on the imagination of unsteady diffusion of the solute into the liquid behaving like a semi-infinite slab for the period of exposure time τ after which the liquid is well mixed. The corresponding mass-transfer coefficient, k_L , is obtained through time averaging of the molar flux as

$$k_L = \sqrt{\frac{4D}{\pi\tau}} \quad (3)$$

As can be seen, the penetration theory doesn't directly describe either the influence of the liquid flow rate or the physical properties of the liquid on the mass-transfer rate. Their possible effect can be anticipated to be hidden in the value of the exposure time which can be assumed to be dependent on the liquid film hydrodynamics. The lack of prediction tools for determination of the exposure time on the basis of viscosity and surface tension can be attributed to relatively low number of experiments performed with different solvents.

Forces acting on the liquid film flowing down the column on the packing and forming its velocity profile and thickness are the inertial, gravitational, viscous and surface ones. The hydrodynamics of the film is

therefore described by the values of *Reynolds*, *Froude* and *Weber* numbers, Re , Fr , We . The relation of the hydrodynamics to the mass-transfer is usually expressed by introduction of the *Sherwood* and *Schmidt* numbers, Sh , Sc . The final correlation can take dimensionless form

$$Sh = C_1 Re^{C_2} Fr^{C_3} We^{C_4} Sc^{C_5} \quad (4)$$

4 Experimental

Experimental set-up

Experiments have been carried out in the PVC-U column with of an i.d. of 150 mm (see Fig. 3). The packing was stainless steel Mellapak 250.Y, 350.Y and 500.Y with height of the element equal 0.21 m provided by Sulzer in the year 2000 and equipped with plain sheet extra wall-wipers designed at the workplace (see Fig. 4)¹. The liquid phase flowing in the closed circuit has been pumped from the main tank into the column by centrifugal pump P1 through the set of turbine flowmeters F1, F2 and distributed over the packing by pipe-type distributor with 25 openings (1415 openings/m²). The liquid flowing from the packing has been collected in the liquid collector / chimney gas-distributor and flew back to the main tank by means of gravity. The nitrogen from the cylinder has been fed under the gas-distributor. Its flow was controlled by control valves.

When tap water was the liquid phase, the batch of 300-400 l was maintained at constant temperature of 20 °C and saturated with oxygen in the agitated 500 l stainless steel tank H1.

In the case of alcohols the main storage tank was a 30 l PVC-U vessel H2 submerged in the H1 tank filled with water. The water served as a supportive heat transfer medium and also as a safety measure for the case of a major leak from the vessel. The alcohols used were of p.a. grade. The vessel was also connected to the saturation loop made of i.d. 25mm glass tube with a total length of about 3 m in which the liquid was cooled down and saturated with oxygen. The flow of the liquid was driven by centrifugal

¹ The extra wall-wipers were made from the 0.1 mm thick stainless steel band with the total height of 40 mm. The lower half was cut to form downward pointing triangles. The tips of the triangles were manually bended into the packing element. The extra wall-wipers were placed just under the lower original ones.

pump P2 and cooled by means of the plate heat exchanger. The oxygen with a flow of approximately 10 nl/min was introduced into the pipeline by i.d. 4 mm blunt ended steel capillary.

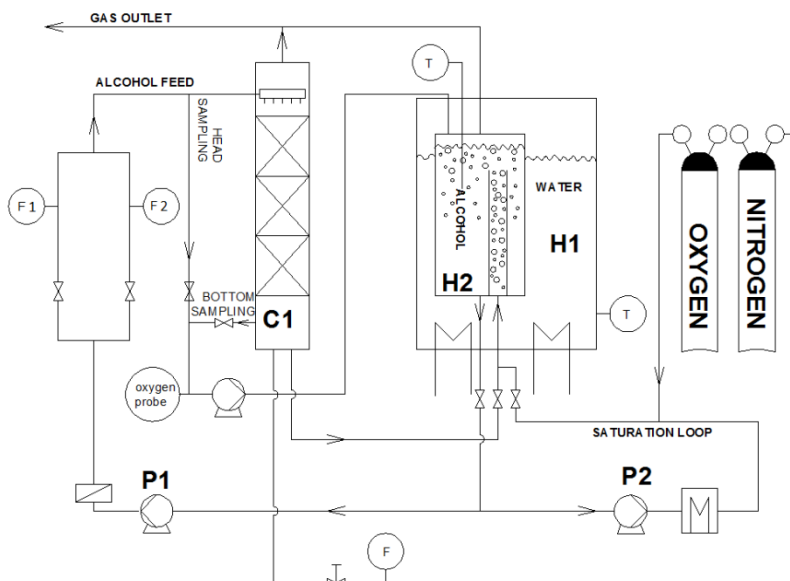


Fig. 3 – Sketch of the experimental apparatus.



Fig. 4 – Packing with the extra wall-wipers.

Sampling and analytics

The sample of the liquid from the head had been taken from the pipe-line feeding the liquid distributor. The sample from the bottom had been taken from five points of the liquid collector. Sampling lines were switched by the three-way valve to the oxygen probe overflown by the sample. Sampling rate from each line was 6 g/s.

Optical oxygen probe (PSt3, PreSens) was calibrated using its response in the environment of the measured liquid saturated with nitrogen, air and oxygen at 20 °C. The probe is manufactured for measurement of the oxygen fugacity in the gas or aqueous media in the fugacity range of 0 – 21 kPa, but it proved to be stable in the alcohol environment with the daily drift of 1.5 % on average. In the environment of alcohols the probe did not provide linear response with the oxygen fugacity in the entire 0 – 100 kPa range and the second order polynomial calibration curve was therefore constructed during the calibration process and further utilized for evaluation of the results. The accuracy of the probes is assumed to be around 1.5 % for the 100 kPa oxygen fugacity and around 10 % for the 0.5 kPa oxygen fugacity.

Table 2: Physical properties of the systems

	ρ [kgm ⁻³]	η [mPas]	σ [mNm ⁻¹]	D_{O_2} [m ² /s]	H [Pa]
<i>MeOH</i>	794	0.579	22.5	5.34x10 ⁻⁹	2.46x10 ⁸
<i>EtOH</i>	790	1.207	21.6	3.61x10 ⁻⁹	1.70x10 ⁸
<i>PrOH</i>	804	2.215	23.7	2.35x10 ⁻⁹	1.47x10 ⁸
<i>WATER</i>	998	1.002	72.8	1.90x10 ⁻⁹	4.05x10 ⁹

The physical properties of the systems used for the calculations are summarized in the Table 2. The densities have been calculated according to the Yaws et. al. [23], viscosities according to the Reid et. al. [24], surface tension according to the Yaws [23], the diffusivities have been calculated according to the Tyn and Calus method [25] and the solubility data were taken from APV86 HENRY database of Aspen SW.

Volumetric mass-transfer coefficient, k_La , evaluation

As the samples were not taken out of the packing there was necessary to subtract end-effects formed by 1) mass-transfer in the space between the packing and the liquid distributor as well as between the packing and the liquid collector and in the liquid collector itself; 2) different gas-liquid behavior of the packing close to its upper and bottom edge and the middle section of the packed bed. The subtraction was performed by measurements with ‘short’ and ‘long’ bed under the same liquid flow rate, as discussed e.g. by Billet [26]. The distance of the liquid distributor and liquid collector from the edge of the packed bed was kept constant for both cases. The k_La for given liquid flow rate with the end-effects subtracted has been evaluated using

$$k_La = \frac{u_L}{H^{long} - H^{short}} \ln \left(\frac{f_h^{long} f_b^{short}}{f_b^{long} f_h^{short}} \right) \quad (5)$$

For measurements performed with water the ‘short’ bed consisted of two packing elements while the ‘long’ bed consisted of four, six or eight elements. For the measurements with methanol and ethanol the maximum number of elements forming the ‘long’ bed were three, otherwise the bottom fugacity of

oxygen in the liquid was below that the probe was able to determine with reasonable accuracy. For measurements with all three alcohols the ‘short bed’ therefore consisted of only a single element ($H = 0.21$ m) in order to keep possibility to check agreement of the results obtained with at least two heights of the ‘long’ bed (two elements: $H = 0.42$ m, three elements: $H = 0.63$ m) and to keep procedure for all alcohols constant. The $k_L a$ data evaluated from the measurements with the ‘long’ bed consisting of two elements (together with the data measured with single element ‘short’ bed) are further denoted as “2-1” while the data evaluated from the measurements with three elements ‘long’ bed (together with the data measured with single element ‘short’ bed) are denoted “3-1”. Intense mass-transfer in the liquid collector and mass transfer in subtracted part of the packing was also found sufficient for saturation of the nitrogen with alcohol vapors.

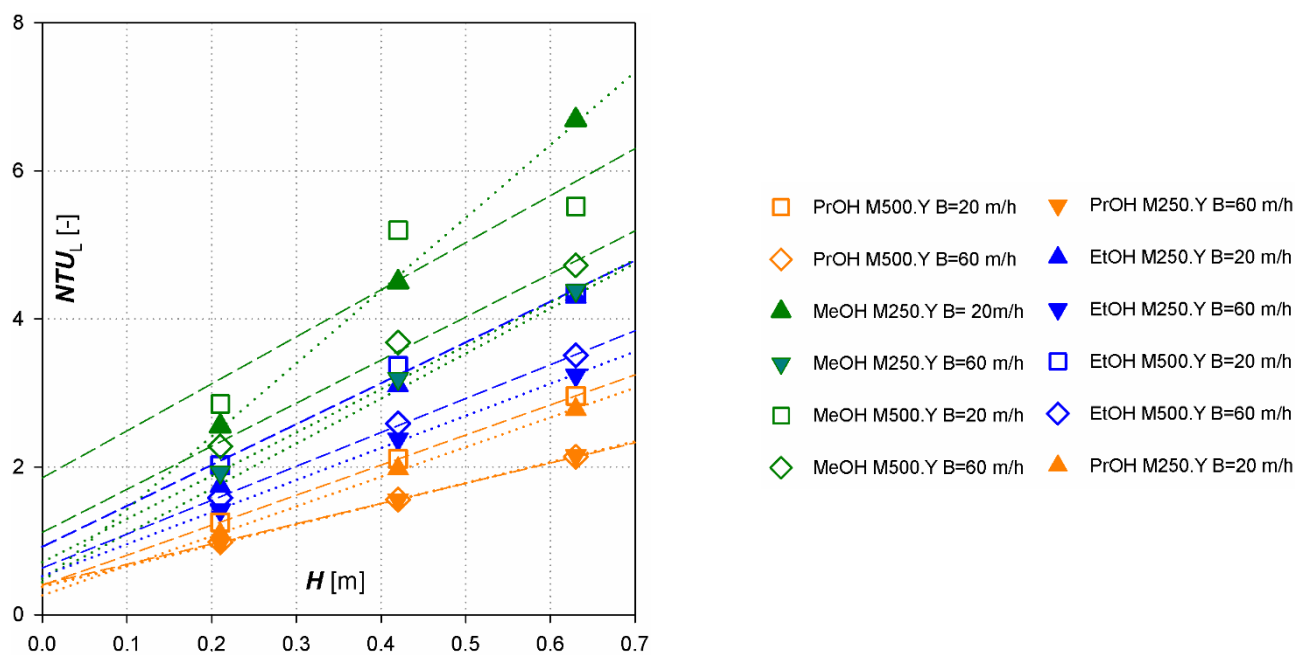


Fig. 5 – End-effect evaluation. NTU_L dependence

on the height of the packed-bed, H .

Legend for Fig. 5

Following treatment by Billet [26], the raw number of transfer units in the liquid, NTU_L , was evaluated out of the head and bottom fugacity of the oxygen for all three heights of the packed bed and $NTU_L(H)$ dependence was fitted with a straight line. For several conditions it is depicted in the Fig. 5.

$$NTU_L = \ln \frac{f_h}{f_b} \quad \text{Eq. (6)}$$

With exception of the outlying data for Mellapak 500.Y and B=60 m/h one can observe that 1) the dependences $NTU_L(H)$ is linear; 2) these dependences intercept the vertical axis in the limited range (around value of $NTU_L = 1$).

This suggests that 1) all the external and internal end-effects affecting the measurements with 'long' bed configuration are completely developed already for the 'short' bed, and can be therefore subtracted according to (5); 2) the end-effects are of similar magnitude for all conditions.

5 Results

The PVC-U i.d. 150 mm column was a new experimental set-up and therefore the measurement method has been first thoroughly verified by the experiments with water on Mellapak 250.Y packing. A significant dependence of the obtained $k_L a$ on the packed bed height as well as noticeable wall flow have been observed. Both phenomena have been suppressed by installation of additional plain sheet wall wiper on each packing element – after its installation the dependence of $k_L a$ on the height of the packed bed almost vanished and $k_L a$ values nearly matched those found for the i.d. 300 mm column (Fig. 6a).

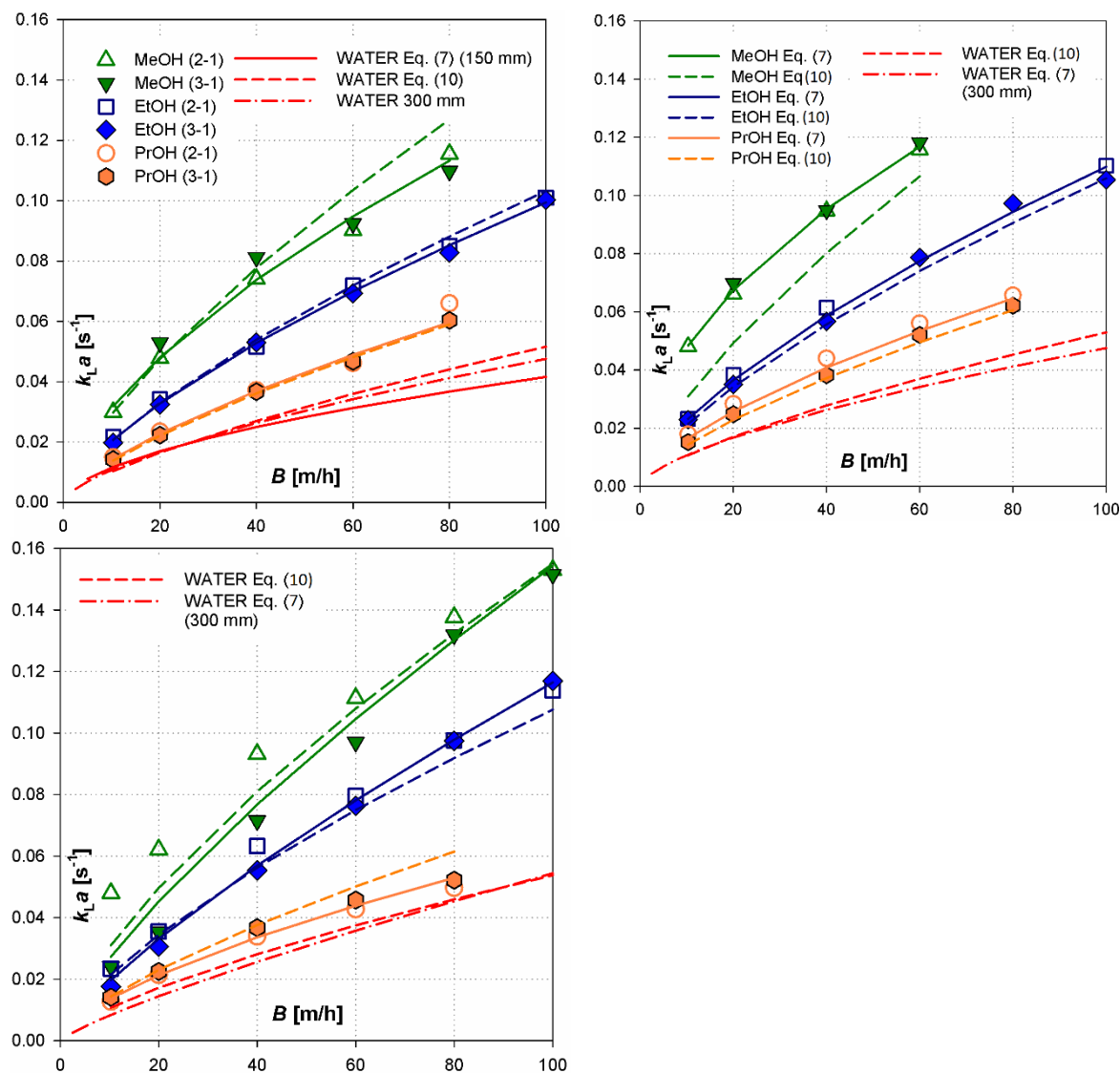


Fig. 6 – Experimental data and their correlation (Eqs. 7, 10). (a) Mellapak 250.Y. (b) Mellapak 350.Y. (c) Mellapak 500.Y.

Measurement of the $k_L a$ with methanol proved difficult due to its high degree of desaturation and corresponding low values of the bottom oxygen fugacity, which determined the lowest achievable liquid flows. High oxygen solubility in alcohols necessitated introduction of the maximum achievable stable flow of the nitrogen and it became a limiting factor for the achievement of the valid data for high liquid superficial flow.

Conversely, liquid collector / gas distributor hydraulic behavior was a limiting factor for achievement of high liquid superficial flows with n-propanol.

The measured $k_L a$ data for all three packings and all three alcohols in the dependence on liquid superficial velocity are depicted in the Fig. 6a,b,c. The numbers in the brackets depict number of

elements for ‘long’ and ‘short’ bed. Out of the 97 obtained experimental $k_L a$ values 5 were discarded as outlying. The data have been correlated with average *R.S.D.* of 8% by the equation

$$k_L a = C_1 B^{C_2} \quad (7)$$

with optimized constants provided in the Table 3.

Table 3: Constants of the Eq. (7)

	Mellapak 250.Y		Mellapak 350.Y		Mellapak 500.Y	
	C ₁	C ₂	C ₁	C ₂	C ₁	C ₂
MeOH	0.00754	0.618	0.00151	0.499	0.00463	0.761
EtOH	0.00416	0.689	0.00472	0.684	0.00319	0.781
PrOH	0.00286	0.693	0.00355	0.662	0.00298	0.656
WATER	0.00320	0.557	0.00243	0.720	0.00124	0.821

The $k_L a$ values measured for “2-1” and “3-1” configurations are almost identical for the most experimental conditions, proving successful subtraction of the end-effects. The behavior also suggest sufficient stripping gas saturation with alcohol vapors before entering the measurement section and maintenance of the sufficiently low oxygen concentration in the stripping gas in the frame of the bed. Only exception are the data for the methanol and Mellapak 500.Y packing (Fig. 6c).

The $k_L a$ values for the same liquid superficial velocity and the same alcohol do not increase systematically with packing geometrical area. Such behavior has been observed for water [5] as well. Assuming increasing effective interfacial area with the geometrical one this finding implies decreasing values of the mass-transfer coefficient, k_L . This in turn can be explained by decreasing liquid interstitial velocity on the individual packing sheet as the same liquid flow needs to be distributed on the higher number of sheets for the packing with higher geometrical area.

Mass-transfer coefficient, k_L , evaluation

The data processing consisted of extraction of the mass-transfer coefficient k_L from $k_L a$ data

$$k_L = \frac{k_L a}{a_{CH}} \quad (8)$$

The effective area for the conditions of $k_L a$ measurement (physical absorption) is not known and its closest experimentally achievable approximation was assumed to be the effective area, a_{CH} , obtained for the same packing types by the chemisorption of CO_2 into aqueous solution of sodium hydroxide [5]. The values were calculated according to (9) with constants from the Table 2, the data are graphically presented in the Fig. 7.

$$a_{CH} = C_3 B^{C_4} \quad (9)$$

Table 4: Constants of the Eq. (9)

PACKING	C_3	C_4
Mellapak 250.Y	135	0.110
Mellapak 350.Y	206	0.0626
Mellapak 500.Y	186	0.112

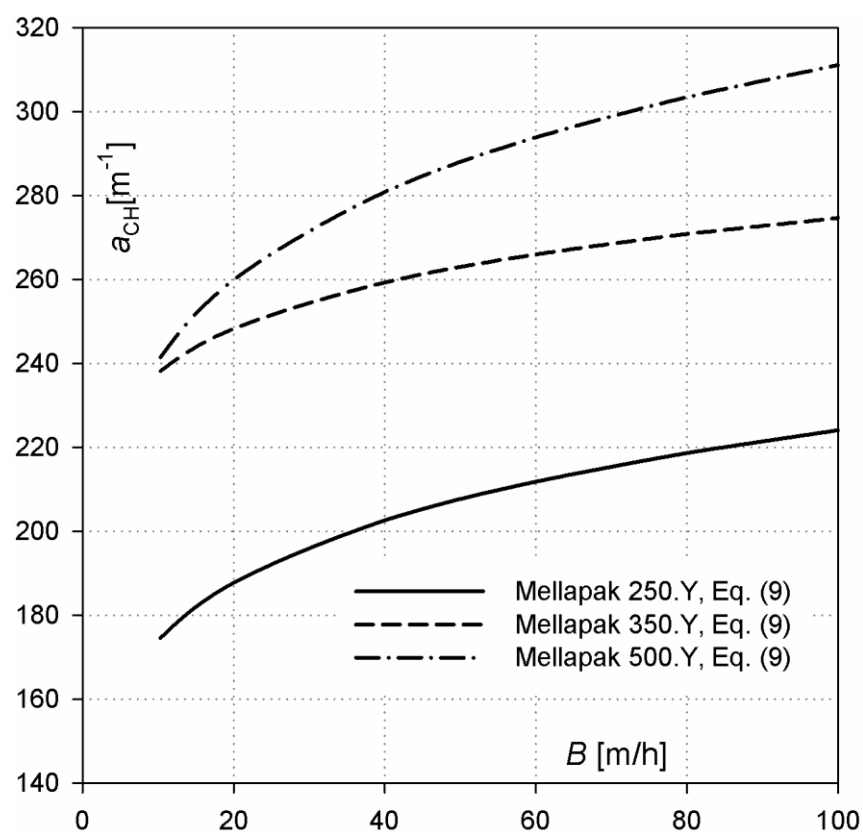


Fig. 7 – Effective interfacial area, a_{CH} , from chemisorption experiments.

Usage of chemisorption area, a_{CH} , for extraction of the k_L from $k_L a$ data mostly proved to provide better match of developed correlations to the experimental data than usage of geometrical area, a_G .

Mass-transfer coefficient, k_L , dimensionless correlation

The k_L data for all alcohols, water and for all packings have been correlated in the dimensionless form

$$Sh_L = 0.207 \cdot Re^{0.798} Fr^{-0.563} We^{0.511} Sc^{0.528} \quad (10)$$

with *R.S.D.* of 12.1%, its success in data fit is shown in the Fig. 8. The similar absolute value of the powers at Froude and Weber numbers suggests that the effect of the inertia forces connected with these numbers cancels and the numbers can be combined into the Eötvös (Bond) number. After the optimization the correlation results in

$$Sh_L = 0.586 \cdot Re^{0.695} Eo^{0.505} Sc^{0.468} \quad (11)$$

with only minor increase in *R.S.D.* to the value of 12.6%.

The dependence of k_L on individual physical quantities as predicted by the Eq. (10) in the form of simple power-law dependence

$$k_L \propto quantity^{power} \quad (12)$$

is shown in the Table 5.

Table 5: Dependence of k_L on physical quantities according to the Eq. (10).

Quantity	u_L	ρ	a	η	g	σ	D
Power	0.694	0.781	-0.872	-0.27	0.563	-0.511	0.472

The correlation Eq. (10) predicts dependence of the mass-transfer coefficient on the diffusivity with power of 0.472 and 0.538, respectively, close to the prediction of the penetration theory (which predicts 0.5). The correlation predicts the same magnitude but opposite effect of the gravity acceleration and the surface tension and relatively small effect of the viscosity. An increase in the packing area has significant negative effect on the value of the mass-transfer coefficient. This corresponds with the fact that the $k_L a$ values for the packings of different sizes (i.e. of geometrical areas) do not change

significantly. The 0.7 dependence of the k_L on the liquid superficial velocity is in approximate agreement with prediction of the models but is too low for explanation of the data evaluated from the distillation experiments. In the same time it is higher than predicted by penetration theory. Significant dependence on density suggests, that out of various extensive quantities it is in fact mass-flow what directly affects mass-transfer on the packing.

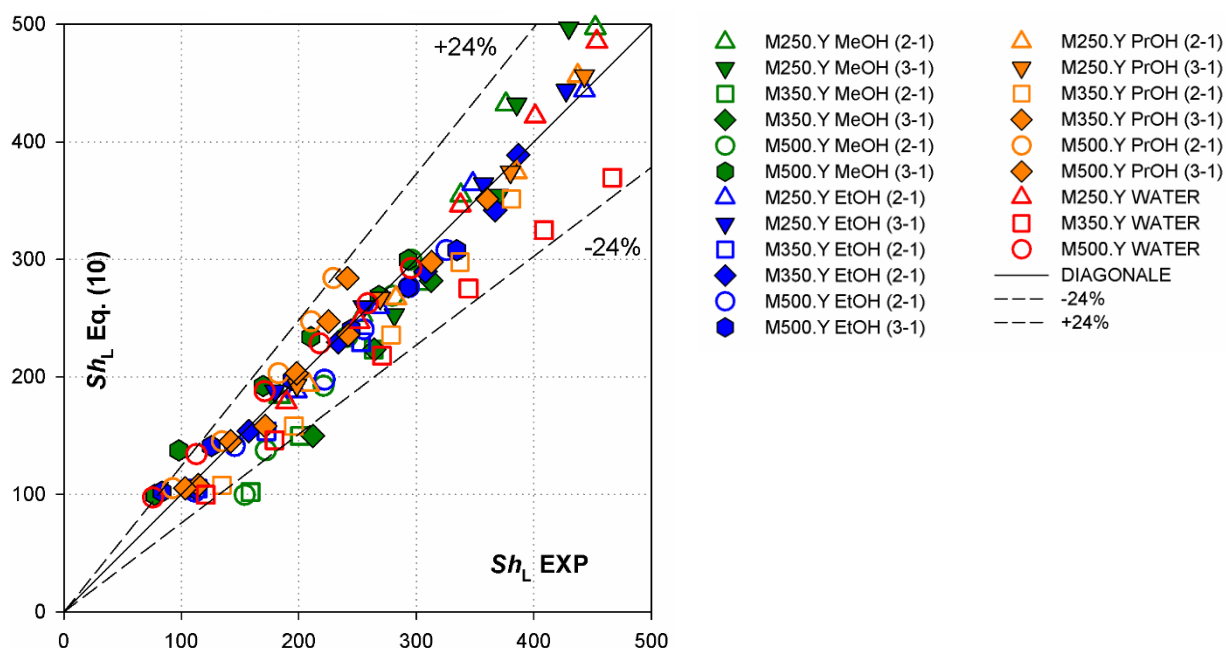


Fig. 8 - Sh_L parity graph. Experimental data and those calculated from Eq. (10).

Legend for Fig. 8

6 Conclusions

Volumetric mass-transfer coefficient, $k_L a$, were measured on three sizes of metal Mellapak structured packing (M250.Y, M350.Y, M500.Y) by desorption of oxygen from water and also from methanol, ethanol and n-propanol. When the packing was equipped with plain sheet extra wall-wipers suppressing significantly wall-flow of the liquid, the $k_L a$ values were found constant for different heights of the packed bed and for different diameter of the column.

Under otherwise similar conditions the $k_L a$ values are similar for all packing sizes as the higher interfacial area is compensated by lower mass-transfer coefficient with possible slight optimum found for M350.Y.

The mass-transfer coefficient has been successfully correlated for all packings, alcohols and water with R.S.D. of 12 % by dimensionless equations (10) and (11) based on the analysis of the forces acting on the liquid film flowing down the packing. The diffusivity power of just a little under 0.5 supports the

penetration theory model for mass-transfer description. The inertial and gravitational forces were found supportive for the mass transfer while the viscous and surface forces were found to be suppressive for the process. On the basis of the correlations the k_L values rise approximately with 0.7 power of the liquid load. The 0.7 power dependence of the k_L on the liquid superficial velocity is close to the prediction of many mass-transfer models but doesn't bring any new clue for explanation for the $k_L a$ behavior evaluated from the distillation experiments.

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8 List of symbols

a	interfacial area of the packing, (m^2/m^3)
a_G	geometrical area of the packing, (m^2/m^3)
a_{CH}	effective area measured by chemisorption experiments, (m^2/m^3)
B	liquid superficial velocity, (m/h)
C_i	multiplication constant or power of a quantity
c	molar concentration of a solute in the liquid, (mol/m^3)
D	Diffusivity, (m^2/s)
f	fugacity of the oxygen in the liquid, (Pa)
g	gravity acceleration, $g = 9.81 \text{ m}/\text{s}^2$
H	height of the packed bed, (m)
H	Henry constant, (Pa)
k_L	liquid phase mass-transfer coefficient, (m/s)
$k_G a$	volumetric gas phase mass-transfer coefficient, ($1/\text{s}$)
NTU_L	number of transfer units in liquid, $NTU_L = \ln \frac{f_h}{f_b}$

$k_L a$ volumetric liquid phase mass-transfer coefficient, (1/s)

u_L liquid superficial velocity, (m/s)

Greek symbols

α power of a quantity

ν kinematic viscosity, (m²/s)

η dynamic viscosity, (Pa·s)

ρ density, (kg/m³)

σ surface tension, (N/m)

τ exposure time, (s)

Subscripts

b bottom

h head

s superficial

i interstitial

i, j individual solutes

Superscripts

long long bed

short short bed

Dimensionless numbers

$Eu = \frac{\rho g}{a_{CH}^2 \sigma}$ liquid phase Eötvös (Bond) number

$Fr = \frac{u_L^2 a_{CH}}{g}$ liquid phase Froude number

$$Re = \frac{u_L \rho}{\eta a_{CH}} \quad \text{liquid phase Reynolds number}$$

$$Sc = \frac{\nu}{D} \quad \text{liquid phase Schmidt number}$$

$$Sh = \frac{k_L}{D a_{CH}} \quad \text{liquid phase Sherwood number}$$

$$We = \frac{\rho u_L^2}{a_{CH} \sigma} \quad \text{liquid phase Weber number}$$

Abbreviations

i.d.	inner diameter
M250.Y	Mellapak 250.Y
M350.Y	Mellapak 350.Y
M500.Y	Mellapak 500.Y
EtOH	ethanol
MeOH	methanol
PrOH	<i>n</i> -propanol
R.S.D	relative standard deviation

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