

## GAS-LIQUID HYDRODYNAMICS IN COUNTER-CURRENT COLUMNS WITH KATAPAK<sup>®</sup>-S and BX STRUCTURED PACKING

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The hydrodynamics in a counter-current gas-liquid flow structured packing column was experimentally investigated on a laboratory scale set-up. The experimental set-up included a 0.1 diameter column with a 1 m height of KATAPAK<sup>®</sup>-S or BX structured packing equipped with the necessary measurement instrumentation. The two fluids used in the experiments were water and air. The liquid phase mixing was investigated by the residence times distribution method, using a saturated NaCl solution as tracer.

### INTRODUCTION

Packed towers with gas-liquid counter-current flow are equipments very much utilized, generally, in distillation and rectification and, in recent years, in catalytic distillation and petroleum refining operations.<sup>1,2</sup>

Random packing is often either cylindrical (ring) shaped or half cylinder (saddle) shaped; these features and other details can provide relatively high surface area within the structure, increased resistance to deformation and represent factors that prevent nesting. The ordered packing can be made of woven or solid material which typically is corrugated and bundled into the column shell; most of these have perforations to help promote mixing and liquid spreading, as well as to increase surface area. The advantage of structured packing is that it eliminates much of the form drag associated with dumped packings. As opposed to the classical random packing, ordered packing offers low pressure drops without sacrificing efficiency or capacity, very high vapour and good turndown characteristics, with the result of an economy in column scale and operating costs.<sup>3</sup>

The ordered geometry evolved from the early Stedman packing to the modern Sulzer types, fabricated from gauze/sheet carbon steel (for noncorrosive services) or stainless steel, carbon, ceramics (resistant to high temperature) or various plastics. The high cost per unit volume of these packings was a disadvantage when compared with high efficiency random packings and with tray and, consequently, their utilization in the chemical and petroleum process industries was rather slow. The non catalytic Sulzer BX packing is often mentioned in the literature for absorption or rectification separations.

Combining reaction and distillation, reactive distillation is often carried out heterogeneously to avoid catalyst separation and recycling and to define exactly the position and height of the reaction zone.<sup>4</sup> The catalyst particle sizes used in catalytic distillation are usually in the 0.5–2 mm range and counter-current operation in fixed beds packed with such particles is difficult because of flooding restrictions. To immobilize the catalyst in the column/reactor several methods are utilized:<sup>4</sup> catalyst baskets on tray, catalytic random packing, catalyst enclosed in cloth wrapped in the form of bales or catalyst sandwiched between corrugated sheets of wire gauze (catalytic structured packing) which combine the features of a catalyst support with the advantage of structured separation packing. Such structures are made by Sulzer, called KATAPAK<sup>®</sup>-S and Koch-Glitsch called KATAMAX.<sup>5</sup>

Although structured packing have been known for almost 30 years, a limited amount of experimental data have been published in the open literature concerning their hydrodynamic and mass transfer characteristics and current models for predicting their performances have not been validated adequately.<sup>3</sup> As the performance characteristics of the structured packing became better understood and it was recognized that their efficiency/pressure drop ratios could be quite favourable, more applications followed.

Numerous attempts have been made to describe the hydrodynamic behaviour (pressure drop, loading and flooding limits, liquid hold-up) of packed columns operating as counter-current gas-liquid contactors. Pressure drop models for non catalytic structured packings have been developed independently at Delft and at the University of Texas in the Separations Research Program,<sup>6,7</sup> both being based on the preloading region where there is minimal interaction between the vapour and the liquid.<sup>8</sup> Gas-liquid hydrodynamics and mass transfer in counter-current columns packed with KATAPAK<sup>®</sup>-S structures was investigated by Moritz and Hasse,<sup>4</sup> van Baten *et al.*,<sup>5</sup> Ellenberger and Krishna,<sup>9</sup> Kolodziej<sup>10</sup> *et al.* Depending on the liquid flow rate Moritz and Hasse<sup>4</sup> identified two liquid flow regimes in a KATAPAK<sup>®</sup>-S structured packing, delimited by the so-called “load point”.

In this work an experimental investigation is presented of the hydrodynamics in a gas-liquid-solid counter-current structured packing column.

### EXPERIMENTAL SET-UP

The main element of the experimental set up is a column of approx. 0.1 m inner diameter filled with five KATAPAK<sup>®</sup>-S or BX packing elements disposed with an angle of 90° between the orientation of sheets in adjacent elements. In the hydrodynamic experiments, the solid particles inside the sandwiches KATAPAK-S consisted of glass spheres having similar size distribution with the catalyst Amberlyst 35 utilized for synthesis of gasoline additives.

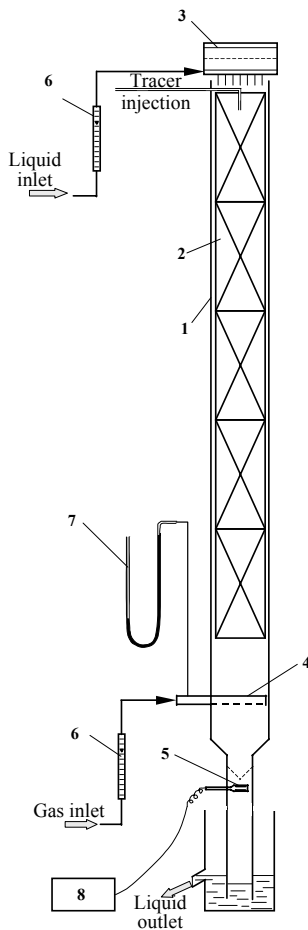


Fig. 1 – Experimental set-up:

- 1–column; 2–KATAPAK<sup>®</sup>-S elements;
- 3–liquid distributor; 4–gas distributor;
- 5–conductometric cell; 6–flowmeters;
- 7–manometer; 8–digital conductometer

The two fluids used were air and water flowing in counter-current. All the experiments were performed at gas pressures very close to the ambient one ( $\sim 1$  bar). The water was fed in column through a distributor with 37 spray nozzles of 2 mm diameter. The gas flow is fed to the bottom and leaves the top of the column at atmospheric pressure. The liquid and gas flow rates and temperatures at column inlet and outlet points were continuously measured by flow meters and thermometers. Also an inclined differential manometer was used to measure the pressure drop of the gas phase in the column.

The liquid hold-up was measured by the volumetric method. When steady-state conditions were established the dynamic liquid hold-up was measured by stopping the gas and liquid supply simultaneously; then the liquid in packed bed was collected in the bottom of the column. The hold-up was calculated as the liquid volume collected divided by the overall volume of the packing elements.

To measure the residence times distribution (RTD), a small quantity of saturated aqueous sodium chloride solution was injected in the centreline, at the top of the packed bed. A conductivity cell with approximately  $1 \text{ cm}^3$  liquid retention was used to measure the tracer concentration in the liquid leaving the last packing element. The cell was connected to the digital conductivity meter and further to a computer for data acquisition.

The column was operated at liquid loads of  $1.3 - 75 \text{ m}^3/(\text{m}^2\cdot\text{h})$  and gas loads of  $675 - 12000 \text{ m}^3/(\text{m}^2\cdot\text{h})$ .

## RESULTS AND DISCUSSION

### 1. Dry packing

A relation for pressure drop in dry structured noncatalytic packing is that proposed by Bravo et al.<sup>11</sup> and Rocha *et al.*:<sup>7</sup>

$$\frac{\Delta p_d}{H} = \frac{f \cdot \rho_g \cdot w_{g,e}^2}{S} \quad (1)$$

where  $\Delta p_d/H$  is the dry bed pressure drop per height of packing,

$$f = C_1 + \frac{C_2}{\text{Re}_{g,e}}, \quad (2)$$

$$\text{Re}_{g,e} = \frac{w_{g,e} \cdot S \cdot \rho_g}{\eta_g} = \frac{w_{g,s} \cdot S \cdot \rho_g}{\epsilon_p \cdot \sin \alpha \cdot \eta_g} \quad (3)$$

Combining equations (1–3) the final relation for dry pressure drop is:

$$\frac{\Delta p_d}{H} = \frac{C_1 \cdot \rho_g \cdot w_{g,s}^2}{S \cdot \epsilon_p^2 \cdot (\sin \alpha)^2} + \frac{C_2 \cdot \eta_g \cdot w_{g,s}}{S^2 \cdot \epsilon_p \cdot \sin \alpha} \quad (4)$$

For BX packing utilized in this work the best correlation, in accordance with relation (4), was:

$$\frac{\Delta p_d}{H} = 56.543 \cdot w_{g,s}^2 + 10.288 \cdot w_{g,s} \quad (5)$$

The experimental and calculated values for dry packing are represented in Fig. 2.

With the characteristics of BX packing utilized (side dimension of corrugation:  $S = 0.007 \text{ m}$ ; angle for corrugation channel:  $\alpha = 60^\circ$ ; void fraction of packing:  $\epsilon = 0.9$ ) the values for constants  $C_1$  and  $C_2$  are:  $C_1 = 0.133$ ;  $C_2 = 17.823$  and finally for the friction factor  $f$  was established the relation (Fig. 3):

$$f = 0.133 + \frac{17.823}{\text{Re}_{g,e}} \quad (6)$$

Ellenberger and Krishna<sup>9</sup> proposed, for dry pressure drop in catalytic packings, the relation:

$$\frac{\Delta p_d}{H} = f \cdot \frac{1}{d_h} \cdot \rho_g \cdot \left( \frac{w_{g,s}}{\epsilon_{oc} \cdot \sin \alpha} \right)^2 \quad (7)$$

Using this relation, the correlation of friction factor in respect to the Reynolds number, based on our experiments (Fig. 1) with KATAPAK<sup>®</sup>-S packing, was:

$$f = 1.6335 \cdot \text{Re}_{g,e}^{-0.3082} \quad (8)$$

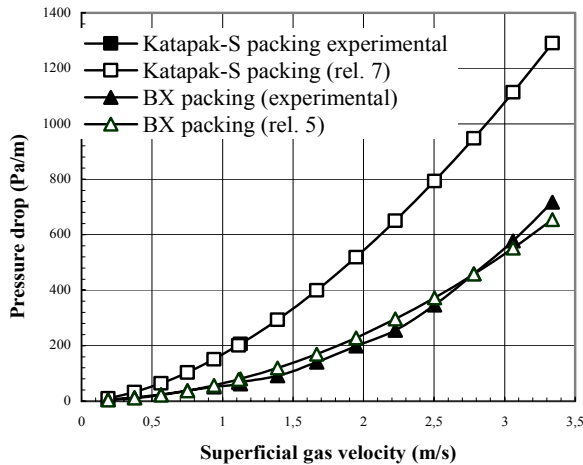


Fig. 2 – Pressure drop in dry structured packing.

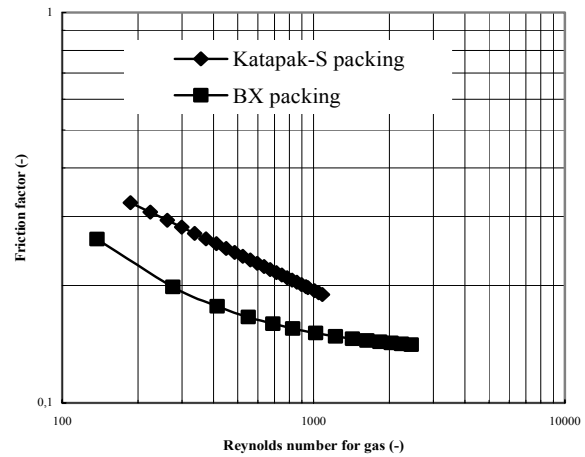


Fig. 3 – The friction factor correlations.

## 2. Irrigated packing

The gas phase pressure drop in an irrigated packing is higher than in a dry bed, this augmentation being explained by the presence of liquid in the interstices of the bed.

### 2.1. Liquid hold-up

The behaviour of the liquid hold-up in the structured packing studied for different liquid and gas loading is shown in Fig. 4. It is evident that, depending on the liquid loading, for a certain domain of superficial gas velocities the hold-up is constant, the point where the hold-up starts to increase being named loading point. Below the loading point the hold-up is function of the liquid velocity only. The region where there is an influence of gas rate is commonly known as loading region. The limit of loading for the two packings used in this

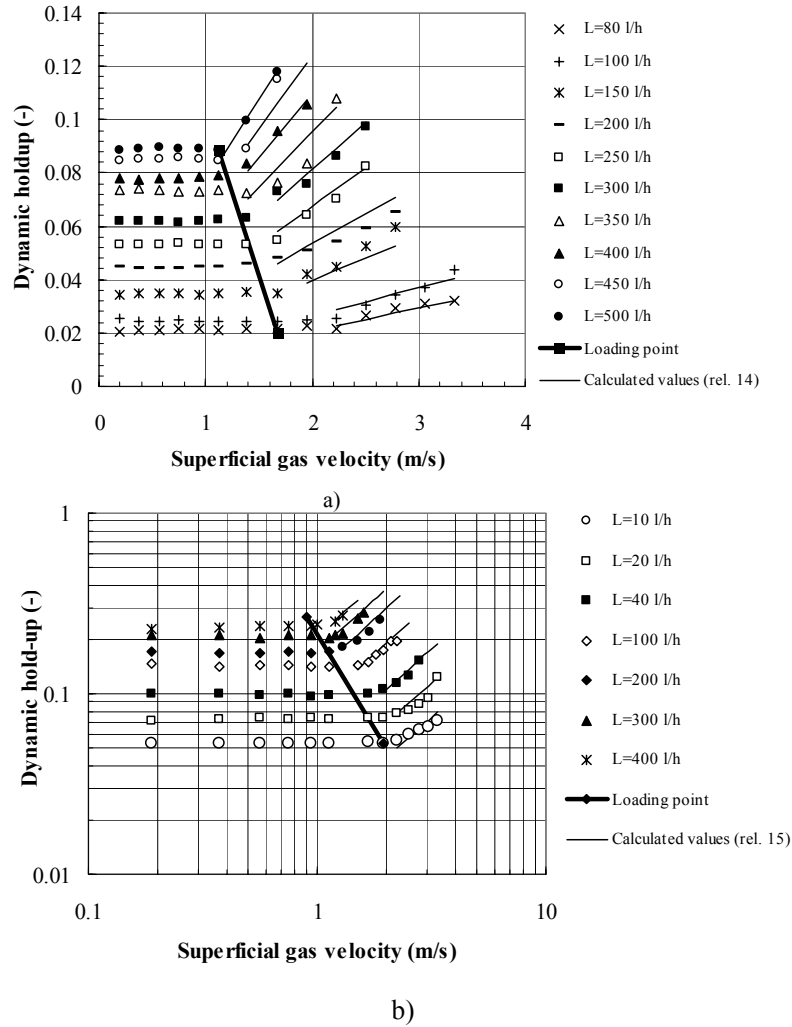


Fig. 4 – Liquid dynamic hold-up: a) BX packing; b) KATAPAK<sup>®</sup>-S packing.

work is represented in the Fig. 5, based on the method suggested in Fig. 4.

Based on the values measured in this work for the preloading zone, the correlations of the dynamic hold-up in respect to the liquid velocity are the following:

– for BX packing:

$$h_d = 1.0665 \cdot w_{l,s}^{0.6393} \tag{9}$$

– for KATAPAK<sup>®</sup>-S packing:

$$h_d = 1.226 \cdot w_{l,s}^{0.3907} \tag{10}$$

These correlations are represented in Fig. 5.

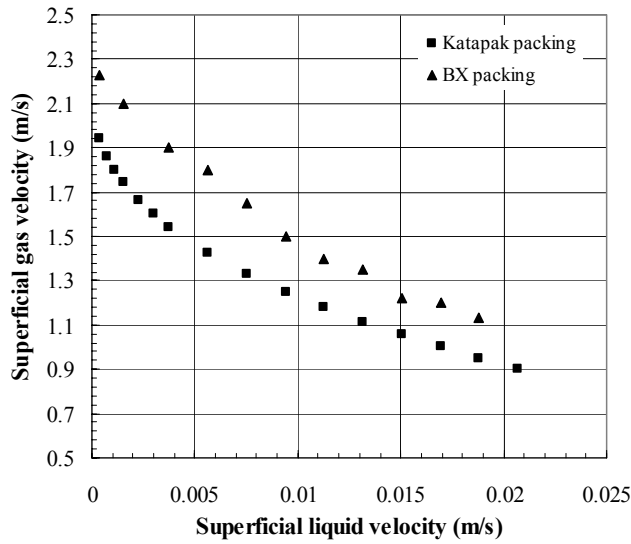


Fig. 5 – Superficial liquid and gas phases at loading limit.

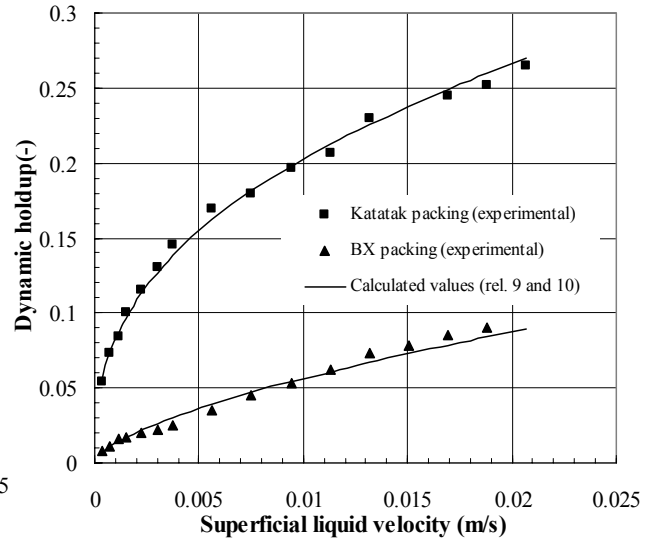


Fig. 6 – Dynamic hold-up for studied packing in preloading zone.

Billet,<sup>12</sup> cited by Stichlmair *et al.*,<sup>13</sup> correlate the hold-up for different packings by the relation:

$$h_d = 0.555 \cdot Fr_1^{1/3} \tag{11}$$

validated for air/water.

The Froude number for structured packing is defined by the relation.<sup>14</sup>

$$Fr = \frac{w_{l,s}^2}{S \cdot g} \tag{12}$$

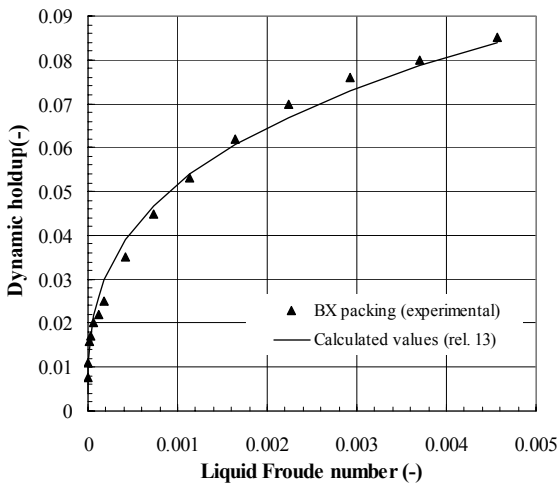


Fig. 7 – Dynamic hold up below the loading point.

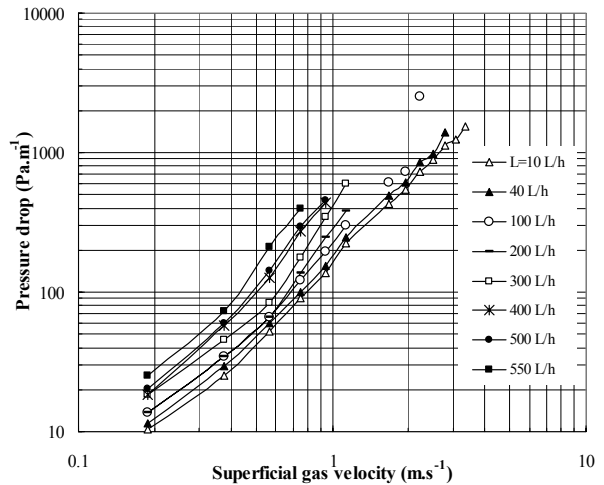


Fig. 8 – Measured pressure drop in KATAPAK®-S packing.

Using for the studied BX packing the Froude number, defined by relation (12), relation (13) was obtained utilized for calculated values of the dynamic hold-up in Fig. 7:

$$h_d = 0.4705 \cdot Fr_1^{0.3607} \tag{13}$$

In the above zone of the loading, the hold-up is affected by the gas velocity (Fig. 4) and increases with the gas rate at a constant liquid rate. For this region the dynamic hold-up was correlated by the relations:

– for BX packing:

$$h_d = 4.677 \cdot w_{g,s}^{0.854} \cdot w_{l,s}^{1.035} \quad (14)$$

– for KATAPAK<sup>®</sup>-S packing:

$$h_d = 1.949 \cdot w_{g,s}^{0.967} \cdot w_{l,s}^{0.557} \quad (15)$$

The corresponding values calculated with relations (14) and (15) are represented in Fig. 4.

## 2.2. Pressure drop

The pressure drop for irrigated packings has been studied extensively and a number of models have been proposed.<sup>11, 13-15</sup>

The specific gas phase pressure drop ( $\Delta p/H$ ) in irrigated packing increases with the increase of superficial gas and liquid velocities as seen from the experimental data presented in Fig. 8. Generally, the models proposed for the pressure drop calculation in irrigated packed beds are based on the correction of the dry bed pressure drop:

$$\frac{\Delta p}{H} = \frac{\frac{\Delta p_d}{H}}{(1 - C \cdot h_d)^n} \quad (16)$$

where  $C$  is an adjustable constant and  $n$  takes different values according to the authors.<sup>14</sup>

From our measurements for KATAPAK<sup>®</sup>-S packing, the values  $C = 0.65$  and  $n = 4.5$  were estimated for the parameters in this equation.

## 3. Liquid phase Residence Times Distribution (RTD) measurements

The RTD in the liquid phase was measured by injection of a pulse signal of NaCl solution at the liquid feed section and registering the NaCl concentration in the liquid at the bottom of the column. The RTD were determined at different liquid and gas flow rates. A typical measured residence times distribution is presented in Fig. 9.

To describe the mixing in the liquid phase, the axial dispersion model proved to be sufficiently accurate. The Peclet ( $Pe$ ) number, which is indicating the importance of the axial dispersion phenomenon in the column was estimated by fitting the measured values of the RTD density function,  $E_m(\theta)$ , to the expression:

$$E(\theta) = \frac{1}{2 \cdot \sqrt{\pi \cdot \theta}} \cdot \sqrt{\frac{w_1 \cdot H}{D_L}} \exp\left(-\frac{(1-\theta)^2}{4 \cdot \theta} \cdot \frac{w_1 \cdot H}{D_L}\right), \quad \theta = \frac{t}{\bar{t}} \quad (17)$$

With this aim the following square sum error function,  $\sigma$ , was used:

$$\sigma = \sum_{i=1}^N [E(\theta_i) - E_m(\theta_i)]^2 \quad (18)$$

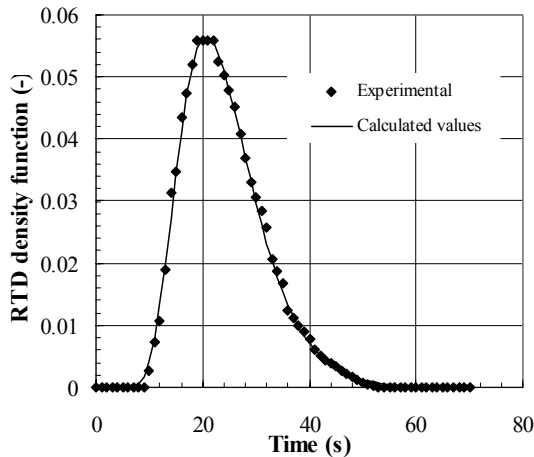


Fig. 9 – Experimentally measured and calculated RTD density function (liquid flow rate 0.3 m<sup>3</sup>/h; gas flow rate 25 m<sup>3</sup>/h).

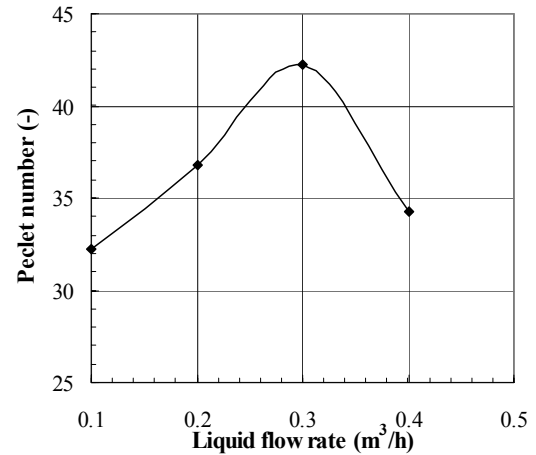


Fig. 10 – The dependence of the Pe number on the liquid flow rate (gas flow rate 10 m<sup>3</sup>/h).

The minimisation of the error function in respect to the Pe number was performed by the Marquardt algorithm.<sup>16</sup> A fairly good quality of the fit was obtained for all measurements (as in Fig. 9) so that the axial dispersion model can be considered as appropriate for the investigated working domain.

The dependence of the Pe number on the liquid flow rate presents a maximum, as in the Fig. 10, the maximum point approximately corresponding to the load point. Moritz<sup>4</sup> reported also a similar dependence.

## CONCLUSIONS

The results of an experimental study of the hydrodynamics in a gas-liquid counter-current flow column with BX and KATAPAK<sup>®</sup>-S structured internals were presented.

The gas phase pressure drop measurements, in dry and irrigated packing, loading point localization, liquid phase residence time distributions measurements, estimation of the Peclet number in the axial dispersion flow model and the development of correlation equations for several operating parameters were reported.

The liquid flow and mixing was described by the axial dispersion model. The values of the Pe number demonstrate a significant mixing intensity in the liquid phase, which has to be considered in the modelling of the column.

The obtained results are in good agreement with the published data of other authors.

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### Nomenclature

|                |  |
|----------------|--|
| a              | specific area, m <sup>-1</sup>                   |
| D              | axial dispersion coefficient                     |
| E              | RTD density function                             |
| d              | hydraulic diameter, m; $d_h = 4 \epsilon_{oc}/a$ |
| g              | acceleration of gravity, m·s <sup>-1</sup>       |
| H              | height of the packing, m                         |
| h <sub>d</sub> | dynamic liquid hold-up, dimensionless            |
| Δp             | pressure drop, Pa                                |



|           |   |
|-----------|---|
| Pe        | Peclet number, dimensionless              |
| Re        | Reynolds number, dimensionless            |
| S         | side dimension of corrugation, m          |
| t         | residence time of liquid in the vessel, s |
| $\bar{t}$ | mean residence time, s                    |
| w         | velocity, m s <sup>-1</sup>               |

*Greek Letters*

|               |  |
|---------------|--|
| $\alpha$      | angle with horizontal for corrugation channel, degrees |
| $\varepsilon$ | void fraction, dimensionless                           |
| $\eta$        | dynamic viscosity, Pa·s                                |
| $\theta$      | dimensionless RTD                                      |
| $\rho$        | density, kg·m <sup>-3</sup>                            |
| $\sigma$      | square sum error function                              |

*Subscripts*

|    |                        |
|----|------------------------|
| d  | dry                    |
| e  | effective              |
| g  | gas phase              |
| h  | hydraulic              |
| L  | longitudinal direction |
| l  | liquid phase           |
| m  | measured               |
| oc | open channel           |
| p  | packing                |
| s  | superficial            |

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