MODELLING OF MALDISTRIBUTION
IN STRUCTURED PACKINGS:
FROM DETAIL TO COLUMN DESIGN

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MAMMIE en PAPPIE
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The aim of this thesis was to determine the origin and extent of both small scale and large scale maldistribution in corrugated sheet structured packings and to develop a complete predictive model for hydraulic and separation performance of industrial scale columns. MONTZ-PAK B1-250 and RALU-PAK 250 YC were chosen out of many types of corrugated sheet structured packings as representatives of two extremes with respect to surface design. The first one has a shallow embossed surface without perforations while the second one has jalousie-like openings which makes it highly permeable for both phases.

A discrete cell model was used as basis for modelling of gas and liquid spreading. A cell size corresponding with the hydraulic diameter of the packings tested was chosen to incorporate small scale maldistribution effects. The calculated local liquid and gas loads enable predicting of the separation efficiency using a local equilibrium stage approach.

The communication between the cells in this detailed model, which proved to be strongly dependant on the packing surface design, was quantified with simple point source and tracer spreading experiments using two sheets of packing. From these experiments it was found that for RALU-PAK the liquid remains on the sheet on which it was introduced and that there is a strong communication between the two sheet sides through the slits. For MONTZ-PAK no communication is possible between sheet sides due to the closed surface. However we found a pronounced communication between adjacent sheets, with opposite channel orientation, via the
contact points resulting in an even distribution over adjacent sheets and a smaller lateral spreading than for RALU-PAK.

Tests with water (high surface tension) have shown that the spreading increases with a decrease of the liquid load due to forming of rivulets following the channel direction. For low surface tension liquids the spreading was less dependant on the liquid load.

For both packings gas distribution remained uniform in the case of a uniform initial distribution. Tracer spreading tests with both packings showed comparable results for the normal operating conditions.

Knowing the distribution parameters for the cells, the detailed model can be used for the prediction of the gas and liquid flow patterns in small columns and their influence on wetting and separation efficiency.

The predicted liquid distribution was checked on the basis of experimental data obtained in a 0.5 ID column with water. The liquid distribution was measured in 332 compartments (cross section = 2.5 x 2.5 cm²) in the bottom section. For MONTZ-PAK we found a much more uniform liquid distribution than for RALU-PAK, especially at low liquid loads. From calculation with the detailed model we found that the non uniformity of the liquid flow found for RALU-PAK is directly related to the large liquid spreading in this packing. In general we can conclude that in case of a uniform initial distribution large spreading causes less uniform liquid flow. However, a severe initial maldistribution is faster restored in case of large spreading.

Wetted areas were measured for water in a 0.45 ID column with a chemical absorption method. It was found that the wetted area for RALU-PAK was about 0.7 times as large as that for MONTZ-PAK, both measured at a low liquid load (1 mm/s). At high liquid loads (10 mm/s) the wetted areas for both packings become almost equal. From calculation with the detailed model it was found that the difference in wetting behaviour can be attributed to the difference in liquid flow pattern.

The HETP was measured in a 0.45 ID distillation column operated under total reflux with the system methanol-ethanol. The HETP for RALU-PAK (0.31 m) was about 1.5 times of that for MONTZ-PAK (0.19 m), measured in the normal operating range with a uniform initial distribution. Sealing half of the distributor results in a HETP of about 0.4 m for both packings. The detailed model shows that separation efficiency for both packings can be predicted on basis of the flow distribution and an equal value of the basic HETP.

The communication of the cells in the design-scale model depends on the sheet orientation only. The distribution parameters were determined by fitting the large scale spreading data obtained in a rectangular column simulator (3 x 0.5 x 4.4 m³).

The gas and liquid distributions were measured in 15 compartments in the top and bottom section respectively. For MONTZ-PAK we found a rather smooth large scale liquid distribution. Due to liquid accumulation around gaps between packing blocks small liquid peaks were found in the centre of the column. In the loading range these peaks developed to large peaks of about two times the average liquid load. From tracer point source distributions we obtained a spreading coefficient of
about 0.012 m for the normal operation range. In the loading range the spreading coefficient reduces to 0.006 m. The results of RALU-PAK were influenced by excessive wall flow due to the small depth of the column simulator, which makes interpretation of spreading results for this packing difficult.

For both packings the gas distribution was found uniform in case of a uniform initial gas distribution. From tracer point source tests a gas spreading coefficient was obtained for RALU-PAK of about 0.008 m which increases with the liquid load. The gas spreading coefficient for MONTZ-PAK was found to be independent of the liquid load and equal to 0.017 m. For both packings the gas spreading increases at gas velocities below 1 m/s.

The effect of liquid maldistribution on the separation performance is predicted with the design scale model in an analogous way as in the detailed model. The predictions of both models are pretty well in agreement with each other, indicating the suitability of the design scale model for simulations of large scale industrial columns.

Knowing the importance of the initial liquid-distribution a distributor model was developed to obtain realistic estimates for the initial distribution. For a single drip pipe the discharge liquid as function of the liquid level and type of discharge opening was studied and modelled. On the basis of the model for a single drip pipe a model for a two stage multi compartment distributor was developed. Simulations clearly show the influence of discharge apertures and compartmentalization in case of an inclined distributor. The distributor model can be incorporated into the design scale model to evaluate the influence of the distributor performance on the separation efficiency.
CHAPTER 1

INTRODUCTION TO MALDISTRIBUTION

1.1 INTRODUCTION

The key to success in separation of liquid mixtures by distillation depends on the creation and utilization of vapour-liquid contact area. The choice of equipment for contacting vapour and liquid is vast. There are three principal types of distillation equipment: trays, random packing, and structured packing. In terms of pressure drop per equilibrium stage, the corrugated sheet structured packings are beating both random packing and trays in performance, however at higher cost. Nevertheless, the low pressure drop proved to be highly rewarding in vacuum and atmospheric applications, so that, during the last decade, the corrugated sheet packings became an established vapour-liquid contacting device. Presently, structured packings are widely used for large diameter columns (up to 12 m.).

The interfacial area for mass transfer is created by spreading a vertical liquid stream radially over the packing surface as the liquid progresses down the column. The spreading occurs along the length of the sheets with corrugations inclined mostly under 45°. Entering the new packing element the liquid streams leaving an element
change direction for 90°. This means that after several layers of packing a thorough redistribution of the liquid in all directions should occur.

However, the expected uniform liquid distribution is not achieved easily, and the tendency of liquid to maldistribution seems to increase with increasing column diameter. Some early failures have shown that initial liquid distribution is crucial to the performance of structured packings. In fact, a packed column must rely on the distributor to spread the liquid uniformly over the packing and to ensure a complete utilization of available packing area. Unfortunately, a good initial distribution/redistribution cannot be maintained for long, and, in practice, the problem is solved by installing shorter beds with liquid catchers and redistributors in between. Such experiments raised questions about the source, extent, and the nature of liquid maldistribution in large diameter columns.

Large beds consist of layers comprising shorter sections (segments), and gaps between the longitudinal sections represent a major structural deviation from the situation present in pilot scale columns. Therefore, an experimental evaluation of column hydraulics on largest possible scale using air/water system was recommended as the only way to collect information which could help to understand the function of large diameter beds, a modelling effort based on equilibrium stage concept was considered to be a reasonable approach. However, to predict the hydraulic and separation performance of diverse types of structured packing, the small scale information, such as the effect of the packing surface design and physical properties on the liquid spread within the height of a packing element or segment, proved to be essential. Experiments have been set up to provide the desired information, and the experimental effort focused on two packing types representing extreme cases with respect to design of surface.

Since the performance of a packed bed depends strongly on the performance of the liquid distributor, an additional experimental effort has been done to collect information essential for modelling of the function of proven types and designs of liquid distributors.

When putting together all of the pieces it is expected that a complete predictive model will be formed to evaluate the performance of a column containing structured packings.

1.2 MALDISTRIBUTION IN DISTILLATION COLUMNS

In general maldistribution influences the performance of equipment in which nonlinear processes occur. In distillation this nonlinear process is the mass transfer between vapour and liquid. The liquid flows from top to bottom and is in contact with a counter current vapour flow. During the passage through the column there is a continuous mass exchange between both phases.

In order to demonstrate the influence of liquid maldistribution we will make use of a simple equilibrium stage model. The composition change over a theoretical stage is a nonlinear function of the liquid-gas ratio. Following Mullin [1], Huber and
Hiltbrunner [2] and Meier and Huber [3], we will split the column in two parallel columns, as shown in Fig.1.1. We calculated the composition change over 5 theoretical stages for total reflux conditions (liquid-gas ratio = 1). The vapours introduced at the bottom and the liquids introduced at the top are totally mixed which means that introduced compositions are equal for both parallel columns. Fig.1.2 shows the McCabe-Thiele diagrams for the left column with a liquid-gas ratio of 1.2 and the right one with a liquid-gas ratio of 0.8. So the average liquid-gas ratio is again equal to 1. The steps in the McCabe-Thiele diagram show the composition change, for the liquid on the X-axis and for the gas phase on the Y-axis, for each theoretical stage. It is clear that, in case of maldistribution, the mass transfer process in both parallel columns proceeds in a different way. That we will get a worse
Fig. 1.2 McCabe-Thiele diagram for left and right column in case of non-uniform distribution; the liquid-gas ratio for the left column is 1.2 and for the right column 0.8.

Fig. 1.3 Stage composition plotted in a McCabe-Thiele diagram (a) for a uniform distribution and (b) for a non-uniform distribution (average of left and right column in Fig. 1.2).

Separation efficiency becomes clear if we plot the average compositions of the left and right column in a McCabe-Thiele diagram and compare this with the McCabe-Thiele diagram of situation with an ideal distribution as shown in Fig. 1.3. This figure shows that, though the number of theoretical stages in both parallel columns is 5, we get a composition change over the column which is equal to the separation of 4 theoretical stages for a column with a ideal distribution. This example clearly demonstrates the influence of maldistribution on the separation efficiency.
1.3 HISTORY OF MALDISTRIBUTION RESEARCH ON STRUCTURED PACKING

The history of maldistribution studies on structured packings started about 30 years ago. It has its origin in the development and testing of the first Sulzer's gauze packing. The general conclusions of the first experimental investigation of (liquid) maldistribution in structured packing carried out by the researchers of Sulzer [2.,4] are summarized here:

- In small columns the effect of maldistribution on the separation efficiency is compensated by the lateral mixing of the gas phase.
- The packing cannot restore an initial liquid maldistribution. If the initial liquid distribution is good enough only insufficient wetting of the packing can cause reduction of the separation performance.
- No diameter effects are found when the packing (BX or MELLAPAK) is installed and operated properly.
- The separation efficiency obtained with maldistribution with partial reflux is equal to or better than with total reflux condition.

An alternative (hydraulic) approach, i.e. a liquid spreading investigation programme based on collecting the liquid below the packing, was initialized by F. Zuiderweg about 15 years ago at TU Delft.

The first results of experimental work carried out by Hoek [5] on structured packing were summarized in an article by Hoek, Wesselingh and Zuiderweg [6]. They suggest that the potential detrimental effect of the small scale maldistribution on the separation performance is largely compensated by radial mixing and that the flow equilibrium between the column's main body and the wall zone is rapidly established. This results in a much more uniform flow pattern in comparison to random packings.

A very rapid flattening of the liquid profiles, of initially maldistributed liquids, was observed. The profiles obtained for gauze packing BX, were smoother than those found for the corrugated metal sheet MELLAPAK.

Most of the experiments of Hoek were carried out in a 0.5 ID perspex column without a counter current gas flow.

The first maldistribution study on structured packing which gives some insight in the gas-liquid interaction, was published by Kouri and Sohlo [7]. They used BX packing to study the liquid and gas flow patterns as a function of the bed height (up to 3 m), flow rates and initial maldistribution. Their conclusions can be summarized as follows:

- The quality of the liquid and gas distribution in BX packing is nearly ideal in most cases.
- There is no effect of the liquid load, but an increase of the gas flow will promote liquid maldistribution.
- A radial reduction of the number of drip points will be compensated by the packing within four to six packing layers. However a pour initial liquid distribution seems to cause gas maldistribution. This effect becomes more evident at high gas loads.

A more thorough study by Stikkelman [8] of BX packing and a number of metal sheet structured packings, undertaken at Delft University of Technology in 1988 with a 0.5 ID perspex column, shed more light on the distribution behaviour of structured packings. In most cases two packing elements were enough to establish a 'natural flow', which was not detrimentally influenced by the gas within the normal operating range of the packing. However in the loading range a large scale liquid segregation was observed. These observations confirmed the assumption which was earlier made by Hoek [6]. Further he came to the conclusion that the gas maldistribution is negligible with respect to the liquid maldistribution.

1.4 AIM AND STRUCTURE OF THIS THESIS

Aim of the thesis

The aim of this thesis is twofold:

- To quantify the effects of surface design on the distribution behaviour and whether it is possible to calculate the effects of the surface design on the separation performance.
- To study the origin and extent of large scale maldistribution to solve distribution problems occurring in large diameter columns.

It was not possible to combine both objectives in one model; hence we presented two models with different fields of application. In the first model detailed information of the flow distribution can be taken into account. This so called detailed model can simulate the maldistribution on the scale of the hydraulic packing diameter and can be used as packing design tool. The second model takes only simplified information on the scale of the height of a packing element into account. This design scale model can be used as diagnostic and design tool for industrial applications.

Structure of the thesis

In this thesis the extent of liquid and gas maldistribution in both small (scale of the hydraulic diameter of the packing channels) and large scale and its influence on the separation performance is investigated.
For the small scale effects we mainly look at the influence of the design of the packing surface. For that reason we have investigated the distribution mechanisms for two types of structured packing with the same over all structure but with a principally different surface design.

The large scale effects of the research is focused on the large scale spreading behaviour of the two packing types. Also the effects of the initial maldistribution and the capability of the packing to restore from initial maldistribution, as well as the influence of structural deviations in the packed bed are considered.

In Chapter 2 the small scale spreading and mixing mechanisms for both liquid and gas phase are discussed. The experiments are carried out between two (or four) packing sheets. In this chapter we will show that we can simulate the experimental results with a discrete cell model, consisting out of small cells, in which we can take all details into account.

The validity of this so called detailed model is tested in Chapter 3 on the basis of (liquid) distribution data obtained in a 0.5 m ID column.

In Chapter 4 large scale distribution data obtained in a large scale column simulator (3x0.5x4.4 m³) are presented. For the liquid phase we will compare the distribution data with the distribution model based on small cells. Beside the detailed model we will present a design scale distribution model which is based on large cells. For both the liquid and the gas phase we will determine the parameters needed for the design scale model and compare the results with the experimental results.

In Chapter 2 to 4, only liquid and gas distribution are studied. In Chapter 5 we will take into consideration the wetting of the packing surface and look how the small scale liquid distribution influences the total wetted area available for mass transfer in columns containing structured packing.

The liquid distribution in the packed bed depends strongly on the initial liquid distribution. For that reason we also investigate the distribution characteristics of common liquid distributors. In Chapter 6 a theoretical model is presented that can predict the performance of gravity distributors.

Now that we have investigated the most relevant aspects of the liquid and gas distribution and the available wetted surface in the column, we can calculate the influence of these phenomena on the mass transfer performance. This is done in Chapter 7, where we compare the calculated mass transfer efficiency with the measured one obtained in a 0.45 m diameter distillation column with the system methanol-ethanol.

The last chapter, Chapter 8, demonstrates the practical aspects of the design scale model.
CHAPTER 1

NOTATION

SYMBOLS
x  Liquid mol fraction of the component
y  Vapour mol fraction of the component

INDICES
D  Distillate (top product)
B  Bottoms (bottom product)
l,r  Left or right column
1,...5  Number of theoretical stage

REFERENCES


2.1 INTRODUCTION

The last ten years a large variety of structured packings was introduced on the market. Most of these packings use well known and proven geometry consisting of corrugated metal plates, which lay counter course in such a way that they create channels crossing each other under an angle of 45° with column axis. A distinct quality of each packing should be a specially designed surface. The textures and the apertures of the packing surface are claimed by manufacturers to ensure extra positive effects, but from published sources it is not clear in which way the proprietary surface enhancements influence wetting and lateral spreading of the liquid.

To enable the prediction of liquid and gas distribution in a packed bed a flow model for both phases was developed in which all relevant effects can be taken into account.
The theory on which the model is based and the comparisons between calculated and measured flow and tracer distribution profiles for both liquid and gas phase are given in Sections 2.2 and 2.3 respectively.

2.2 LIQUID DISTRIBUTION BETWEEN TWO SHEETS OF CORRUGATED SHEET PACKING

We studied the liquid flow over the packing surface of two packings with a completely different surface design. Both packings (MONTZ-PAK B1-250 and RALU-PAK 250 YC) have a surface area of 250 m²/m³. The main difference between the two types is that RALU-PAK has an open surface, which allows the liquid to flow to the other side of the plate, and the MONTZ-PAK not. RALU-PAK is made of corrugated sheets provided with slits and a sharp corrugation ridge. MONTZ-PAK has a totally closed and highly roughened (shallow embossed) surface and a rounded corrugation ridge. Here we will show that these differences exhibit substantial influence on the lateral spreading and small scale mixing behaviour. Most other types of surface design of corrugated sheet packings can be placed between these two extremes.

2.2.1 EXPERIMENTAL SET UP

The liquid spreading characteristics were studied using a very simple apparatus shown in Fig.2.1. The essential part of this apparatus consists of two corrugated sheets which create the typical channel geometry of a structured packing. The liquid flow rate is measured by a rotameter and is introduced at the top of the packing sheet(s). The outgoing liquid is collected in a simple measuring device (Fig.2.1). The levels in the compartments of the measuring device are a measure of the local liquid load coming out of the packing.

For the majority of the spreading tests the point source was introduced centrally i.e. at a position coinciding with the location of the liquid collecting compartment No. 7. To investigate the phenomena at the packing periphery, the liquid point source was introduced in the periphery zone of the packing. For these experiments the packing was moved to the right in such away that the packing periphery was located above compartment No. 5 and the point source remains on the same location. Besides the measurements with water we also performed some experiments with a water-ethanol mixer and water-glycol mixture to investigate the influence of the liquid properties on the liquid spreading behaviour.

In addition to point source tests we also carried out some tracer tests to investigate whether there is a difference between the liquid spreading and mixing mechanisms. For these tests the liquid (water) was introduced at five equidistantly placed pour points with equal liquid flow rates. The point source in the middle was fed with a salt solution in stead of demineralized water. The salt concentration in
Fig. 2.1 Equipment for single/double plate point source and tracer source experiments

Each compartment was measured with a conductivity probe. In some tests the liquid was introduced in the periphery zone to investigate the effects at the packing periphery.

The packing sheets used in these tests were carefully degreased and washed. Each of above described experiments was started after prewetting the packing and sufficient stabilization time was allowed before samples were taken at new flow rates. At low liquid loads the distribution of the outlet flow was fluctuating with time. The average flow distribution calculated from 10 repeated experiments was reproducible. At higher flow rates the flow distribution proved to be very stable.
2.2.2 **THEORETICAL BACKGROUND OF THE LIQUID FLOW OVER CORRUGATED SHEETS**

It is very difficult if not impossible to describe exactly the liquid flow over a packing surface on a pure theoretical basis. Until now we only succeeded in determining the film flow direction over one single corrugated sheet under influence of gravity when neglecting the influence of the surface tension.

Such basic study was carried out by Zogg [1,2]. To facilitate the calculation of different angles with the vertical, we used vector analysis. An example is the angle of the liquid flow direction as it occurs in structured packing under influence of gravity. In Section 2.2.3 we will use this flow direction in the liquid distribution model.

![Diagram of a corrugated sheet](image)

*Fig.2.2 Schematic representation of a corrugated sheet, indicating major packing dimensions and flow angles*

The liquid flow over the packing surface results from a component of gravitational force in the channel direction and a component perpendicular to the channel ridge as shown in Fig.2.2. This figure shows also the channel dimensions and the different angles used for the calculation. The two components of the gravitational force are given here.
Vector $F_1$ in the channel direction

$$F_1 = \begin{pmatrix} 0 \\ \sin(\alpha) \\ -\cos(\alpha) \end{pmatrix} \rho_i \ g$$ \hspace{1cm} (2.1)

Vector $F_2$ towards the channel ridge

$$F_2 = \begin{pmatrix} \sin(\beta_1) \\ -\cos(\beta_1) \cos(\alpha) \\ -\cos(\beta_1) \sin(\alpha) \end{pmatrix} \rho_i \ g$$ \hspace{1cm} (2.2)

where $\rho_i$ is the liquid density, $g$ the gravity acceleration and $\alpha$ the channel angle with the vertical. The resultant $F_{1,2}$ of these two vectors yields the force vector that directs the liquid over the packing surface under the smallest angle with the vertical.

$$F_{1,2} = \cos(\alpha) \ F_1 + \cos(\beta_2) \ F_2 = \begin{pmatrix} \cos(\beta_1) \sin(\beta_1) \\ \cos(\beta_1) \cos(\beta_1) \cos(\alpha) - \cos(\alpha) \sin(\alpha) \\ -\cos(\beta_1) \cos(\beta_1) \sin(\alpha) - \cos^2(\alpha) \end{pmatrix} \rho_i \ g$$ \hspace{1cm} (2.3)

The angles $\beta$, $\beta_1$ and $\beta_2$ can be calculated from the channel dimensions ($h_{cor}$, $b$) and the channel inclination angle ($\alpha$).

$$\beta = \arctan\left( \frac{h_{cor}}{\frac{1}{2} \frac{b}{\tan(\alpha)}} \right)$$ \hspace{1cm} (2.4)

$$\beta_1 = \arctan\left( \frac{\tan(\beta)}{\sin(\alpha)} \right)$$ \hspace{1cm} (2.5)
\[ \beta_2 = \arccos(\cos(\beta_1) \sin(\alpha)) \]  

(2.6)

Because we assume that no further forces but the gravitational force and the friction force are involved in the calculation of film flow, the flow direction is the same as the direction of the vector \( F_{1,2} \). By taking the in-product of the vertical unity vector and the vector \( F_{1,2} \) we can calculate the angle with the vertical under which the liquid flows downwards.

In-product

\[
\cos(\phi) = \frac{\begin{pmatrix} 0 \\ 0 \\ -1 \end{pmatrix} \cdot \begin{pmatrix} 0 \\ 0 \\ -1 \end{pmatrix}}{|F_{1,2}|} = \frac{\cos(\beta_2) \cos(\beta_1) \sin(\alpha) + \cos^2(\alpha)}{\sqrt{\cos^2(\beta_2) + \cos^2(\alpha)}}
\]  

(2.7)

To calculate the angle of liquid spread in structured packing under influence of gravity only, we have to exclude the x-direction (see Fig. 2.2). The result is:

\[
\cos(\phi') = \frac{\cos^2(\beta_2) + \cos^2(\alpha)}{\sqrt{\cos^4(\beta_2) - \cos^2(\beta_2) \cos^2(\alpha) \sin^2(\alpha)}}
\]  

(2.8)

With this angle we can calculate the flow path for the flow over the channel ridge for MONTZ-PAK. For RALU-PAK we also have to take into account the angle between the slits and the channel sides, which causes a more vertical flow. Table 2.2 gives the results of the calculation if we use the packing dimensions from Table 2.1.

<table>
<thead>
<tr>
<th>( \alpha )</th>
<th>( b )</th>
<th>( \text{h}_{\text{cor}} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>45(^\circ)</td>
<td>0.032 m</td>
<td>0.012 m</td>
</tr>
</tbody>
</table>

Table 2.1 Main packing dimensions
<table>
<thead>
<tr>
<th>$B$</th>
<th>$B_1$</th>
<th>$B_2$</th>
<th>$\Phi$</th>
<th>$\Phi'$</th>
</tr>
</thead>
<tbody>
<tr>
<td>36.9°</td>
<td>46.7°</td>
<td>61.0°</td>
<td>31.0°</td>
<td>19.8°</td>
</tr>
</tbody>
</table>

*Table 2.2* Calculated angles

### 2.2.3 Liquid Distribution Model

The liquid distribution model presented here is based on a discrete cell model in which all cells are interconnected. This approach was first introduced by Albright [3] for simulation of liquid distribution in randomly packed columns, and later verified.

![Diagram of liquid flow paths](image)

*Fig.2.3* Observed liquid flow paths 1) flow path in the channel direction 2) flow path through the perforations 3) flow path over the channel ridge 4) flow path via the crossings
experimentally by Hoek et al. [4] and Stikkelman [5,6]. A similar but more probabilistic method, developed by Crine [7], also proved to be good for description of liquid distribution in random packed beds. However, a bed consisting of structured packings has a structure consisting of quite regular channels and liquid flows should be much less random in nature. Hoek et al. [4] have made an attempt to model liquid distribution in beds consisting of structured packings on the basis of point source studies with Sulzer BX and MELLAPAK 250Y.

The specific feature of the model developed here for structured packings is that the dimensions of the cells and their communication depend on the type, size and surface characteristics of the packing.

Based on visual observation we make a distinction between the following flow paths:

- Flow in the channel direction; i.e. liquid follows the channel direction and remains consequently in one (same) channel, shown schematically in Fig.2.3(1).

- The flow through the perforation (slits) in the packing sheets. The liquid flows in the direction of the smallest angle with the column axis (gravity flow) and the communication is possible between the two sheet sides, as illustrated in Fig.2.3(2).

- The flow over the channel side of the packing. Here the liquid also flows in the direction with the smallest angle of the column axis (gravity flow). The liquid has to change channel by flowing over the channel ridge as illustrated in Fig.2.3(3).

- The flow over the crossings of adjacent corrugations. At the crossings a stream may mix with the stream from the neighbouring sheet. Such a situation is illustrated schematically in Fig.2.3(4).

![Fig.2.4](image) Possible cell outlet flows for a cell with a right downward orientation and communication with the sheet in front. For a,b,c and d see text.
For all of these different flow paths in our cell model we assume that the liquid available in a cell can leave only via distinct outlet ports. This results in four different cell (outlet) flows as shown in Fig.2.4.

(a) The liquid can leave the cell under 45° in the direction of the channels.
(b) The liquid can flow in vertical direction to the cell located below.
(c) It can leave the cell under 45° opposite to the channel orientation. This simulates the flow over a corrugation ridge.
(d) It can flow in the channel direction to the sheet at the front or back side. This outlet simulated the transport via the crossing

Fig.2.5 Simulated liquid flow paths 1) flow path in the channel direction 2) flow path through the perforations 3) flow path over the channel ridge 4) flow path via the cross points
We can simulate a distinctive flow path by a repeating cluster of cell flows in series. The four different flow paths of Fig.2.3 can now be simulated as follows.

1. The flow in the channel direction is simulated by cell flow (a) as shown in Fig.2.5(1).
2. The flow through the perforation is simulated by a repeating cluster of cell flow (a) and cell flow (b) as shown in Fig.2.5(2). This figure also shows that in the case the liquid flows via cell flow (b) it enters a cell with a dark shade which means that the liquid changes sheet side.
3. The flow over the channel ridge is simulated by repeating clusters of cell flows (a) and (c) as shown in Fig.2.5(3). When the liquid flows via cell flow (c) it will change packing channel but will remain on the same sheet side.
4. The flow via the crossings is simulated by cell flow (d) as shown in Fig.2.5(4).

The flow in the channel direction (flow path 1), is important for both RALU-PAK and MONTZ-PAK and is promoted by high surface tension. The main flow for RALU-PAK is the (gravity) flow through the perforations (flow path 2). How often we have to choose cell flow (b) downwards compared to the cell flow (a) under 45° depends on the angle the flow through the perforations makes with the vertical axis. This also determines how often the liquid changes plate side. For RALU-PAK we found both experimentally and theoretically (Section 2.2.2) that the flow path through the perforations can be simulated by a repeating cluster of 1 x cell flow (a) + 3 x cell flow (b). Main flow path for MONTZ-PAK is the (gravity) flow over the channel ridge (flow path 3). We found that the flow path over the channel ridge can be simulated by repeating cluster of 2 x cell flow (a) + 1 x cell flow (c). So the liquid will change channel every 3 cells. The flow via the crossings (flow path 4) has not been observed with RALU-PAK.

How much liquid attributes to a certain flow path does not depend on the packing size or type only. Also the liquid load has an influence. Later on in this section we will see that the influence of the liquid load depends on the liquid properties.

As mentioned above the flow paths is modelled by a combination of certain cell flows. For the normal flow conditions the liquid assigned to the cell flow (b) (c) or (d) is determined by its splitting factor. The splitting factor determines the fraction of the total liquid flow that leaves a cell via a certain cell outlet flow. In the case that a cell flow is not used by a flow path its splitting factor is zero, which means that no liquid is assigned to this cell flow. Because the liquid assigned to the cell flow in channel direction (a) is determined from the mass balance over a cell, the liquid that does not flow via one of the cell flows (b) (c) and (d) automatically flows via cell flow (a). Each cell flow except the cell flow in the channel direction (a) has besides a splitting factor also a minimum value. When the total flow of liquid into a cell is below this limit no liquid is added to this cell flow. For the same reason as mentioned above this liquid then will follow the channel direction (cell flow (a)). In this way we can describe rivulet flow which dominates the flow behaviour at low
liquid loads. From observation we know that small rivulets prefer the flow path in the channel direction.

In the same way as the minimum critical value we defined a maximum flow rate for each cell flow except for cell flow (a). When the amount of liquid determined by the splitting factor exceeds the maximum critical value only the amount of liquid determined by the maximum critical value is assigned to the cell flow. The surplus of liquid (the difference between the amount of liquid determined by its splitting factor and its maximum value) is assigned via the mass balance calculation to the cell flow in the channel direction (a). This maximum limitation is important for the flow through the perforations (RALU-PAK). Observation indicated that only a limited amount of liquid can pass through the slits.

In the model the different flow regimes are described by the next three equations.

If the liquid load is lower than the critical minimum flow ($\text{Flow}_{\text{min}_n}$) of cell flow (b) (c) or (d) this cell flow is equal to zero.

$$\text{CellFlow}_{n,xyz} = 0 \quad \text{for } n=b,c,d$$  \hspace{1cm} (2.9)

If the flow rate is higher than the minimum critical flow we can write for the cell flows.

$$\text{CellFlow}_{n,xyz} = S_{1,n} \left( \text{Flow}_{xyz} - \text{Flow}_{\text{min}_n} \right) \quad \text{for } n=b,c,d$$  \hspace{1cm} (2.10)

The splitting factors of the cell flows ($S_{1,n}$) are not equal to zero only if a flow path makes use of this cell flow. So, for instance, the splitting factor of the cell flow through the perforation is 3 out of 4 times not equal to zero, because the liquid will leave a cell layer 3 times via the cell flow through the perforation (b) and 1 time via the cell flow in the channel direction (a).

In the case that the flow calculated with Eq.(2.10) is higher than the highest possible flow rate for that particular flow path, we can write.

$$\text{CellFlow}_{n,xyz} = \text{Flow}_{\text{max}_n} \quad \text{for } n=b,c,d$$  \hspace{1cm} (2.11)

Because the liquid that is not assigned to cell flow (b), (c) or (d) flows via the cell flow in the channel direction (a), we can write for the cell flow in the channel direction.

$$\text{CellFlow}_{n,xyz} = \text{Flow}_{xyz} - \text{CellFlow}_{b,xyz} - \text{CellFlow}_{c,xyz} - \text{CellFlow}_{d,xyz}$$  \hspace{1cm} (2.12)
When the liquid reaches the packing periphery it has two different possibilities.

a) To leave the packing and flow further downward via the column wall.
b) To flow along the packing periphery.

This is shown in Fig.2.6.

How much liquid stays at the periphery of the packing is determined by the maximum possible periphery flow \( \text{PerFlow}_{\text{max}} \). So if the local flow rate in a periphery cell exceeds the maximum periphery flow the surplus of liquid will leave the packing and will flow further downward via the column wall.

In this case the wall flow \( \text{WallFlow}_{\text{w}} \) can be calculated with:

\[
\text{WallFlow}_{\text{w}} = \text{WallFlow}_{\text{w}} + \text{Flow}_{\text{w, w}} - \text{PerFlow}_{\text{max}} \quad (2.13)
\]

And the new value of the periphery flow is equal to the maximum periphery flow.

\[
\text{Flow}_{\text{w, w}} = \text{PerFlow}_{\text{max}} \quad (2.14)
\]
If the inflow of liquid in a periphery cell is smaller than the maximum periphery flow all the liquid will stay on the packing and the wall flow remains unchanged.

In all cases the liquid in the periphery cell can flow back into the packing via the crossings (cell flow d) or can flow downwards via the packing periphery and overflow the channel ridge (cell flow c) and/or flows through the perforations (cell flow b). The extra liquid that flows back into the packing via crossings (cell flow d) is now calculated by the return factor (Rf).

\[
\text{CellFlow}_{d} \text{ per, } z = \text{CellFlow}_{d} \text{ per, } z \times \text{Rf Flow}_{\text{per, } z} \tag{2.15}
\]

The remaining liquid flows downward along the packing periphery.

If a wall wiper is installed (at \( z = z_{\text{wiper}} \)) and it works properly, a part of the liquid flowing down the wall is brought back to the periphery of the packing. The quantity of the liquid that remains on the wall is determined by the wiper leakage factor (Lf). The other part of the liquid re-enters the packing.

\[
\text{Flow}_{\text{per, } z_{\text{wiper}}} = \text{Flow}_{\text{per, } z_{\text{wiper}}} + (1 - \text{Lf}) \text{ WallFlow}_{z_{\text{wiper}}} \tag{2.16}
\]

The remaining wall flow is calculated with:

\[
\text{WallFlow}_{z_{\text{per}}} = \text{Lf WallFlow}_{z_{\text{per}}} \tag{2.17}
\]

Now we know how much liquid is distributed over the different cell flows we can calculate the liquid distribution in the next layer of cells. For the transfer of liquid from layer \( z \) to layer \( z+1 \) we distinguish four different situations.

Cells that represent the front side of packing sheet with a channel orientation downwards to the right.

\[
\text{Flow}_{x, z+1} = \text{CellFlow}_{x+1, z} + \text{CellFlow}_{x, z} + \text{CellFlow}_{x+1, z+1} + \text{CellFlow}_{x+1, z+1} \tag{2.18}
\]
Cells that represent the back side of a packing sheet with a channel orientation downward to the right.

\[(2.19)\]

\[
\text{Flow}_{x,y,z+1} = \text{CellFlow}_{a \ x-1,y,z} + \text{CellFlow}_{b \ x,z} + \text{CellFlow}_{c \ x+1,y,z} + \text{CellFlow}_{d \ x+1,y-1,z}
\]

Cells that represent the front side of a packing sheet with a channel orientation downwards to the left.

\[(2.20)\]

\[
\text{Flow}_{x,y,z+1} = \text{CellFlow}_{a \ x+1,y,z} + \text{CellFlow}_{b \ x,z} + \text{CellFlow}_{c \ x-1,y,z} + \text{CellFlow}_{d \ x-1,y+1,z}
\]

Cells that represent the back side of a packing sheet with channel orientation downward to the left.

\[(2.21)\]

\[
\text{Flow}_{x,y,z+1} = \text{CellFlow}_{a \ x+1,y,z} + \text{CellFlow}_{b \ x,z} - \text{CellFlow}_{c \ x-1,y,z} + \text{CellFlow}_{d \ x-1,y-1,z}
\]

The liquid on the wall flows straight downwards.

The liquid distribution is now calculated from layer to layer till the bottom layer is reached.

To establish the values of the different maximum and minimum critical flows and the splitting factors we have to fit the calculated distribution profiles on the measured data. In practice not all of the observed flow paths appear. For the bulk cell of RALU-PAK for instance we found that only the cell flow in the channel direction (a) and the cell flow through the perforation (b) are important. This means that we only have to fit three different parameters; one splitting factor and the minimum and maximum critical flow for the cell flow through the perforation. For MONTZ-PAK we found that all cell flows except the cell flow through the perforations (b) are important. This means that theoretically we have to fit six parameters. The only limitation we found was a minimum critical value for the cell flow via the crossings (d). This leaves for MONTZ-PAK three parameters to fit on the measured distribution profiles; two splitting factors and one minimum flow rate below which rivulet flow appears.
2.2.4 COMPARISON BETWEEN MEASURED AND CALCULATED DISTRIBUTION PROFILES

As mentioned in Section 2.2.3 we had to fit the splitting factors and capacities on the measured liquid distribution profiles. This was done for both packings and here we will compare and discuss the measured and calculated distribution profiles.

![Experimental RALU-PAK and Calculated RALU-PAK](image)

Fig. 2.7 Measured (left) and calculated (right) point source distribution profiles of RALU-PAK for water at different liquid loads

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Point source distribution profiles

Fig.2.7 shows the measured and calculated distribution profiles at three liquid loads for two sheets of RALU-PAK. Here as well as in the other spreading experiments discussed later on, the liquid was supplied on the sheet with a right downwards orientation only. From the measured data (left) it is obvious that there is no liquid distribution to the left side, which indicates that the liquid remains on the sheet with a right downwards orientation on which it was introduced. From this observation we concluded earlier (Section 2.2.3) that flow path 3 (flow over the channel ridge) and 4 (flow via the crossings) are not relevant for RALU-PAK. For very low liquid loads the liquid is mainly collected in compartment 13. This compartment collects the liquid of the channel which was fed by the point source from which we may conclude that the liquid is flowing in the channel direction. This is caused by a rivulet flow through the inner part of the corrugation ridge. With an increasing liquid load we have a decrease in the lateral spreading which is a

<table>
<thead>
<tr>
<th></th>
<th>RALU-PAK</th>
<th>RALU-PAK</th>
<th>MONTZ-PAK</th>
<th>MONTZ-PAK</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>water</td>
<td>water + 6% ethanol</td>
<td>water</td>
<td>water + 6% ethanol</td>
</tr>
<tr>
<td>$S_l_b$</td>
<td>0.95</td>
<td>0.95</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>$S_l_c$</td>
<td>0</td>
<td>0</td>
<td>0.65</td>
<td>0.65</td>
</tr>
<tr>
<td>$S_l_d$</td>
<td>0</td>
<td>0</td>
<td>0.15</td>
<td>0.15</td>
</tr>
<tr>
<td>$Flow_{b_{\text{min}}}$ [l/h]</td>
<td>2.7</td>
<td>1</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>$Flow_{c_{\text{min}}}$ [l/h]</td>
<td>0</td>
<td>0</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>$Flow_{d_{\text{min}}}$ [l/h]</td>
<td>0</td>
<td>0</td>
<td>1</td>
<td>0.6</td>
</tr>
<tr>
<td>$Flow_{b_{\text{max}}}$ [l/h]</td>
<td>32</td>
<td>32</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>$R_f$</td>
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<td>0.05</td>
<td>0</td>
<td>0</td>
</tr>
<tr>
<td>$PerFlow_{\text{max}}$ [l/h]</td>
<td>24</td>
<td>24</td>
<td>17</td>
<td>17</td>
</tr>
</tbody>
</table>

| Table 2.3 Liquid spreading parameters |
consequence of the fact that a part of the liquid is flowing through the slits under a smaller angle with the vertical axes. This change of channel is repeated several times, resulting in a maximum load nearby the centre line. However if we further increase the liquid load the spreading increases again due to the fact that the slits cannot handle all the liquid. The surplus of liquid is flowing then in the channel direction. The calculated profile fits the measured profile fairly well over the entire

**Fig.2.8** Measured (left) and calculated (right) point source distribution profiles of MONTZ-PAK for water at different liquid loads
CHAPTER 2

flow range, if we use the splitting factors and minimum and maximum capacity rates given in Table 2.3.

Fig.2.8 shows the results obtained with MONTZ-PAK. From the measured profile it is clear that the liquid flow behaviour is quite different in comparison with RALU-PAK. The liquid, leaving the packing at the left side of the point source, indicates that there is a liquid transport from the sheet with a right downwards orientation to the sheet with a left downwards orientation. This liquid transport can only occur via the crossings (the place where two adjacent corrugations are in contact with each other), as discussed in Section 2.2.3. The spreading behaviour is rather constant with the liquid load. Only at very low liquid loads we have again rivulet flow. The small rivulets do not reach crossings, which means that more liquid follows the channel direction. This results in larger spreading. If we fit the splitting factors and the minimum flows on the measured results we see again that there is a good agreement between the measured and calculated results for the entire flow range. The obtained splitting factors and minimum and maximum flows are given in Table 2.3.

Fig.2.9 Measured (left) and calculated (right) wall zone point source distribution profiles of RALU-PAK for water at different liquid loads (left side packing periphery) above compartment 5
Liquid point source distribution in the periphery zone

To calculate the flow in the periphery zone we have to fit the periphery flow parameters as maximum periphery flow and return factor on the measured periphery flow data. The experiments were carried out by introducing the liquid point source at about 6 cm from the periphery of the packing. For this reason we moved the packing sheets to the right and held the point source on the same place. The liquid was introduced on the sheet with a channel orientation towards the wall (in this case a left downward orientation).

Fig.2.9 shows the measured and calculated results for RALU-PAK. For RALU-PAK it is clear that the liquid which is transported towards the packing periphery accumulates in the periphery region.

For a similar test with MONTZ-PAK (Fig.2.10) we see that the results are totally different. The main reason for this difference is again the flow path via crossings of

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*Fig.2.10* Measured (left) and calculated (right) wall zone point source distribution profiles of MONTZ-PAK for water at different liquid loads (left side packing periphery above compartment 5)
the corrugation ridges. A part of the liquid never reaches the packing periphery because it takes a crossing and is transported to the sheet with an orientation downwards from the wall. The liquid that reaches the wall has also the possibility to change sheet via a crossing which means that it can flow back into the packing. The increase of wall flow with an increasing liquid load is a consequence of the fact that if the local liquid load on the periphery of the packing exceeds the maximum periphery flow the liquid will fall from the packing. In real columns the liquid that leaves the packing will cause wall flow and can only be brought back on the packing by the wall wipers. To investigate whether the crossings in the periphery zone are important, we arranged the packing sheets in two different ways as shown in Fig.2.11. The difference is that in Fig.2.11a we have a number of crossings directly on the periphery; this is not the case in Fig.2.11b. As expected we found for the situation in Fig.2.11b a larger liquid accumulation in the periphery zone. For reason of simplification we have averaged the data of both situations to fit the parameters (PerFlow and Rf). The resulting average profile is shown in Fig.2.9a. Table 2.3 shows the values of the used maximum periphery flow and return factor for both RALU-PAK and MONTZ-PAK.

![Diagram](image)

**Fig.2.11** Two possible situations at the periphery a) crossings (the place where two adjacent corrugation are in contact) are located exactly at the packing periphery b) crossings are located a little bit from the packing periphery inside the packing

**Influence of the liquid properties**

Because of the fact that the liquid properties can influence the flow behaviour of the liquid over the packing, we looked at the influence of surface tension and viscosity.

Most of the distillation systems have a considerably lower surface tension than water. To lower the surface tension of water with ± 30% we added some ethanol (6%) to the water (for properties see Table 2.4). The results for RALU-PAK carried out with this water-ethanol mixture for the same test situation as mentioned earlier are given at the left side of Fig.2.12. The right side shows the model calculation.
Fig. 2.12 Measured (left) and calculated (right) point source distribution profiles of RALU-PAK for a water-6%ethanol mixture at different liquid loads.

<table>
<thead>
<tr>
<th></th>
<th>water</th>
<th>6% ethanol (by vol.)</th>
<th>37% glycol (by vol.)</th>
<th>88% glycol (by vol.)</th>
</tr>
</thead>
<tbody>
<tr>
<td>density [kg/m³]</td>
<td>1000</td>
<td>1000</td>
<td>1050</td>
<td>1100</td>
</tr>
<tr>
<td>surface tension [N/m]</td>
<td>0.072</td>
<td>0.052</td>
<td>0.045</td>
<td>0.044</td>
</tr>
<tr>
<td>viscosity [Pa.s]</td>
<td>0.001</td>
<td>0.001</td>
<td>0.002</td>
<td>0.011</td>
</tr>
</tbody>
</table>

Table 2.4 Liquid properties
From the measurements it is clear that the larger lateral spreading (caused by rivulets that follow the channel direction) at low liquid loads disappear at lower surface tension. In the model calculation we only lowered the minimum flow below which rivulets occur to simulate this effect.

Fig. 2.13 Measured (left) and calculated (right) point source distribution profiles of MONTZ-PAK for a water-6% methanol mixture at different liquid loads
If we look at the results for MONTZ-PAK shown in Fig.2.13 we may come to the same conclusion. The model calculations were carried out with a lower minimum flow below which rivulet flow becomes dominant. The used splitting factors and maximum flows are given in Table 2.3.

In general we may say that rivulet flow at low liquid loads will disappear at lower surface tensions. This means that for a liquid with a low surface tension the spreading behaviour of the packing will be less dependent on the liquid load as long as the local liquid load is kept below the maximum flow.

We also studied the influence of the viscosity by using water-glycol mixtures. Table 2.4 gives the properties of the used water-glycol mixtures. Fig.2.14 and 2.15 give the measured point source distribution curves for RALU-PAK for a mixture of 37% glycol and 88% glycol respectively. From the measurements with the 37% glycol mixture we see that there is no principal difference compared to the measurements obtained with water. We only find a lower critical value below which rivulet flow starts, which can be explained from the lower surface tension of the

![Fig.2.14 Measured point source distribution profiles of RALU-PAK for a water-37%glycol mixture at different liquid loads](image1)

![Fig.2.15 Measured point source distribution profiles of RALU-PAK for a water-88%glycol mixture at different liquid loads](image2)
glycol mixtures. At high liquid loads we see that the lateral transport will not increase as we found for water. This can be explained from the fact that the surplus of liquid that can not flow through the slits is now flowing over the channel ridge in stead of in the channel direction. The liquid that flows over the channel ridge may contact a crossing which gives the opportunity to flow to the neighbouring sheet with an opposite orientation. This is the explanation for the appearance of liquid on the left hand side of the point source at high liquid loads (Fig.2.15b).

The results for MONTZ-PAK are given in Fig.2.16 and 2.17. From these figures we see that for MONTZ-PAK the spreading behaviour is almost independent of the liquid load, which means a further reduction of the minimum limit below rivulet flow occur.

Our observations indicated that the influence of the liquid properties on the periphery capacity and the return factor is negligible.

**Fig.2.16** Measured point source distribution profiles of MONTZ-PAK for a water-37%glycol mixture at different liquid loads

**Fig.2.17** Measured point source distribution profiles of MONTZ-PAK for a water-88%glycol mixture at different liquid loads
Fig. 2.18 Measured liquid distribution of 5 equidistantly (6.4 cm) placed point sources (left) and the corresponding tracer distribution (right) of the central point (above compartment 7) source of RALU-PAK at different liquid loads.

Tracer test with water

Up to now we only studied the spreading of liquid point sources. The question is whether the mixing mechanisms between different liquid streams will be the same as the spreading mechanisms found from our point source distributions. For this reason we also carried out some tracer experiments. The tracer we used was a NaCl solution and the tracer concentration was measured by conductivity probe. The liquid was introduced by five pour points which delivered liquid to both adjacent sheets. The central point source was fed with the water salt solution instead of demineralized water. The liquid quantity and the salt concentration was measured for each collecting compartment. The measured results for RALU-PAK are presented in Fig. 2.18. On the left side we see the liquid distribution for all the point sources and on the right side we see the tracer concentration distribution from the central point source. It was impossible to supply both adjacent sheets with exactly half of the liquid delivered by point sources, which explains the asymmetrical shape of the distribution curves. If we compare the results in Fig. 2.18 with the calculated results.
obtained with the parameters found from the water point source tests (Fig.2.19) we see that there is a good agreement between measured and simulated tracer distributions. The same conclusion can be made if we compare the measured results in Fig.2.20 and the calculated results in Fig.2.21 for MONTZ-PAK.

Because for water the spreading mechanism depends on the liquid load, this local different loads can result in differences between tracer and liquid point source distributions. Since there is a good agreement between the measured and simulated tracer distributions, although the parameters for the simulation were obtained from liquid point source tests, we may conclude in general that for the liquid loads of interest the distribution and mixing mechanisms are the same.

2.2.5 CONCLUDING REMARKS

For RALU-PAK the liquid will stay on the sheet on which it is introduced and is transported in the same direction as the sheet orientation which results in a large lateral transport. Due to the open surface, the liquid will continuously change side of the packing sheet. For MONTZ-PAK there is a exchange of liquid between
Fig. 2.20 Measured liquid distribution of 5 equidistantly (6.4 cm) placed point sources (left) and the corresponding tracer distribution (right) of the central point (above compartment 7) source of MONTZ-PAK at different liquid loads.

adjacent sheets. Due to the closed surface the liquid film is forced to flow over the channel ridge where it is split at the crossings on which the adjacent sheets contact each other and a part of the liquid is transported to adjacent sheet. This continuous splitting on crossings will reduce the lateral transport but improve the liquid distribution over adjacent sheets.

In general the liquid properties have no influence on the principal difference in
Fig. 2.21 Simulated liquid distribution of 5 equidistantly (6.4 cm) placed point sources (left) and the corresponding tracer distribution (right) of the central point (above compartment 7) source of MONTZ-PAK at different liquid loads

the flow distribution between RALU-PAK and MONTZ-PAK. Lower surface tension will reduce the critical liquid load below which rivulet flow appears. This means that the lateral transport at low liquid loads is less and the spreading behaviour is more constant with liquid load. For RALU-PAK at high viscosity and high liquid loads the surplus of liquid that can not flow through the slits will flow over the channel ridge in stead of the channel direction. Allied with this change of flow path, the flow path via the crossings will become relevant.
2.3 GAS DISTRIBUTION IN CORRUGATED SHEET PACKING

Up to now little attention is paid to the gas distribution in packed columns, especially columns packed with structured packings. Most research is focused at the liquid distribution and the gas distribution is assumed to be uniform. However, if we look more closely at a small scale the assumption of uniform gas distribution seems to be a rough approximation for instance due to wall effects [4,5]. That is why we tried to get a better understanding of how the gas flows through the packing, which mechanisms are responsible for the gas distribution and how spreading and mixing mechanisms effect each other. Tracer spreading is paid more attention to in literature. The first fundamental approach was presented by Zogg [1,2] who investigated the tracer spreading behaviour between two packing sheets with dye injections. Quantitative tracer spreading results in a 0.45 m ID column with MELLAPAK 250Y were first presented by Meier et al. [8]. They used CO₂ as tracer gas. Stikkelman et al. [5,6] presented some tracer spreading results measured for comparable conditions as Meier, but also for different types of structured packings. He concluded that gas spreading profiles, comparable to the results of Meier are found for MELLAPAK 500Y but, for the other types of packing, a different spreading mechanism can be present. So there are still some questions about gas spreading behaviour which ask for a more fundamental study.

In the first part of this section a model will be presented that can predict the gas flow pattern between 2 and 4 sheets. The measured and simulated profiles will be compared. In the second part we will present some tracer spreading tests obtained between 2 and 4 sheets and will discuss the results.

2.3.1 EXPERIMENTAL SET UP

To investigate the gas distribution behaviour of structured packings we investigated the two principal different packing types. One with a totally closed surface (MONTZ-PAK B1 250) and one with an open surface (RALU-PAK 250 YC). The gas distribution was investigated in a simple piece of equipment shown in Fig.2.22. The main part of the equipment is the open frame where we can put one or more packing sheets with a length of 0.5 m and a height of 0.2 m. At the bottom we have 17 compartments which are fed separately. The two outside compartments are considered as left hand and right hand side wall compartments, respectively. The outlets of these two compartments are connected with the wall channels, while the 15 inlet compartments which fit exactly with the width of a packing element (0.5 m) feed actually 17 outlet packing channels. We measured both the inlet flow rate and the pressure for each compartment. Directly above the packing we measured the concentration of tracer gas (CO₂) and/or the outlet velocities of each air stream leaving a channel. For the latter purpose a specially designed micro-Pitot tube was used. Thanks to its construction this tube enabled also the determination of the direction and the static pressure of the outlet air flows.
For the tests of MONTZ-PAK, the packing with the closed surface, we used two packing sheets to create the characteristic structure of crossing channels. The air was simply introduced between the two sheets. To investigate the flow behaviour we did some tests with a configuration in which the packing was fitted perfectly so there was no open space between the wall and the end of the packing channels. With this configuration we investigated the mixing in the packing by introducing a homogeneous inlet profile to the bottom compartments and a local tracer injection. The influence of the initial maldistribution was investigated by introducing point sources at different locations. To investigate in more detail the influence of a wall channel and wall wipers we did some tests with an open space between the outside boundary of the packing and the wall and put a restriction in the wall channel to
simulate a wall wiper. The three different investigated wall configurations are given in Fig.2.23.

To create a characteristic structure for RALU-PAK, the open surface packing, we had to use at least four plates to enable communication between the two sides of the packing sheet. The outside face of the two outside sheets was closed to avoid air escape to the surrounding. Because of the gas inlet location, some initial gas maldistribution could not be avoided during these experiments. With RALU-PAK, only the configuration with the perfectly fitted packing was investigated.

Fig.2.24 Structure of interconnected cells (left) and corresponding structure of corrugated sheet packing (right)
2.3.2 Gas flow distribution model

The gas distribution model is based on a cell model. The structure of the model is the same as that used for the liquid-distribution model. How the cells are interconnected to create the packing channels is shown in Fig.2.24. For the gas flow modelling the pressure distribution has to be incorporated, because the pressure drop is the driving force for the gas flow. Since the gas and pressure distribution are inherently dependent an exact description requires the simultaneous solution of the mass, momentum and energy balance for each crossing of flow channels. Because of the differences in geometry, a distinction has to be made between the bulk and wall zone of the packing.

Situation in the bulk zone

The geometry of the flow channels between two corrugated sheets is of a complex structure. Up to now nobody succeeded in simulating this flow pattern with a computational flow simulator as Fluent or Phoenix. That is why we had to solve the integral balance over the outlets of intersections of two crossing channels. The different crossings (cells) are linked by the principle of mass conservation. First a short description is given of the equations which describe the distribution in one crossing. A schematic presentation of the control volume over which the equations are solved is given in Fig.2.25. Numbers 1 and 2 denote inlets, and 3 and 4 denote outlets of a crossing of gas flow channels. The shaded area is the area where two flows contact each other.

In what follows a short description is given of the mathematical model describing the gas flow distribution over two crossing channels. In our model we assume a constant gas density ($\rho_g$) and a uniform velocity ($u$) over the channel cross section area ($A$). With these assumptions we can find for a steady-state one dimensional gas flow the next four balance equations.

![Fig.2.25 Control volume for a cross point in the bulk of the packing (left) and the corresponding packing channels (right)](image)

40
MASS BALANCE

$$A \rho_s (u_1 + u_2) = A \rho_s (u_3 + u_4) \quad (2.22)$$

MOMENTUM BALANCE

y-direction

$$F_y + A \sin(\alpha) (P_1 - P_3 - P_2 + P_4) = A \sin(\alpha) \rho_s (u_3^2 - u_1^2 + u_2^2) \quad (2.23)$$

z-direction

$$F_z + A \cos(\alpha) (P_1 - P_3 + P_2 - P_4) = A \cos(\alpha) \rho_s (u_3^2 - u_1^2 + u_2^2) \quad (2.24)$$

ENERGY BALANCE

$$A \frac{1}{2} \rho_s (u_1^2 + u_2^2 - u_3^2 - u_4^2) + A (u_1 P_1 - u_2 P_2 - u_3 P_3 - u_4 P_4) = E_{loss} \quad (2.25)$$

where $F$ denotes acting force, $P$ the absolute pressure and $\alpha$ the channel inclination.

The loss of mechanical-energy per time ($E_{loss}$), caused by frictional resistance, can be expressed as:

$$E_{loss} = A \frac{1}{2} \rho_s \sum_{i=1}^{4} u_i^2 \quad (2.26)$$

where $\xi$ is a overall pressure loss factor.

Since above five equations have seven unknowns $v_3, v_4, P_1, P_2, E_{loss}, F_y$ and $F_z$, two additional equations are needed to completely describe the problem. The y- and z-directed friction forces $F_y$ and $F_z$ may be expressed as follows:

$$F_y = \sin(\alpha) (-F_1 + F_2 - F_3 + F_4) \quad (2.27)$$
\[ F_z = \cos(\alpha)(-F_1-F_2-F_3-F_4) \]  \hspace{1cm} (2.28)

With:

\[ F_i = A \frac{1}{\mu_i^2} \]  \hspace{1cm} i=1..4 \hspace{1cm} (2.29)

\( F_y \) and \( F_z \) can now be written as a function of \( u_i \) (i=1..4).

By substituting Eq.(2.26) to (2.29) into Eq.(2.22) to (2.24) and some manipulations we can write all the unknowns as a function of \( u_i \). By substitution of \( u_i \) in the energy balance Eq.(2.25) we get a polynomial of the third order. The solution of this polynomial may result in more than one real root. The value for \( u_i \) which gives the lowest energy losses over a crossing is chosen as solution.

**Situation in the wall zone**

The flow pattern in the wall zone is even more complex than the bulk zone because of the bend losses and channel cross-section changes.

We can distinguish two situations in the wall zone.

1) The situation that the packing fits perfectly which means that there is no wall channel between the periphery of the packing and the column wall. This can be simulated by assuming total reflection.

2) In a more realistic situation that there is a wall channel between the packing periphery and the column wall. This can be simulated by partial reflection.

![Fig.2.26](control_volume.png)  
*Fig.2.26* Control volume for a cross point at the wall (left) and the corresponding packing and wall channel(s) (right)
We will discuss now the situation of partial reflection. Fig.2.26 illustrates schematically the situation without a wall wiper, represented in form of two packing channels, one (1) reaching and one (4) leaving the wall channel. The vertical channels with the numbers 2 and 3 are a consequence of the wall channel between the packing and the column wall. For the wall zone we can find the same balance equation as the bulk zone. Only the momentum balance equations differ somewhat.

**MOMENTUM BALANCE**

**y-direction**

\[ F_y + \frac{A}{\sin(\alpha)} (P_1 + P_4) = A \sin(\alpha) \rho_g (-u_1^2 - u_4^2) \]  

(2.30)

**z-direction**

\[ F_z + A_{\text{wall}} (-P_1 + P_2) = A_{\text{wall}} \rho_g (u_2^2 - u_3^2) + A \cos(\alpha) \rho_g (u_4^2 - u_1^2) \]  

(2.31)

The friction forces are calculated with Eq.(2.29) where we have to replace \( \zeta \) by \( \zeta_{\text{bend}} \) for channel inlet 1 and outlet 4 and by \( \zeta_{\text{wall}} \) for channel inlet 2 and outlet 3 to take into account the different pressure loss for the bend and wall channel respectively. \( F_z \) is calculated in the same way as a crossing in the packing.

\[ F_z = \cos(\alpha)(-F_1 - F_4) + (-F_3 - F_2) \]  

(2.32)

For \( F_y \) we have besides the friction forces a wall force \( (F_{\text{wall}}) \)

\[ F_y = \sin(\alpha)(-F_1 + F_4) - F_{\text{wall}} \]  

(2.33)

The additional wall force is calculated from the average pressure in the wall channel and the wall area of the control volume:

\[ F_{\text{wall}} = (P_2 + P_3) \frac{A}{\sin(\alpha)} \]  

(2.34)

These equations can be treated in the same way as described for the bulk flow.
A wall wiper can now be simulated by a local crossing section reduction in the wall channel shown in Fig.2.27. From the mass balance we can calculate the velocity change at the wall channel cross section reduction and from the momentum balance we can calculate the pressure change at the wall channel cross section reduction. The balances are determined over the control volume 1 (or 3) in Fig.2.27.

\[
u_{\text{wiper}} = \frac{A_{\text{wall}}}{A_{\text{wiper}}} u_{\text{wall}}
\]

\[
P_{\text{wall}} - P_{\text{wiper}} = \rho g u_{\text{wall}} \left( \frac{A_{\text{wall}}}{A_{\text{wiper}}} - 1 \right)
\]

The reduced wall channel cross section is taken into account in the calculation over the wall crossing (compare control volume 2 in Fig.2.27 with the control volume in Fig.2.26).

**Gas inlet zone**

The channel crossings in the bottom layer are somewhat different from the crossings in the bulk or wall zone. These crossings can be considered as an inlet channel with two outlets (Y-junction). This means that two neighbouring channels in adjacent sheets are fed by the same inlet compartment (Fig.2.28), thus they have equal inlet pressures. Since the inlet flow has no horizontal component in y direction, the horizontal momentum of the inlet channel is assumed to have a zero value.
The corresponding balance equation is:

**MOMENTUM BALANCE**

**y direction**

\[ F_y = A \sin(\alpha)(-P_y + P_i) = A \sin(\alpha) \rho \frac{A^2}{L} (u_y^2 - u_i^2) \]  

(2.37)

\( F_y \) is calculated from the friction forces

\[ F_y' = \sin(\alpha)(-F_y + F_i) \]  

(2.38)

\( F_i \) and \( F_i' \) are calculated with Eq.(2.29) where we have to replace \( \zeta \) by \( \zeta_{\text{inlet}} \) to take the extra pressure loss due to inlet losses into account. Eq.(2.37) together with the mass balance enables the calculation of the outlet, the velocities \( u_y \) and \( u_i' \) from the pressures \( P_y \) and \( P_i' \) and the inlet flow. Knowing the outlet velocities we can calculate the pressures at the inlet from the vertical momentum balance (Eq.2.24) and the assumption that the pressures at the inlet are the same (\( P_i = P_i' \)).

*The calculation procedure*

The flow field in the packing is now solved by successive substitutions of the crossing outlet pressures (\( P_i \) and \( P_i' \)) beginning from the bottom layer of cells, going
Fig. 2.29 Iteration scheme for calculation of the gas velocity field

upwards, layer by layer, until the top layer of the packing element is reached. The crossing outlet pressures at the top are known and taken as reference pressures. The inlet flows for the crossing in the bottom cell layer are also known and have to be given as input value. The iteration schema is shown in Fig. 2.29. Convergence is only reached if enough damping is allowed for the calculation of the new values for both the pressures and the velocities, which means that the calculation procedure is rather time consuming.
Fig.2.30 Measured outlet velocities in left and right upward orientation (a) and inlet pressures in bottom compartments (b) for two sheets of MONTZ-PAK without wall channel ($u_p=2.8$ m/s)

Fig.2.31 Simulated outlet velocities in left and right upward orientation (a) and inlet pressures for the packing channels (b) for two sheets of MONTZ-PAK without wall channel ($u_p=2.8$ m/s)

2.3.3 RESULTS AND COMPARISONS FOR THE GAS DISTRIBUTION

In this section we first discuss the results measured between two sheets of MONTZ-PAK B1 250 and compare them with the simulations.

Fig.2.30a and b show the velocity distribution and pressure distribution measured with MONTZ-PAK in the case of a perfectly fitted packing obtained with a uniform initial gas profile. The two velocity profiles denoted "right" and "left" refer to the channel orientation. From Fig.2.30a we may conclude that the outlet velocity distribution is almost uniform. There is only a small disturbance near the wall. If we look at the pressure distribution at the bottom we see that the pressure measured in the compartments which feed a channel that ends at the wall is a little bit higher than in the channels which are connected directly with the outlet. This is due to the additional pressure drop caused by the 90 degrees turn in the gas flow [1]. The measured (static) pressure distribution at the top of the packing was flat for all measurements. The measured outlet direction of the gas flow confirms that the gas is
flowing in the channel direction (45 degrees with the vertical axes). Fig. 2.31a and b show the simulated velocity distribution at the top and the simulated pressure distribution at the bottom respectively. The simulations were carried out with a cell pressure loss coefficient of 0.32 ($2\xi$) and an inlet loss coefficient of 1.0 ($\zeta_{inlet}$). With 13 crossings in series for one packing layer (height = 0.2 m) this corresponds with a overall friction factor for this packing of 1.1 which is equal to the Overall friction factor found for a similar type of structured packing by Zogg [1]. For the cells at the wall a pressure loss coefficient ($\zeta_{brea}$) two times the pressure loss factor for the bulk was assumed to take the extra loss due to the turn into account.

The same experiment has been repeated with a spacing of 0.01 m between packing and the wall. This open space creates a wall channel with a cross section of two times the cross section of the packing channels. The packing was fed by a uniform initial gas profile. The compartments which are situated below the wall channels were closed to avoid direct feeding of the wall channels. The velocity profile measured at the top and the pressure profile measured at the bottom are shown in respectively Fig. 2.32a and b. From the velocity profile it is clear that only a part of the gas that reaches the wall flows back into the packing. The low velocity

![Graphs showing velocity and pressure distributions](image-url)
measured for channel 6 indicates that this is more pronounced for channels starting at the bottom of the wall channel than for other channels. The low pressure at the bottom of the wall channel is a consequence of the absence of entrance effects (no inflow in the wall compartment). The low pressure in the bottom of the wall channel causes a positive pressure gradient towards the wall which results in a gas transport in the wall direction. If we compare the measured profiles with the simulated profiles for the velocity distribution at the top (Fig.2.33a) and the pressure distribution in the bottom (Fig.2.33b) respectively, we see that there is a reasonable agreement if we use the same pressure loss factor for the wall channel as for the packing ($\zeta_{wall} = \zeta$).

In order to quantify the interaction of crossing gas flows at the interface within a packing element we inserted a flat plate between the two corrugated sheets. In this way channels with a pure triangular cross section channel were created. So the outcoming flow profile is a consequence from the inflow and the return flow at the wall only.

The resulting flow profile shown in Fig.2.34a is obviously similar to that

---

*Fig.2.34* Measured outlet velocities in left and right upward orientation (a) and inlet pressures in bottom compartments (b) for two sheets of MONTZ-PAK with wall channel and a plate between the adjacent sheets ($u_p=2.7$ m/s)

*Fig.2.35* Simulated outlet velocities in left and right upward orientation (a) and inlet pressures for the packing channels (b) for two sheets of MONTZ-PAK with wall channel and a plate between the adjacent sheets ($u_p=2.4$ m/s)
obtained with an open channel structure (Fig. 2.31a). The comparison of the measured inlet pressure profiles (Fig. 2.34b vs. Fig. 2.32b) indicates a considerably lower value of the pressure drop in the case of closed channel flow, i.e. with a plate inserted between the corrugated sheets. With respect to the normal situation this means approximately a 50% lower value of the pressure loss factor. The explanation can be found in the fact that the insertion of a plate eliminated the influence of rather rough corrugation ridges. A lower pressure drop factor means a decrease in the amount of wall flow, because of the increase of driving force in the channels transporting gas from the wall. The corresponding simulation results are shown in Fig. 2.35.

To avoid excessive wall flows structured packing is usually provided with wall wipers. The influence of a wall wiper was investigated by inserting a restriction in the wall channel at a distance of 8 cm from the top, which reduced the wall channel cross section locally to 10%. Fig. 2.36a and b show respectively the velocity profile measured at the top of the packing and pressure profile measured at the bottom of the packing. Comparison with the velocity profile obtained without the wall wiper

Fig. 2.36 Measured outlet velocities in left and right upward orientation (a) and inlet pressures in bottom compartments (b) for two sheets of MONTZ-PAK with a restriction (wiper) in the wall channel ($u_p=2.4$ m/s)

Fig. 2.37 Simulated outlet velocities in left and right upward orientation (a) and inlet pressures for the packing channels (b) for two sheets of MONTZ-PAK with a restriction (wiper) in the wall channel ($u_p=2.4$ m/s)
(Fig.2.36a vs. Fig.2.32a) suggest two main differences. Firstly, channel 3 and 14 both with an upward orientation towards the centre and inlets directly below the restriction, exhibit a local maximum. The pressure just below the restriction is that high that the gas is forced to flow back into the packing. Secondly, we see that the velocities measured in channel 3 to 14 are higher than is the case without the wall channel restriction. Obviously, the wall restriction causes an increased pressure in the whole section below the restriction. This means that a larger part of the gas flow that reaches the wall is forced to flow back into the packing. Above the wall wiper the pressure in the wall channel will reduce to the outlet pressure. Due to the lower pressure in the wall channel above the wiper the wall channel sucks the air out of the packing which results in very low velocities in channel 1 to 3 and 14 to 17 with an upward orientation towards the centre. This effect may be seen as an outlet effect. Simulation results shown in Fig.2.37 do not differ considerably from the measured ones. However, simulations were carried out with a local wall channel reduction of 75%. This was done to compensate for deviation in the height of the restriction used in the experiment and in the model. The latter one was equal to the height of a cell (Fig.2.27), and the real one was smaller.

**Fig.2.38** Measured outlet velocities in left and right upward orientation (a) and inlet pressures in bottom compartments (b) for two sheets of MONTZ-PAK without wall channel and a point source in the left side wall compartment

**Fig.2.39** Simulated outlet velocities in left and right upward orientation (a) and inlet pressures for the packing channels (b) for two sheets of MONTZ-PAK without wall channel and a point source in channels 1 (left and right)
To investigate the response of the packing on severe initial maldistribution we introduced a point source in the left wall compartment (compartment 1) and no flow to the other compartments. For these tests we used the configuration in which the packing fits perfectly to the wall so there was no wall channel. If we look at the velocity profile (Fig. 2.38a) it is clear that the channels 7 and 8 both with a right upward orientation, which are fed by the point source (the flow in channel 7 is reflected at the wall) exhibit peak velocities. An interesting point here is the gas velocity measured in the channels 8 to 17, which is obviously an attestation of a horizontal gas transport towards other channels right of channel 7. This is confirmed by the pressure profile (Fig. 2.38b) which shows that there is a positive pressure gradient towards the right hand side. Fig. 2.39 shows simulated velocity and corresponding pressure profile, for this situation. Because of the structure of the model a horizontal flow can only be created as combination of back- and forward flow (zig-zag flow). Although the simulation showed small fluctuations in the bottom velocities, the velocity profiles at the top were constant. If we compare the measurements with the simulations we may say that the peaks are predicted fairly well by the model.

![Fig. 2.40 Measured outlet velocities in left and right upward orientation (a) and inlet pressures in bottom compartments (b) for four sheets of RALU-PAK without wall channel (u_in=2.4 m/s)](image)

The model is generally not suitable for all types of structured packings, because it assumes channel flow and some packing types with an open surface structure do not have this channel flow. This is clear if we look at the data measured for RALU-PAK 250 YC measured for the situation that the packing fits perfectly at the wall and that we used a homogeneous initial distribution. Fig. 2.40a gives the velocity profile at the top. The difference between the velocities measured for the top sheets is probably due to unequal initial distribution, but in general we may consider this profile as flat. Fig. 2.40b shows the pressure distribution in the bottom which indicates that the pressure drop for RALU-PAK is a lower one than the pressure drop for MONTZ-PAK. The obtained pressure loss factor for RALU-PAK is almost 30% lower than the one for MONTZ-PAK. The explanation for this lower pressure drop can be found in the more vertical flow direction measured for the outlet flows. Due to the fact that the gas flows more vertical we have a lower actual gas velocity which results in a lower pressure drop.
2.3.4 MODELLING OF MASS TRANSFER BETWEEN TWO CROSSING CHANNELS

In the previous section it has been shown that, for closed surface packings, the gas generally follows the direction of the channels and that only at severe imposed maldistribution it can change channel. In the normal situation there will be no net flow between the channels but there is always some mixing between the channels due to turbulent diffusion. The mass transfer between crossing channels can be described with the basic mass transfer equation:

\[ N = k_{g \text{ ex}} a_{\text{ ex}} \left( c_{1,3 \text{ average}} - c_{2,4 \text{ average}} \right) \]  \hspace{1cm} (2.39)

where \( N \) is mass flow, \( k_{g \text{ ex}} \) mass transfer velocity, \( a_{\text{ ex}} \) is area of the exchange surface between two crossing channels, \( c_{1,3 \text{ average}} \) and \( c_{2,4 \text{ average}} \) are the average concentrations in channel 1,3 and 2,4 respectively. Fig.2.41 shows two crossing channels.

![Fig.2.41 Mixing (mass transfer) between two crossing packing channels](image)

On the other hand we can describe the mass transfer between two crossing flows in terms of splitting factors. In the case of a uniform gas distribution (\( u_1 = u_2 = u_3 = u_4 = u \)) and equal channel cross sections (A) we can write

\[ c_3 = (1-S_g) c_1 + S_g c_2 \]  \hspace{1cm} (2.40)

\[ c_4 = (1-S_g) c_2 + S_g c_1 \]  \hspace{1cm} (2.41)

where \( S_g \) is the gas splitting factor, \( c_1 \) and \( c_2 \) are concentrations at the channel inlets and \( c_3 \) and \( c_4 \) are concentrations at the channel outlets.
Equations (2.40) and (2.41) together with the mass balance over a crossing, give

\[ N = S_g u A (c_2 - c_1) \]  

(2.42)

The mass transfer described by the splitting factor should be the same as calculated from Eq.(2.39). We can derive therefor, the following relation between the splitting factor and the mass transfer coefficient for the case of a uniform gas distribution and equal channel cross sections

\[ S_g = \frac{k_{ge} a_{we}}{k_{ge} a_{we} + A u} \]  

(2.43)

So, the splitting factor can be expressed in terms of usual mass transfer parameters. For reasons of simplicity we will further use the splitting factor approach (with a constant splitting factor) to describe the mixing between crossing channels.

In the case that we have a perfectly fitting packing the gas flow, that reaches the wall, just turns and its concentration remains unchanged. In the case of an open space between the packing periphery and the wall the gas which reaches the wall will be partly reflected and partly bypass the packing. We assume that the degree of turbulence in the wall channel is that high that the flows are totally mixed. In the next section we will check this assumption.

2.3.5 Comparison between measured and calculated tracer distributions

The tracer spreading tests were carried out with the same equipment as described for the gas distribution test. The tracer (CO₂) was injected in the bottom of a certain channel and the outlet concentration for each channel was measured by an infrared gas analyzer. Two series of tests were carried out to investigate the bulk flow and the wall flow. The measured results and comparisons with the model calculation are discussed below.

Point source tracer spreading tests

Fig.2.42 shows the results for MONTZ-PAK with no wall channel and a (initial) uniform gas profile (Fig.2.30a). The tracer is injected in channel 16 with a right upward orientation. It is clear that most of the tracer gas is following the channel direction causing a peak at the outlet of channel 16 in which the tracer was
introduced. The presence of tracer in the other channels indicates mixing (mass transfer) between crossing channels. The calculated tracer distribution is shown in Fig.2.43. A good fit of the experimental tracer profiles was obtained with splitting factor of 0.15.

If we use Eq.(2.43) to calculate the mass transfer coefficient from the splitting factor we find that the mass transfer coefficient is about 5% of the gas velocity in the channels. This is typical the value for the fluctuating velocity in a turbulent flow [9]. From our experiments we know that if we do the same test with a much lower gas velocity there is hardly any change in the velocity profile but we see a difference in the spreading behaviour. Due to the lower gas velocity the turbulence intensity in the channels will decrease, especially the turbulence at the exchange surface. Due to the
lower turbulence intensity we have less mixing between channels which results in a larger lateral spreading. This becomes clear if we compare Fig. 2.44, which gives the tracer spreading for a low gas velocity of 0.5 m/s (Reynolds = 550), with Fig. 2.42. Above 1 m/s (Reynolds = 1100) the gas spreading was found independent of the Reynolds number as earlier concluded by Meier et al. [8] and Stikkelman [6]. The relatively low Reynolds number for the transition to turbulent flow is in agreement with Zogg [1].

\[ \text{Fig. 2.45 Measured tracer distribution in left and right upward oriented sheet for two sheets of MONTZ-PAK with a wall channel and a tracer injection in channel 16 with a right upward orientation} \]

\[ \text{Fig. 2.46 Calculated tracer distribution in left and right upward oriented sheet for two sheets of MONTZ-PAK with a wall channel and a tracer injection in channel 16 with a right upward orientation} \]

To see how the velocity profile can influence the mixing behaviour we did a similar test for the situation with a wall channel between the packing periphery and the wall. The tracer profile is shown in Fig. 2.45, the corresponding velocity profile in Fig. 2.32a. The tracer was again injected in channel 16 with a right upward orientation. We see that there is no principal difference in mixing behaviour in comparison to the situation of a perfectly fitted packing (no wall channel). This is also predicted by the simulation in Fig. 2.46. With this configuration we investigated how the tracer profile develops over the height. For this purpose we placed a plate between the two sheets as earlier described in Section 2.3.2. There we found already that the plate between the packing sheets has hardly any influence on the velocity profile at the top. However the plate prevents contact between crossing gas flows. If we lift the plate mixing is only possible on crossings in the lower part of the packing sheets below the plate. From the outlet tracer concentrations we can calculate the tracer distribution at the location of the plate periphery. In this way we can obtain tracer distributions at different heights. The results are given in Fig. 2.47a to f. If we compare the profiles in Fig. 2.47 with the calculated profiles in Fig. 2.48 we see that there is more mixing between crossing channels in the bottom part of the packing. The more intensive mixing in the lower part is a consequence of the extra turbulence due to inlet effects.
If we look at the influence of a counter current liquid flow of 6 mm/s as shown in Fig.2.49 we may conclude that there is hardly any effect on the mixing behaviour of MONTZ-PAK in the normal operation range.
Fig. 2.49 Measured tracer distribution in left and right upward oriented sheet with a counter current liquid flow for two sheets of MONTZ-PAK without a wall channel and a tracer injection in channel 16 with a right upward orientation ($u_p=2.4$ m/s, $u_a=6$ mm/s)

We also investigated the mixing behaviour for RALU-PAK. In Section 2.3.3 we conclude that the velocity profile at the top is in principle not different from MONTZ-PAK but that the direction of the outlet gas velocity was more vertical than we should expect if we assume channel flow and that the pressure drop coefficient was much lower. From smoke tests, we found that the gas flows through the slits in the packing surface as shown in Fig. 2.50. The tracer distribution in Fig. 2.51, measured under same condition as the velocity profile in Fig. 2.40a, confirms the fact that the gas is not flowing in the channel direction because the peak found in the tracer distribution was found closer to the centre line. The tracer was introduced in channel 15 with a right upward orientation, however no tracer was found at the right side of channel 13.

With respect to the velocity effect on the gas spreading behaviour RALU-PAK showed behaviour similar to MONTZ-PAK.

Fig. 2.50 Gas flow ('slit flow') direction for RALU-PAK 250 YC
In practical applications the slits in RALU-PAK are more or less filled with liquid and gas is forced to flow in the channel direction. This means that the gas flow in irrigated RALU-PAK is similar to that of a closed surface packing. This is shown in Fig.2.52 where we introduced a counter current liquid flow of 3 mm/s. If we compare this figure with Fig.2.51, which was measured under the same conditions without a counter current water flow, it is clear that the lateral transport of gas (tracer) is increased. Apparently sufficient irrigated RALU-PAK behaves, in respect to the gas flow, more like a packing with a totally closed surface.
Point source tracer test in the wall zone

To check the assumption of total mixing for the crossings in the wall zone we introduced tracer approximately 3 cm from the wall in a channel with an upward orientation toward the wall. The measured and calculated results corresponding to the velocity profile shown in Fig.2.32a with a wall channel between the packing periphery are shown in Fig.2.53 and Fig.2.54 respectively. From the comparison between the measured and calculated profile we may conclude that the assumption of total mixing in the wall channel holds. The difference in the measured and simulated profile is a consequence of the extra turbulence in the channels that leave the wall due to inlet effects. The simulated profile was obtained with a splitting factor of 0.15 as we used for the bulk of the packing. A better fit can be obtained if we increase the splitting factor for the crossings nearby the wall however for reasons of simplicity we have kept the splitting factor constant for each crossing.

2.3.6 Concluding Remarks

For a uniform initial gas distribution and no disturbances in the packing the gas distribution profile will remain uniform. Flow differences between crossing channels will equalize only if sufficient large pressure gradients are present.

For normal conditions, i.e. the presence of a counter current liquid flow, for both MONTZ-PAK and RALU-PAK the prevailing flow direction is the channel direction. However, for RALU-PAK with a dry packing surface the gas flows through the slits.

The splitting factor of 0.15 calculated for superficial gas velocities above 1 m/s indicates that over the total loading range the channel flow prevails as lateral transport mechanism. This is in accordance with the observations of Stikkelman [6].

Above 1 m/s (Reynolds = 1100) the gas spreading is independent of the Reynolds number as earlier concluded by Meier et al [8] and Stikkelman [6]. At lower Reynolds numbers the tracer spreading increases due to a decrease of turbulence intensity. The relatively low Reynolds number for the transition to turbulent flow is in agreement with Zogg [1].

How these phenomena will influence the large scale spreading behaviour is discussed in Chapter 4.
NOTATION

SYMBOLS

α  Packing channel direction with the vertical axis  °
β  Base angle of a triangular packing channel  °
ϕ  Film flow direction with the vertical axis °
ϕ̂  Lateral spreading direction (gravity only) with the vertical axis °
ρ_l  Liquid density  kg/m³
ρ_g  Gas density  kg/m³
ζ  Cell pressure loss coefficient -
ζ_bend  Cell pressure loss coefficient due to 90 degree turn at wall -
ζ_inlet  Cell pressure loss coefficient due to inlet effects -
ζ_wall  Cell pressure loss coefficient for the wall channel -
A  Channel cross section area  m²
a_ex  Exchange area between crossing channels  m²
b  Width of a triangular packing channel  m
C  Tracer concentration  mol/m³
CellFlow  Liquid that follows a certain cell flow direction  l/h
df  Damping factor -
E_loss  Energy loss  Watt s
F  Force  N
Flow  Liquid flow in a bulk cell  l/h
Flow_in  Inlet gas mass flow  kg/s
g  Gravity acceleration  m/s²
h_τ  Height of a triangular packing channel  m
Lfi  Liquid wiper leakage factor -
N  Mole flow  mol/s
P  Pressure  Pa
Perflow  Liquid flow in a packing periphery cell  l/h
Rfi  Liquid periphery return factor -
S_g  Gas splitting factor -
S_l  Liquid splitting factor -
u  Gas velocity  m/s
WallFlow  Liquid flow in a wall cell  l/h
INDICES

a,b,c,d  Different cell flow outlets
min      Minimum possible flow
max      Maximum possible flow
x,y,z    Cell coordinates, c.q. flow direction
x_per    x cell coordinate of packing periphery cell
z_wiper  z cell coordinate where a wiper is located
1,2      Channel inlet numbers
3,4      Channel outlet numbers
wall     In wall channel
wiper    In wall channel at wiper location

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CHAPTER 3

MALDISTRIBUTION IN SMALL DIAMETER COLUMNS

3.1 INTRODUCTION

Most of the researchers [1-4] that investigated maldistribution in small packed distillation columns assume that the small scale irregularities in the flow are of minor importance, because the detrimental effect on separation performance is compensated by lateral mixing. However, from our sheet experiments (Chapter 2) we know that small differences in the surface texture and corrugation shape can lead to different lateral spreading and mixing behaviour for both gas and liquid phases. In this chapter we will consider the small scale (mal)distribution in a 0.5 ID column. The effect of small scale maldistribution on the separation performance of a column containing structured packing is discussed in Chapter 7.

In Section 3.2 we will concentrate on the small scale liquid distribution data for RALU-PAK and MONTZ-PAK measured by Stikkelman [5,6]. We did some additional measurements to see how these two packings with different lateral
spreading behaviour will react on severe initial maldistribution. The experimental results are compared with model calculations based on the theory described in Section 2.2.

The small scale gas distribution and lateral spreading in columns is examined in Section 3.3. The calculated tracer distributions, elaborated in Section 2.3, will be compared with experimental tracer distributions obtained by Meier et al. [2].

3.2 LIQUID MALDISTRIBUTION IN SMALL DIAMETER COLUMNS

3.2.1 EXPERIMENTAL SET-UP

To measure the small scale maldistribution we used the equipment, shown in Fig.3.1, developed for such purposes by Stikkelman. The equipment is described in detail in [5], so only a brief description is given here. The test column consists of 0.5 m ID perspex cylinders (height 1 meter) and may be packed up to a maximum height of 3 meter. The liquid is supplied to the packing by a (pressure type) liquid distributor, which is provided with 149 drip points (760 drip points per square meter). It can be reduced down to a point source distribution in an arbitrary manner. The main feature of the apparatus is the liquid collecting device, consisting of the support grid, divided into 332 compartments with a cross section of 2.5 x 2.5 cm² each. The liquid collecting compartments are connected to a corresponding number of U-tubes equipped with level sensors. In this way for each tube the time is measured to reach a certain liquid level.

![Simplified flow sheet of the small-scale maldistribution equipment, with enlarged details of liquid distributor and liquid collecting device (from [5])](image-url)
3.2.2 Parameters to Characterize the Distribution

To compare the model calculations and the experiments we need some characteristic parameters for a distribution profile. It is very difficult to put all the characteristics of a certain profile in one characteristic parameter. We follow the approach of Stikkelman who used two parameters:

- Maldistribution factor
- Wall flow factor

For both parameters we will give a short description.

Maldistribution factor

The Maldistribution factor (Mf) is the normalized standard variance of the collected flows, expressed as local superficial velocities \( u_i \) (based on 2.5 x 2.5 cm\(^2\) collecting compartments).

\[
Mf = \frac{1}{n} \sum_{i=1}^{n} \left( \frac{u_i - u_{\text{bulk}}}{u_{\text{bulk}}} \right)^2
\]

(3.1)

where \( n \) is the number of compartments and \( u_{\text{bulk}} \) the superficial liquid velocity based on the inner 90% of the column cross section.

If the maldistribution is zero, this means that all the flows are the same and that we have a totally flat distribution profile. On the other hand a large maldistribution factor means that there are considerable differences between local flows. However, the absolute value of the Maldistribution factor depends on the scale of detection. A practical disadvantage of this definition is the fact that it only takes into account the variance of the local velocities, and not their spatial distribution orientation. To see the actual distribution profile a three dimensional plot of velocity distribution is needed. This problem is worked out more in detail in [5].

As long as we have a rather good initial distribution and no local irregularities we know that the liquid distribution in the bulk of the packing (the inner 90% of the column cross section) is rather uniform, which means that the Maldistribution factor can be used as measure of the distribution quality. In the wall zone (the outer 10% of the column cross section) we can have high local liquid loads which very much influence the value of the maldistribution factor. For that reason we present the maldistribution factors calculated over the inner 90% of the cross section.

Wall flow factor

As mentioned above, the maldistribution factor is only applicable for the inner 90% of the column cross section. To compare the wall flow for different packings,
Stükkelman [5] defined a wall flow factor (Wf-factor):

$$Wf = \frac{u_{wall}}{u_{bulk}}$$ (3.2)

where the superficial velocity in the wall zone ($u_{wall}$) is based on the outer 10% of the column cross section.
For the ideal case the Wf-factor equals one. In the case the Wf-factor is larger than one we have a pronounced wall flow. When the Wf-factor is smaller than one we have small wall flow in comparison to the bulk flow.

It is risky to draw conclusions from comparisons of Wf-factors obtained for different column diameters.

### 3.2.3 Modelling of the Liquid Distribution

The packed bed in columns containing structured packing consists of stacked packing elements, which are rotated over 90 degrees to create a good large scale mixing.

In Section 2.2 we showed how we can calculate the liquid distribution between two sheets by a network of interconnected cells. To model the liquid distribution in a packing element, we placed more cell networks parallel to each other, simulating more parallel sheets. From the dimensions of the channels it follows that the cells have a rectangular shape.

The liquid leaving the distributor is supplied to the first packing element. In which cells the liquid is introduced is determined by grid mapping. For the initial distribution this means that the cell directly under a drip point and its neighbour on the adjacent sheet are supplied with liquid. We distributed the liquid over the two neighbouring cells to simulate the splash effect due to the high local liquid load from a drip point. The approach of grid mapping is also followed for the liquid that falls out of one element and enters into the next element, which has a perpendicular orientation. Here the local liquid loads of the rivulets are much lower, which means that we can neglect the splash effect here. In the model we did not turn the element, but we turned the liquid profile. So, liquid that leaves the cells in the bottom layer of an element is turned over ninety degrees (or any other angle) and is supplied to the cells of the top layer of the next element. We assume that the liquid that leaves the cells in the bottom layer flows in form of small rivulets, which means that the outflowing liquid is concentrated in one single point. The single point is located at random determined location in the cell cross section to avoid interference patterns, which can occur as consequence of the rectangular shape of the cell cross sections.
Fig. 3.2  Liquid redistribution over cells (channels) at the transition of one packing element to the next element which is rotated over 90 degrees. a) The liquid leaves a cell in the centre which can lead to interference patterns b) The liquid leaves the cell in a randomly determined location of the cell cross section.

Fig. 3.2 shows how interference patterns, as consequence of the liquid rotation, can lead in the model to totally dry packing sheets, which is not realistic, and how we avoid these interference by using randomly leaving rivulets from a cell.

To compare the simulated liquid profile with the measured one we have to convert the simulated liquid distribution in the packing to the discrete collecting grid. We used the same approach of grid mapping, so the liquid leaves the cells in the bottom layer as a rivulet located at a randomly determined location in the cell cross section and is collected by the collection grid. The fact that the cross section of the collecting tubes is larger than the cross section of the packing channels (or the cells that represent these channels), will result in two or more packing channels ending in the same collecting tube.
3.2.4 RESULTS AND COMPARISONS

Fig.3.3a shows the maldistribution factor measured by Stikkelman [5], after 4 packing elements, for both RALU-PAK and MONTZ-PAK, as function of the liquid load. From this figure it is clear that the maldistribution factor of RALU-PAK is much higher than the one measured for MONTZ-PAK, especially at the low liquid loads. For RALU-PAK the maldistribution factor decreases with an increasing liquid load. From section 2.2 we know that the lateral spreading of RALU-PAK shows the same trend, so it seems that the high maldistribution factor at the low liquid loads is a consequence of the high lateral liquid spreading. For MONTZ-PAK we see that the maldistribution factor is more constant which is also in agreement with the more constant lateral spreading of MONTZ-PAK. The reason that there is a relation between maldistribution and lateral spreading can be explained as follows. In case of a large lateral liquid transport, much liquid is transported to the wall. If the liquid that is transported to the wall is not reflected back into the packing, it will accumulate in the wall zone. At the end of a packing element the liquid at the packing periphery falls in the next element with a perpendicular orientation. The probability that the liquid falls in a channel with an inward orientation is 0.5. This means that a large peak can be transported from the wall to the inside of the packing. In the extreme case where the liquid follows one certain flow path, the height of the peak remains unchanged, which results in high peak in the bulk and so in a high maldistribution factor. So, the extreme case is the situation where the liquid follows the flow path in the channel direction. In this case a high lateral transport is combined with unchanged peaks. This is the situation as it appears for RALU-PAK at low liquid (water) loads. At higher liquid loads the liquid is split into different flows, which reduces the lateral transport. But it also transfers large peaks in several smaller peaks, which results in a lower maldistribution factor. The results of the simulations shown in Fig.3.3b shows that these mechanisms can predict the trend in the maldistribution fairly well.
Fig. 3.4 Measured (a) and predicted (b) wall flow factor for RALU-PAK and MONTZ-PAK after 4 packing elements

We have to be careful in drawing conclusions from small differences between simulated and measured maldistribution factors. In the simulation the maldistribution factor depends on the simulated redistribution between the packing elements. For instance for the simulated maldistribution factor obtained after four elements and a liquid load of 3.4 mm/s, we obtained an average maldistribution factor of 1.21 with a standard deviation of 0.18 for RALU-PAK and an average value of 0.29 and a standard deviation of 0.04 for MONTZ-PAK.

The results obtained for the Wf-factor as function of the liquid load are given in Fig. 3.4a. The Wf-factor for RALU-PAK is substantially larger than the one obtained for MONTZ-PAK, which is also a consequence of the large difference in spreading.

Fig. 3.5 Measured (a) and predicted (b) three dimensional liquid distribution profile for RALU-PAK (above) and MONTZ-PAK (below) after 4 packing elements ($u_0=3.4$ mm/s)
Fig. 3.6 Measured (a) and predicted (b) small scale maldistribution factor for RALU-PAK and MONTZ-PAK as function of the number of drip points \( u_w = 3.4 \text{ mm/s} \).

Fig. 3.7 Measured (a) and predicted (b) three dimensional liquid distribution profiles of RALU-PAK after 2, 4, and 6 packing elements for an initial distribution with one half of the distributor inactive \( u_w = 3.4 \text{ mm/s} \).
behaviour as explained before. The simulated results obtained with a wiper leak factor of 0.6 for both packings are depicted in Fig.3.4.b. For the interpretation of these results we can make the same remark as we made for the maldistribution factor. An average WF-factor 1.90 with a standard deviation of 0.26 for RALU-PAK and an average value 1.13 with a standard deviation of 0.11 for MONTZ-PAK was obtained after four packing elements and a superficial liquid velocity of 3.4 mm/s.

The large differences of the flow profile obtained for RALU-PAK and MONTZ-PAK after four layers of packing and a superficial liquid velocity of 3.4 mm/s are shown in Fig 3.2.5a. The simulated profile is shown in Fig.3.5b.

Having the possibility of reducing the number of drip points, we made some tests to evaluate the sensitivity of liquid distribution to initial distribution. As shown in Fig.3.6a, after the second packing layer both packings have a stable distribution, at absolute Mf-values corresponding to their "natural distribution". A perfect initial liquid profile will be disregarded by the packing itself. Fig.3.6b shows that the same trend is followed by simulation.

Fig 3.8 Measured (a) and predicted (b) three dimensional liquid distribution profiles of MONTZ-PAK after 2, 4, and 6 packing elements for a initial distribution with one half of the distributor inactive ($u_0 = 3.4 \text{ mm/s}$)
The sensitivity of the model can be further illustrated by an example which can be considered as a case of most initial maldistribution [7]. Fig.3.7a and Fig.3.8a show the three dimensional profiles after 2, 4 and 6 packing elements for RALU-PAK and MONTZ-PAK respectively, for the case that only half of the column cross section was irrigated with liquid. As expected from the lateral spreading behaviour, RALU-PAK is less sensitive to severe initial maldistribution than MONTZ-PAK. From a comparison of the model calculation in Fig.3.7b and Fig.3.8b it can be seen that for both packings the simulation reproduces the actual shape of the flow pattern fairly well.

3.2.5 CONCLUDING REMARKS

As clearly shown from the measured results, there are large differences between the maldistribution found for RALU-PAK and MONTZ-PAK. From the fact that these differences can be predicted by the model, we may conclude that the simple spreading experiments with packing sheets provide a good basis for predicting maldistribution in small diameter columns.

Due to the larger lateral transport of RALU-PAK, much liquid is transported to the wall, which will cause large wall flow. On the other hand this means that RALU-PAK can easily restore from a severe initial distribution within a small packing height. For MONTZ-PAK we found no significant wall effects, which is a consequence of the smaller lateral transport in this packing. But limited lateral transport makes MONTZ-PAK also more sensitive to initial distributions.

The continuous splitting of streams as it occurs in MONTZ-PAK, will distribute (mix) the liquid over adjacent sheets. Because the measuring grid used here always receives liquid from two or more opposite oriented channels, it is not possible to draw conclusions about the liquid transfer between adjacent sheets. We will come back to these phenomena in Chapter 5 and 7, where we discuss wetting and mass transfer.

3.3 GAS MALDISTRIBUTION IN SMALL DIAMETER COLUMNS

3.3.1 MODELLING OF THE SMALL SCALE GAS DISTRIBUTION AND MIXING IN COLUMNS

Attempts to calculate the small scale gas distribution were not successful due to the random variation of the channel geometry at the transition between packing elements. So, we could not check whether the assumption of a uniform gas distribution, as concluded by Sohlo and Kouri [8] and Stikkelman et al. [5,6,9], can be confirmed by model calculations. However, in Section 2.3 we saw that the detailed gas distribution model predicts a uniform gas distribution in the case of a
uniform initial gas distribution. Further we saw that the increase of the gas wall flow occurs above the wiper we installed in the wall channel (Fig.2.36). This was explained by the fact that above the wiper the pressure in the wall channel becomes equal to the outlet pressure, which means that the pressure energy of the gas entering the wall channel is converted to kinetic energy. So, the rather high gas velocities measured for the wall channel (Section 2.3) can be seen as an outlet effect. The same occurs with the top element of a packed bed consisting of structured packing. This means that the high gas velocities measured in the wall channels by Stikkelman et al. [5,6] (Fig.3.9) are most probably an outlet effect which makes the assumption of a uniform gas distribution in the structured packings even more realistic.

In our detailed cell model, where the structure of the model is related directly to the packing structure, the assumption of a uniform gas distribution can also cause problems. In Section 3.2 we showed that for the detailed liquid distribution model, the rectangular cell cross section will introduce maldistribution, due to redistribution between two packing elements which have a perpendicular sheet orientation. To avoid any introduction of gas maldistribution at the transition between packing elements we divided the cells into 12 equal small square parts and turned each square part separately. As shown in Fig.3.10 each cell in the bottom layer of the next packing element will receive exactly twelve small square parts. In this way a uniform gas distribution will remain uniform if it leaves a packing element and enters the next packing element, which is rotated over ninety degrees.

The mixing is described in the same way as presented in Section 2.3, where we showed that for a uniform gas distribution, we may use the splitting factor approach to describe the gas mixing between crossing channels and that we may assume total mixing in the wall channel. How the wall (duo)cells for the different packing layers are connected is shown in Fig.3.11.
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Fig.3.10 Illustration of how introduction of gas maldistribution due to the redistribution at the transition from one element to the next element, rotated over 90 degrees, can be avoided in the detail gas model by dividing the cell cross sections in 12 square parts.

Fig.3.11 Illustration of wall duo cells in the detail gas model.
MALDISSION IN SMALL DIAMETER COLUMNS

Fig. 3.12 Tracer point source distribution after 1, 2 and 3 packing elements a) measured for MELLAPAK (u_g=2.3 m/s) by Meier et al. [2] b) obtained with the detailed gas (mixing) model with a splitting factor of 0.15

Here we will check whether we can simulate the results of tracer spreading obtained by Meier et al. [2] in a 0.45 m ID column with MELLAPAK 250Y. Fig.3.12a shows the measured results by Meier after 1, 2 and 3 packing layers. It seems that these results are in disagreement with the conclusion earlier made by Stikkelman [6] and confirmed in Section 2.3, that the channel flow (lateral transport) is much more pronounced than the mixing between crossing channels.

Because we did not know precisely how Meier injected and measured the tracer we will assume in the simulation that the tracer is injected in two neighbouring channels in adjacent packing sheets and that the outlet tracer concentrations are the average outlet concentrations of neighbouring channels in adjacent sheets. The lateral spreading was simulated with a splitting factor of 0.15 as obtained in Section 2.3 for MONTZ-PAK at superficial gas velocities above 1 m/s. Fig.3.12b shows the simulated tracer distribution. From the comparison of Fig.3.12a vs 3.12b we may conclude that the model can simulate the gas spreading behaviour fairly well.

The fact that the simulation was carried out with the same splitting factor as used in Section 2.3 we further may conclude that the difference in tracer distribution profiles presented in Section 2.3 and the profiles (obtained after one packing element) presented in this Section are a consequence of the injection and measuring method.

3.3.2 CONCLUDING REMARKS

The details of the small scale gas distribution in structured packing proved to be complex and not predictive. As shown in Section 2.3, and as earlier concluded by Sohlo and Kouri [8] and Stikkelman et al. [5,6,9], a uniform gas distribution seems
to be satisfactory in case of a uniform initial gas distribution.

A comparison between tracer distributions measured for MELLAPAK by Meier et al. [2] and simulated tracer distributions shows that the splitting factor approach enables satisfactory simulation results as earlier concluded in Section 2.3.

The knowledge of the gas mixing in structured packings could be of importance for the prediction of the effects of liquid maldistribution on the separation performance of distillation columns containing structured packing. We will come back on this in Chapter 7.

### NOTATION

<table>
<thead>
<tr>
<th>SYMBOLS</th>
<th>Description</th>
<th>Unit</th>
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</thead>
<tbody>
<tr>
<td>n</td>
<td>Number of collecting compartments</td>
<td>-</td>
</tr>
<tr>
<td>Mf</td>
<td>Maldistribution factor</td>
<td>-</td>
</tr>
<tr>
<td>Wf</td>
<td>Wall flow factor</td>
<td>-</td>
</tr>
<tr>
<td>$u_{gs}$</td>
<td>Superficial gas velocity</td>
<td>m/s</td>
</tr>
<tr>
<td>$u_{ls}$</td>
<td>Superficial liquid velocity</td>
<td>mm/s</td>
</tr>
<tr>
<td>$u_{bulk}$</td>
<td>(Superficial) liquid velocity for the bulk zone</td>
<td>mm/s</td>
</tr>
<tr>
<td>$u_{l}$</td>
<td>Local (superficial) liquid velocity</td>
<td>mm/s</td>
</tr>
<tr>
<td>$u_{wall}$</td>
<td>(Superficial) liquid velocity for the wall zone</td>
<td>m/s</td>
</tr>
</tbody>
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4.1 INTRODUCTION

Most of the research on maldistribution in columns containing structured packing is carried out in small columns, where the wall effect plays an important role. This is not so in large columns, which means that if there are no other irregularities that can create maldistribution, the liquid and gas distribution will depend only on the initial distribution. However, in large columns the packing layers are built of blocks. Between blocks there are small gaps that can disturb the liquid and gas flow. To study the so called gap effects and to investigate how fast the packing can restore from large scale initial maldistribution a large rectangular column simulator (3x0.5x4.4 m³) was built.

To simulate the large scale liquid distribution in our column simulator, we can use the detailed model presented in Section 2.2, but it is not very practical, because of the enormous number of cells that are involved in this kind of calculations. For
that reason we also developed a design scale model that is much simpler and is based on large cells. Both models will be discussed in Section 4.2 and we will make an evaluation how we can use these models to predict large scale maldistribution in large columns.

In Section 4.3 we will discuss the gas spreading and mixing. For the gas distribution we will only present a design scale model based on large cells. In the case of a uniform (initial) gas distribution we will show that we can use a simple splitting factor model to simulate the gas mixing behaviour. Further we will show on the basis of experimental data that we can apply the theory from Section 2.3 on the larger cell in design scale gas flow distribution model.

4.2 LIQUID DISTRIBUTION IN A LARGE COLUMN SIMULATOR

4.2.1 EXPERIMENTAL SET UP

The experimental equipment is shown in Fig.4.1. It consists of the water circuit, air circuit and the rectangular packed column with a cross section of 3.0x0.5 m² and a maximum packing height of 4.4 m. The top of the column is divided in fifteen compartments with equal dimensions (0.2x0.5 m²). The water, pumped from the

![Schematic diagram of large diameter column simulator](image)
storage tank, is supplied to each compartment and the flow to each compartment is measured separately. All the fifteen liquid distributors are narrow trough gravity type distributors with 100 drip points per square meter resulting in 10 drip pipes predistributor. The drip pipes were provided with five vertically placed small holes (ID of holes from bottom to top is 3.5 mm, 4 mm, 5 mm, 8.5 mm, 8.5 mm) resulting in a large operating range. After having passed the packed bed, the water was collected in the fifteen separated bottom compartments. For each compartment the local liquid load is measured. This is done by closing the outlet valve and measuring the time needed to get a certain level in the compartment. The inverse of the time registrations gives us the two dimensional liquid flow distribution profiles. The outlets of the compartments are connected via a central return pipe with the storage tank. The installed centrifugal pump enables volumetric flow rates equivalent to liquid loads up to 20 mm/s (72 m³/m² h).

The packings tested were RALU-PAK 250 YC and MONTZ-PAK B1 250. RALU-PAK has an element height of 0.21 m. The distribution profiles were measured for two bed heights, one consisting out of 19 packing layers (4.0 m) and one out of 11 packing layers (2.2 m). The packing layers were built of different blocks with a maximum width of 0.3 m and a length that varies form 0.5 to 1.6 m.

![Fig.4.2 Bed lay out of RALU-PAK as packed in large diameter column simulator](image1)

![Fig.4.3 Bed lay out of MONTZ-PAK as packed in large diameter column simulator](image2)
The packed bed with small discontinuities between the blocks is shown in Fig.4.2. The element height for MONTZ-PAK B1 250 is 0.195 m. For this packing we measured the distribution profiles after 21 packing layers (4.1 m) and 11 packing layers (2.15 m). The packed bed of MONTZ-PAK was built of blocks with a length of 0.5 and 1.5 m and a width of 0.5 and 1 m. Fig.4.3 shows where the gaps between packing elements will appear. In the next sections we will refer to the two bed heights as 4 and 2 m.

4.2.2 LIQUID DISTRIBUTION MODEL BASED ON SMALL CELLS

The cells and distribution mechanisms of the small scale model are directly related to the packing used, which means that this is independent of the column size. This means that we can build the packed bed model of the large rectangular column simulator directly out of the small cells. Here we only have to place the wall cells in the right position to simulate the walls of the rectangular column. Besides the wall irregularities we also have the irregularities between the packing blocks. These irregularities are simulated in exactly the same way as we simulate the wall, only the gaps do not have a wall wiper. This means that at the gaps more liquid is accumulated than at the wall.

The inlet liquid is supplied to the cells in the top layer, located directly under the drip pipes of the liquid distributors. The liquid that is coming out of the cells in the bottom layer is collected in the same way as described in Section 3.2, however the collecting compartments are here much bigger, which means that one compartment collects liquid from a large number of cells (packing channels). The simulation is carried out from top to bottom in the same way as described in Section 3.2. From our experiments and simulation results we know that the orientation of the outer packing sheets is very important. In the model we took exactly the same channel orientations as in the equipment.

4.2.3 LIQUID DISTRIBUTION MODEL BASED ON LARGE CELLS

The design scale liquid distribution model consists of large cells, which means that one cell represents a large number of packing channels. The channel orientation of corrugated sheet packings permits only a lateral transport in two directions within a cell. This is shown in Fig.4.4. The cells do not represent a physical dimension, which means that we can choose the size. To keep the model as simply as possible, we used cubic cells with a maximum dimension of a packing layer. This means that if we choose the largest possible cells, we have only one mixing cell in vertical direction per packing layer, as proposed by Hoek et al. [1]. Following further the approach of Hoek, we have a splitting between the flow downwards and the flow in the left-right direction for the packing layer in which the channels have a left right orientation, and a splitting between the flow downwards and the flow in the front-
back direction for the layers in which the packing channels have a front-back orientation. This is shown in Fig.4.4. In this way we easily can describe the large scale lateral mixing behaviour for different types of structured packing. Packing types with a large lateral spreading behaviour should be simulated by large splitting factors. Structured packings with less lateral spreading should be simulated with smaller splitting factors. For the cells in the wall zone one direction (the direction towards the wall) is closed. So the wall-cell can transport liquid downwards or back into the packing (Fig.4.4).

As mentioned before, the packing layers of large columns consists of a number of packing blocks. Between these blocks there are always small gaps that restrain the liquid to flow to the adjacent packing block, resulting in an accumulation of liquid around gaps. For this reason the cells that include gaps between packing elements have a smaller splitting factor. Therefore, more liquid will leave the cell in the downward direction.

The use of large cells has the advantage that we have a smaller number of cells. However, we can not consider small scale variations. For instance in case of peak loads, as around gaps and near the wall, this is not very realistic. That is the reason why we built in the possibility of two or more vertically placed cells per packing.
layer. The cells that represent one packing layer, split the liquid in the same direction (e.g. left-right direction). The cells in the next packing layer distribute the liquid in the perpendicular direction (front-back direction). In Chapter 8 we will demonstrate the relation between chosen cell size and the predicted mass transfer efficiency. The emphasis here is calculation of the splitting factors for the different cell sizes.

**Splitting factor determination**

As mentioned before, for the liquid flow we distinguish different splitting factors:
- a splitting factor for cells that represent the bulk of the packing,
- a splitting factor for cells that represent the wall zone,
- a splitting factor for cells that represent the area around the gaps between adjacent packing blocks.

For the design scale model the splitting factors are determined by the results of large scale spreading tests, presented in Section 4.2.4. We can simply determine the splitting factors by fitting our experimental result to the splitting factor model, but in the case that all measured profiles have the appearance of a normal distribution (no irregularities as wall effects and gaps) we can use the concept of lateral dispersion theory.

Because of the geometry of the rectangular column we can assume that our spreading results represent a two-dimensional case. Then the normal distribution is given by:

\[
f(x) = \frac{1}{\sqrt{2\pi}s^2} e^{-\frac{x^2}{2s^2}} \tag{4.1}\]

where \( x \) is the length parameter and \( s \) is the standard deviation.

If we integrate this equation over a certain length we can find the fraction of the liquid that is collected over this length (L).

\[
fraction = \int_{-L}^{L} \frac{1}{\sqrt{2\pi}s^2} e^{-\frac{x^2}{2s^2}} dx \tag{4.2}\]

Because we can calculate the fraction liquid collected over a certain length out of the measurements, the only unknown in this equation is the standard deviation, which can be determined numerically by solving the integral. From the standard
deviation we can calculate the lateral spreading coefficient with:

\[ D_r = \frac{s^2}{2z} \tag{4.3} \]

where \( D_r \) is the spreading coefficient and \( z \) is the bed height.

From Stikkelman [2] and our own evaluation we know that there is a very simple empirical relation between splitting factors and the spreading coefficient for different cell dimensions. For cubic cells we can write this relation as follows:

\[ S_l = \frac{D_r}{0.25 h_{cell}} \tag{4.4} \]

where \( S_l \) is the liquid splitting factor and \( h_{cell} \) is the cell dimension.

An interesting point of this relation is that we can calculate the splitting factor for different cell dimensions, that simulate the same spreading behaviour.

### 4.2.4 Results and Comparisons

We will first present the experimental data and then compare them with simulation results obtained from the detailed model and the design scale model.

#### 4.2.4.1 Experimental Results

*Uniform initial liquid distribution*

For the liquid distribution we did a number of tests to investigate the large scale distribution and spreading behaviour. Fig.4.5 shows the liquid distribution at a moderate liquid load for two different packing heights for both RALU-PAK and MONTZ-PAK. From Fig.4.5 it is clear that there are differences between the large scale distribution behaviour, as it appears in our column simulator, and the small scale behaviour as discussed in Section 3.2. This is in particular true if we look at the results for RALU-PAK. For small scale tests (Section 3.2) we found for RALU-PAK a significant liquid accumulation in the wall zone, and here we see the contrary, i.e. a large scale liquid transport to the centre of the column simulator. For MONTZ-PAK the distribution profile for both small and large scale is rather flat. At low liquid loads the difference between small and large scale as it appears for RALU-PAK is even more pronounced, as shown in Fig.4.6a, which represents the
Fig. 4.5 Measured liquid distribution after 2 (above) and 4 m (below) of packing for RALU-PAK (a) and MONTZ-Pak (b) obtained with a uniform initial distribution ($u_i = 2.5 \text{ mm/s}$)

Fig. 4.6 Measured liquid distribution after 4 m of packing for RALU-PAK (a) and MONTZ-Pak (b) obtained with a uniform initial distribution ($u_i = 0.5 \text{ mm/s}$)

distribution profile for RALU-PAK after 4 m of packing for a superficial liquid velocity of 0.5 mm/s. The large scale distribution profile for MONTZ-PAK at low
liquid loads is shown in Fig.4.6b. It is less flat than the profile found for moderate liquid loads however it does not show serious liquid maldistribution.

The liquid peaks in the compartments 7, 8 and 9 for RALU-PAK and in compartment 8 for MONTZ-PAK represent the liquid accumulation around the gaps. The gaps between packing blocks restrain the liquid from flowing to the adjacent packing block and so the gap will behave similar to a wall without wall wiper. The under-irrigated wall zones are probably a consequence of the rectangular bed. If we look more in detail to the structure of the packed bed of RALU-PAK, shown in Fig.4.7, we see that at the left hand side the outside sheet, perpendicular to the width of the simulator, will transport the liquid wall flow to the front side of the column. For the left hand side of the column this will result in high local liquid loads at the front side wall. In the next packing layer the outside sheets at the front side parallel to the width of the column has an orientation to the right. This sheet receives very high local liquid loads which results in a overall transport of liquid towards the right. On the right hand side we have the same situation, only there it will result in a net liquid transport towards the left. So at both sides there is a net liquid transport to the centre of the column, due to the orientation of the outside packing sheets. From observation we know that, due to the fact that the packing blocks of RALU-PAK fitted very tightly in the column simulator, high wall flows will result in a film.
between the column wall and the channel ridge, which will result in a wall flow that follows the channel direction. This effect will reinforce the effect of the orientation of the outside packing sheets.

The influence of the orientation of the outside sheets can also be found for MONTZ-PAK. At the right hand side we have the same situation as described for RALU-PAK. However at the left hand side the outside sheets perpendicular to the width of the column has also an orientation towards the backside as shown in Fig.4.8. This means that for both, the left and the right hand side, the backside sheet with an orientation towards the right receives more liquid which results in a net liquid transport towards the right which explains the small peak in compartment 15. As discussed in [3] the cut ends of packing sheets may also contribute to the asymmetrical shape of the distribution curves. For MONTZ-PAK the local liquid loads in the wall zones are much lower than for RALU-PAK. This is the reason for the fact that for MONTZ-PAK we find a relatively small effect of the net liquid transport due to the orientation of the outside sheets on the overall liquid distribution profile.

These test show that details of packing lay-out can have a large influence on the distribution behaviour of structured packing. We will come back on this later during discussion of the simulated results of the detailed model.
Fig. 4.9 Measured tracer distribution after 2 (above) and 4 m (below) of packing for RALU-PAK (a) and MONTZ-PAK (b) obtained with a uniform initial distribution and a tracer injection in top compartment 4 ($u_a = 2.5$ mm/s).

Tracer and liquid point source distributions

Besides the liquid distribution we have also studied the tracer spreading behaviour. Fig. 4.9 shows experimental tracer distribution curves at a moderate liquid load (2.5 mm/s) after 2 and 4 m of packing for both RALU-PAK and MONTZ-PAK. The tracer was introduced in compartment 4 to minimize gap and wall effects on the spreading behaviour. If we look at the tracer distribution for RALU-PAK in Fig. 4.9a it is clear that the tracer distribution curve is not symmetrical. The highest tracer concentration is found in bottom compartment 5 and not in compartment 4 where it was initially introduced at the top. This can be explained from the net liquid transport towards the centre of the column as discussed earlier. Further we see that the distribution curve determined after 2 and 4 m are almost the same. This is probably a consequence of the increasing periphery flow. From Section 2.2 we know that for RALU-PAK the lateral spreading of water decreases with an increasing
'local' liquid load. In the upper part of the packed bed the liquid is rather
good distributed, which means that the spreading behaviours of the bulk and the periphery
zone are equal. However, in the bottom part of the packed bed we observed that
almost all the liquid flows through the periphery zone. Due to the high liquid loads
in the periphery zone the lateral spreading will decrease which explains the small
difference between the obtained tracer distribution curves after 2 and 4 m.

Both facts indicate that the spreading behaviour for RALU-PAK, as it appears in
our rectangular column simulator, does not comply with the dispersion theory. This
makes the determination of spreading coefficients in this situation meaningless.

If we look at the tracer distribution curves for MONTZ-PAK in Fig.4.9b, we see
that these curves are almost symmetrical. Table 4.1 shows that for MONTZ-PAK the
spreading coefficients, determined by the two-dimensional curves measured after 2
and 4 m, are equal which indicates that the dispersion theory holds for this packing
configuration.

Fig.4.10 shows the liquid point source distribution after 4 m. The similarity
between these results and that of Fig.4.9b confirms the conclusion made in Section
2.2 that there is hardly any difference between the liquid flow spreading and tracer
spreading mechanisms in structured packing.

<table>
<thead>
<tr>
<th></th>
<th>( u_p = 0 \text{ m/s} )</th>
<th>( u_p = 3 \text{ m/s} )</th>
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</thead>
<tbody>
<tr>
<td></td>
<td>( z = 2 \text{ m} )</td>
<td>( z = 2 \text{ m} )</td>
</tr>
<tr>
<td>( u_p = 0.5 \text{ mm/s} )</td>
<td>0.0155 m</td>
<td>0.0165 m</td>
</tr>
<tr>
<td>( u_p = 2.5 \text{ mm/s} )</td>
<td>0.0155 m</td>
<td>0.013 m</td>
</tr>
</tbody>
</table>

Table 4.1 Liquid spreading coefficients for MONTZ-PAK (design scale model)

![Fig.4.10](image1)  ![Fig.4.10](image2)

*Fig.4.10* Measured point source liquid distribution after 4 m of packing for RALU-PAK (a) and
MONTZ-PAK (b) obtained with a liquid point source in compartment 4
Fig.4.11 Measured tracer distribution after 4 m of packing obtained with a tracer injection in compartment 1 (above) and 8 (below) for RALU-PAK (a) and MONTZ-PAK (b) for a uniform initial distribution ($u_0 = 2.5 \text{ mm/s}$)

To investigate the spreading behaviour near the wall, the tracer was introduced in compartment 1; to see how the gap will influence the spreading behaviour the tracer was introduced in compartment 8. The results for both RALU-PAK and MONTZ-PAK after 4 m of packing are shown in Fig.4.11. If we compare Fig.4.11 vs Fig.4.9b we can see that the liquid spreading around the gap is smaller than the liquid spreading in the bulk of the packing.

Initial liquid maldistributions

To study the influence of the spreading behaviour of structured packings on the sensitivity to initial maldistribution, we did two tests with different initial maldistributions. For one test we plugged the distributors in the uneven compartments. This will result in an initial distribution with an under-irrigated wall region. For the other test we plugged the distributors in the even compartments,
Fig. 4.12 Measured liquid distribution after 4 m of packing for a initially under-irrigated wall zone (above) and a pronounced initially irrigated wall zone (below) for RALU-PAK (a) and MONTZ-PAK (b), obtained with an initial distribution in which alternate distributors were plugged ($u_w = 2.5 \text{ mm/s}$)

which results in an initially over-irrigated wall zone. The results obtained after 4 m for both RALU-PAK and MONTZ-PAK are shown in Fig. 4.12a and b respectively. If we compare the results with the results with a uniform initial distribution it is clear that, even after 4 m of packing, we can still find an influence of the initial maldistribution.

However we have to be careful with the interpretation of the results obtained for RALU-PAK in our rectangular column simulator. As discussed earlier, due to the wall flow and the orientation of the outside sheets, there will be a decreasing spreading tendency over the column height and a net liquid transport towards the middle of the column simulator. These two effects strongly influence the overall distribution profile.
Liquid distributions at loading conditions

Until now we only looked at the large scale liquid distribution without counter current gas flow. From Stikkelman [2], Kouri and Sohlo [4] and our own experimental data we know that below the loading point counter current gas flow has hardly any influence on the distribution behaviour of the liquid phase. However, in the loading range Stikkelman [2] found gas liquid segregation. Fig.4.13 shows the liquid distribution for both RALU-PAK and MONTZ-PAK measured in our column simulator at loading conditions after two different packing heights. From the comparison of Fig.4.13a and Fig.4.5a we see that for RALU-PAK the loading conditions will not change the distribution profile significantly. Only the transport of liquid towards the centre of the column simulator is a little bit more pronounced. For MONTZ-PAK we see an enormous liquid accumulation around the gap in the centre of the column.

Fig.4.13 Measured liquid distribution after 2 (above) and 4 m (below) of packing at loading conditions (pressure drop between 3 and 10 mbar/m) for RALU-PAK (a) and MONTZ-PAK (b), obtained with a uniform initial distribution ($u_0=2.5$ mm/s)
Chapter 4

**Fig. 4.14** Measured tracer distribution after 2 (above) and 4 m (below) of packing at loading conditions (pressure drop between 3 and 10 mbar/m) for RALU-PAK (a) and MONTZ-PAK (b), obtained with a uniform initial distribution and a tracer injection in top compartment 4 (u_e = 2.5 mm/s)

Tracer point source distributions at loading conditions

To see whether the liquid accumulation around these gaps in the loading range is accompanied by a change in the lateral spreading behaviour, we repeated the tracer injection tests for compartment 1, 4 and 8 for loading conditions. Fig. 4.14 shows the tracer distribution curves for loading conditions for both RALU-PAK and MONTZ-PAK after 2 and 4 m of packing in the case that the tracer point source was introduced in compartment 4. From the comparison of Fig. 4.14 vs Fig. 4.9 it is clear that the liquid spreading, obtained in our rectangular column simulator, decreases in the loading range. The spreading coefficient for MONTZ-PAK at this condition is given in Table 4.1. Fig. 4.15 shows the tracer distribution for the case the tracer was introduced in compartment 1 and 8 respectively. The comparison of Fig. 4.15 vs Fig. 4.11 shows that around irregularities we also have a decrease of liquid spreading in the loading range. However, the spreading around the gaps
Fig.4.15 Measured tracer distribution after 4 m of packing at loading conditions (pressure drop between 3 and 10 mbar/m), obtained with a tracer injection in compartment 1 (above) and 8 (below) for RALU-PAK (a) and MONTZ-PAK (b) for a uniform initial distribution ($u_m = 2.5$ mm/s).

decreases more (Fig.4.15b) than the spreading in the bulk of the packing. So in the loading range more liquid is transported from the bulk to the gaps than from the gaps to the bulk in comparison with the normal condition. This results in a larger liquid accumulation around the gaps.

4.2.4.2 RESULTS OF THE DETAILED DISTRIBUTION MODEL

To simulate the large scale liquid transport, as it appears in our rectangular column simulator, we built in the detailed model a specific wall effect which influences the liquid transport at high liquid loads. Above a critical wall flow the gap between the wall and neighbouring parallel sheet of packing is filled with liquid. The liquid follows then the orientation of the sheet. So the orientation of the outside sheets becomes important and should be identical to the experimental situation.

As mentioned earlier, the packing blocks of RALU-PAK fitted tightly in our column simulator, which will promote the above mentioned effect. Here we will show some simulation results for liquid distribution without counter current gas flow.
Fig.4.16 Liquid distribution simulated with detailed model after 2 (above) and 4 m (below) of packing for RALU-PAK (a) and MONTZ-Pak (b), obtained with a uniform initial distribution ($u_{0a} = 2.5 \text{ mm/s}$)

Fig.4.16 shows the simulated distribution profile after 2 and 4 m obtained, with uniform initial liquid distribution and a superficial liquid velocity of 2.5 mm/s for RALU-PAK and MONTZ-PAK respectively. As mentioned in Section 3.2 the simulation model is sensitive to re-distribution between packing elements. This is of special importance in case of relatively high local liquid loads, which we can expect for RALU-PAK. That is the reason why we presented in Fig.4.16 the average profile of 3 simulations. We further assumed in this simulations that, if the local liquid wall flow exceeds a velocity of 25 mm/s, which means here ten times the superficial liquid velocity, the liquid wall flow will follow the channel direction of the adjacent packing sheet. Below this critical wall flow the liquid on the wall flows straight downwards. For MONTZ-PAK the wall flow will only exceed this limit in the first two packing layers, which means that for MONTZ-PAK assumptions concerning the wall flow are of minor importance.

From the comparison of the simulated and experimental profiles (Fig.4.16 vs Fig.4.5) we may conclude, that the trend is predicted reasonably well.
Fig.4.17 Tracer distribution simulated with detailed model after 2 (above) and 4 m (below) of packing or RALU-PAK (a) and MONTZ-PAK (b), obtained with a uniform initial distribution and a tracer injection in top compartment 4 ($v_w = 2.5 \, \text{mm/s}$)

Fig.4.17 shows the distribution of the tracer obtained with a uniform initial irrigated packing at a superficial water velocity of 2.5 mm/s. The tracer was injected in compartment 4. From comparison with the experimental tracer distributions (Fig.4.9a) we can conclude that the lateral spreading simulated for RALU-PAK is much larger than the experimental one, especially if we compare the results after 4 m. In Section 4.2.4.1 we already mentioned that the lateral spreading reduces with increase of the packing height. For the simulated tracer distribution profile we also found a small reduction of the lateral spreading with increasing packing height, however it is considerably less pronounced. The simulated spreading behaviour for MONTZ-PAK is in agreement with the experimental one.
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![Large cell tracer distribution and small cell tracer distribution graphs](image)

Fig.4.18  (Tracer) distribution calculated with design scale model after 2 (above) and 4 m (below) of packing for large cells (0.2 m) (a) and smaller cells (0.067 m) (b) and a (tracer) point source in top compartment 4 for the normal operating range.

4.2.4.3 RESULTS OF THE DESIGN SCALE DISTRIBUTION MODEL

As shown before, the spreading behaviour of RALU-PAK as it appears in our rectangular column simulator, does not obey the dispersion theory, which also implies that it is not possible to calculate the tracer and flow distribution profiles of the column simulator with the design scale model. For this reason we will limit our calculations with the design scale model to MONTZ-PAK only. We will assume the height of a layer equal to 0.2 m. The calculations were carried out for a bed consisting of 21 packing layers (4.2 m) and 11 packing layers (2.2m).

Fig.4.18 shows the calculated tracer distribution profiles after 2.2 and 4.2 m in case the tracer is introduced in compartment 4. For these calculations we used a cell size of 0.2 m (Fig.4.18a) and 0.067 m (Fig.4.18b). The splitting factors for the cells that represent the bulk of the packing were calculated, for both cell sizes, with Eq.(4.4) in which we used the experimentally determined spreading coefficient from...
**Fig. 4.19** (Tracer) distribution calculated with design scale model after 4 m of packing obtained with a (tracer) point source in top compartment 1 (above) and 8 (below) for large cells (0.2 m) (a) and smaller cells (0.067 m) (b) for normal operating range

Table 4.1. The splitting factors for the wall and the gap were determined by fitting the model on the experimental data for the case the tracer was introduced in compartments 1 and 8 respectively. The results are given in Fig.4.19, which shows

<table>
<thead>
<tr>
<th></th>
<th>( u_{gs} = 0 \text{ m/s} )</th>
<th>( u_{gs} = 2.5 \text{ mm/s} )</th>
<th>( u_{gs} = 3 \text{ m/s} )</th>
<th>( u_{gs} = 2.5 \text{ mm/s} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>( h_{cell} )</td>
<td>0.2 m</td>
<td>0.067 m</td>
<td>0.2 m</td>
<td>0.067 m</td>
</tr>
<tr>
<td>( S_{l \text{ bulk}} )</td>
<td>0.29</td>
<td>0.87</td>
<td>0.15</td>
<td>0.45</td>
</tr>
<tr>
<td>( S_{l \text{ gap}} )</td>
<td>0.24</td>
<td>0.42</td>
<td>0.08</td>
<td>0.08</td>
</tr>
<tr>
<td>( S_{l \text{ wall}} )</td>
<td>0.18</td>
<td>0.72</td>
<td>0.13</td>
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</tr>
</tbody>
</table>

**Table 4.2** Liquid splitting factors for MONTZ-PAK
Fig.4.20  Liquid distribution simulated with design scale model after 4 m of packing for large cells (0.2 m) (a) and smaller cells (0.067 m) (b) for normal operating range.

Fig.4.21  Liquid distribution simulated with design scale model after 4 m of packing for a initially under-irrigated wall zone (above) and a pronounced initially irrigated wall zone (below) for large cells (0.2 m) (a) and smaller cells (0.067 m) (b) obtained with an initial distribution in which alternate distributors were plugged.
the distribution curves calculated with a cell size of 0.2 m and 0.067 m respectively. The used splitting factors are given in Table 4.2. To simulate the tracer distribution around a gap, the accumulation of liquid in a small gap cell should be in accordance with reality. Therefore the splitting factor for the gap cell should increase slower with a decrease of the cell size than the splitting factor of the bulk cells. For the same reason the splitting factor of the wall should increase faster with a decrease of the cell size than the splitting factor of the bulk to simulate the same net liquid transport from the wall into the packing.

Now that we know the splitting factors for all cells we can calculate the liquid distribution profile for different initial profiles. Fig.4.20a and b show the simulated liquid distribution after 4.2 m in the case of a uniform initial distribution for the two different cell sizes.

We also carried out simulations for both cell sizes for the case that alternate distributors were plugged as shown in Fig.4.21a and b. The upper diagrams show the results for the case that the uneven distributors were plugged. From the distribution

![Image](image-url)

**Fig.4.22** (Tracer) distribution calculated with design scale model after 2 (above) and 4 m (below) of packing for large cells (0.2 m) (a) and smaller cells (0.067 m) (b) and a (tracer) point source in top compartment 4 for the loading range.
curve after 4.2 m of packing we can still see the effects of an under-irrigated wall zone as we earlier showed for the experimental results. The lower diagrams in this figure show the same kind of simulation but now the distributors in the even compartments were plugged. Here we also may conclude that the results are comparable with the experimental results discussed before and that after 4.2 m the extra liquid in the wall region is still not totally drained away by the lateral liquid transport into the packing.

Although the spreading coefficients calculated from the tracer distribution obtained under loading conditions after 2.2 and 4.2 m of packing are not completely in agreement with each other, we assume that the tracer distribution under loading conditions will follow the dispersion theory. Fig.4.22a and b show the simulated tracer distribution under loading conditions for cell sizes of 0.2 m and 0.067 m respectively. Although for both cases the splitting factor was obtained from Eq.(4.4) with the experimentally determined spreading coefficient, we see that for small spreading coefficients the cell size has a small influence on simulated tracer spreading. Apparently Eq.(4.4) does not hold for large cells in combination with

![Fig.4.23](image)

Fig.4.23 (Tracer) distribution calculated with design scale model after 4 m of packing obtained with a (tracer) point source in top compartment 1 (above) and 8 (below) for large cells (0.2 m) (a) and smaller cells (0.067 m) (b) for the loading range.
small spreading coefficients. The differences however are within the accuracy range of the experiments.

Fig.4.23 shows the calculated tracer distributions after 4.2 m in the case that the tracer was introduced in compartment 1 near the wall and in compartment 8 (in the centre). The obtained splitting factors for the wall and the gap cells are given in Table 4.2.

Fig.4.24 shows the simulated liquid distribution under loading conditions in case of a uniform initial distribution. Although the splitting factor for the gap for both cell sizes was fitted on the experimental tracer distribution, it is clear that the simulated liquid distribution carried out with the small cells gives a better fit of the experimental profile (Fig.4.13b).

From the fact that no big differences are found in the simulated distribution profiles if we adapt the splitting factors to the cell size, we can conclude that the design scale model with a cell size of 0.2 m (layer height) predicts the liquid distribution profile in columns with sufficient accuracy.

![Fig.4.24](image.png)

**Fig.4.24** Liquid distribution simulated with design scale model after 4 m of packing for large cells (0.2 m) (a) and smaller cells (0.067 m) (b) for the loading range

### 4.2.5 CONCLUDING REMARKS

The experimental results obtained in our rectangular column simulator show very clearly that an initial uniform liquid distribution profile deteriorates gradually over the height of the packed bed. The explanation of the deterioration is always related to (small) structural deviations in the packed bed as walls and gaps.

A problem is that the structural deviations in a rectangular column are not the same as the structural deviation in a round column, which implies that we have to be very careful with the interpretation of the results, especially that for RALU-PAK which are dominated by pronounced wall effects. However, the net liquid transport to
the centre of the column was also found by Bennet et al. [5] in a round column (1.25 m ID) irrigated with water.

The influence of a counter current gas flow may be considered as negligible below the loading point as earlier concluded for small columns by Stikkelman [2] and Kouri and Sohlo [4]. For MONTZ-PAK large scale gas-liquid segregation will occur in the regions of structural deviation, if the packing is operated above the loading range. This is in agreement with results of Stikkelman [2] for a small column. For RALU-PAK no significant deterioration of the liquid profile was found above the loading point.

After incorporation of the observed wall effects, the detailed model was able to simulate the trend in the experimental obtained distribution data. However, the simulated lateral spreading for RALU-PAK is too large. For MONTZ-PAK there was a good agreement between the simulated and experimentally determined lateral spreading. This gives the confidence that the spreading behaviour of MONTZ-PAK can be predicted reliably. So when the spreading behaviour follows the dispersion theory, we can determine the spreading coefficient, from both the experimental tracer spreading profiles and tracer spreading profiles calculated with the detailed model.

Spreading behaviour that follows the dispersion theory can be simulated with the design scale model. For the bulk cells the splitting factor can be calculated directly from the spreading coefficient. For the cells, that represent the gap and wall zone, the splitting factor has to be fitted on the tracer distribution profiles obtained for these zones.
4.3 LARGE SCALE GAS DISTRIBUTION IN COLUMN SIMULATOR

4.3.1 EXPERIMENTAL SET-UP AND PROCEDURE

For the large scale gas flow distribution and tracer mixing experiments we used the same equipment as described for the liquid experiments (Section 4.2). The large blower with a maximum capacity of 7 m³/s enables superficial air velocities high enough to flood common structured packings at low liquid loads. The air is supplied via fifteen separated bottom compartments below the packed bed. The inlet velocity for each compartment is measured with an anemometer. We also measured the absolute pressure for each bottom compartment to establish the inlet pressure profile. The gas flows through the packed bed from bottom to top and the outlet gas velocities were measured at the top for each separate top compartment with an anemometer. In this way we could measure the outlet gas velocity profile for different superficial gas velocities. To get more knowledge about the gas flow path through the packed bed, we carried out some experiments with tracer injections in a uniform flow field. The tracer (CO₂) was introduced in one of the fifteen bottom compartments and the outlet concentrations were measured with an infrared gas-analyzer for each top compartment. This tracer distribution gives us information about the dispersion of a tracer gas in a uniform flow field within a packed bed. We repeated the tracer experiments for different superficial gas velocities.

4.3.2 DESIGN SCALE MODEL BASED ON SPLITTING FACTORS

In Section 2.3 we showed that, in case of a uniform gas distribution, there is no difference between the mixing calculated by turbulent diffusion or the mixing calculated with the splitting factor approach.

Although the cells used here have quite other dimensions than the cells used in the detail gas model presented in Chapter 2, we assume that we can calculate the gas tracer spreading by a splitting factor model. The cells used in the design scale splitting factor model for the gas phase are the same as the cells presented in Section 4.2 for the liquid phase. We will show that we can use the design scale liquid splitting model upside down to simulate the gas tracer spreading in a uniform flow field (uniform initial gas distribution).

4.3.3 DESIGN SCALE GAS FLOW DISTRIBUTION MODEL

In case of a non uniform gas flow distribution we cannot use the splitting factor approach, because non uniformity of the gas distribution will introduce pressure differences over the cross section of the column. These pressure differences determine how the gas will flow through the packed bed. In Section 2.3 we showed
that we can describe the gas flow distribution between two adjacent corrugated sheets by solving the mass, momentum and energy balance for each cell. Here we will use the same balance equations applied to large cells to describe the gas distribution in a column containing structured packing.

With other words, we combine the mixing cell model, based on splitting factors, with a distribution model based on the theory of Section 2.3. To combine these two behaviours in one and the same large cell, we have to consider the two principal differences between the mixing model and the flow distribution model.

1) The flow distribution model assumes channel flow (only flow in the channel direction). In the mixing cell model the flow is split in a flow in the channel direction and a flow in the vertical direction.

2) In the flow distribution model we distinguish a flow in the left or back side direction and a flow in the right or front side direction. For the lateral mixing model these flows are equal and are determined by splitting factors, which means that we do not need to handle these flows separately.

We can overcome the first problem by reducing the cell size till we can simulate the lateral spreading behaviour for a certain packing with a splitting factor equal to 1 (no vertical flow). We can use Eq.(4.4) to calculate the cell size for which the splitting factor is equal to 1. For a splitting factor of 1 this equation will reduce to:

\[ h_{cell} = 4 \ D_r \]  \hfill (4.5)

![Diagram](image)

*Fig.4.25* Large bulk and wall cell(parts) of design scale gas flow distribution model
In Section 4.3.4 we will show that for normal to high gas velocities the cell dimension \( h_{\text{cell}} \) will be approximately 0.07 m.

To model the different flow directions, we have to divide the large cells into two parts. One part that represents the channels with a left (back) side orientation and one that represents the channels with a right (front) side orientation. This is shown in Fig.4.25.

So, in case that flow distribution and tracer mixing have to be simulated simultaneously, the dimensions of a cell in the design scale model have to be approximately \( \frac{1}{3} \) of the height of the packing layer (0.2 m) and the outgoing flows have to be calculated with the balance equations (mass, momentum and energy) of Section 2.3.

At the column wall we assume total reflection. This is simulated by transferring the gas flow in a cell from the part with an orientation towards the wall to the part with an orientation back into the packing. Fig.4.25 shows how the gas flows from a wall cell back into the packing in case we have three vertically placed cells per packing layer.

Each packing layer is rotated over ninety degrees to simulate the large scale mixing behaviour. This also means that, for example in the case we have 3 vertically placed cells per packing layer, the orientation of the cells changes after every 3 vertically placed cells.

Fig.4.26 Large bulk cell(parts) of design scale gas flow distribution model at the transfer of one packing layer to the next packing layer with a perpendicular orientation.
CHAPTER 4

We assume that the inlet velocities of the bottom cells of a certain packing layer can be calculated in the same way as we did for the gas inlet calculation in Section 2.3. So the inlet velocities of the two adjacent (cell) parts of the cells in the bottom layer of a packing element are calculated from the corresponding inlet pressures of previous iteration sweep and the outlet flows of the cells underneath. These cells underneath are located in the top layer of the packing element below with a perpendicular orientation. The outlet pressures of the two adjacent (cell) parts of these latter mentioned cells are taken equal and calculated from the corresponding outlet velocities, and the inlet velocities and pressures of the first mentioned cells. Fig. 4.26 shows how the bottom cells of a packing layer are connected with the cells

![Flow scheme for calculation of gas velocity field with the design scale gas flow model.](image-url)

**Fig. 4.27** Flow scheme for calculation of gas velocity field with the design scale gas flow model.
in the top layer of the packing element located underneath below with a perpendicular orientation. The flow scheme for the calculation of velocity field with the design scale gas flow model is given in Fig.4.27.

To simulate the pressure drop over a large cell we have to adapt the cell pressure loss coefficients (ζ) from Section 2.3. These can be taken proportional to the cell dimensions. The pressure loss coefficient for the entrance losses (ζ_{inlet}) for each packing layer remains unchanged.

4.3.4 RESULTS AND COMPARISONS

Here we will present the experimental results and compare them with the calculated results obtained from the splitting model and the flow distribution model based on theory of Section 2.3.

4.3.4.1 EXPERIMENTAL RESULTS

Uniform initial gas and tracer point source distributions

Fig.4.28a and b show the gas distribution after 4 m of packing for an ideal initial distribution obtained for RALU-PAK and MONTZ-PAK respectively. From these two figures it is clear that we may conclude that the gas flow distribution for a uniform initial distribution remains uniform for both packings. From our experiments we know that for low to moderate liquid loads the gas distribution is not influenced by the liquid. Although it is clear that the gas flow distribution is uniform these

![Ralupak and Montzpak gas distributions](image)

*Fig.4.28* Measured gas distribution after 4 m of packing for RALU-PAK (a) and MONTZ-PAK (b) obtained with an uniform initial gas distribution (a) \( u_g = 4 \) m/s (b) \( u_g = 3 \) m/s
Fig. 4.29 Measured gas tracer distribution after 4 m (above) and 2 m (below) of packing for RALU-PAK (a) and MONTZ-PAK (b) obtained with a uniform initial gas distribution and a tracer injection in bottom compartment 8 ($u_e = 3 \text{ m/s}$)

Experiments do not give us any information about the lateral mixing behaviour. For this reason we did some tracer experiments. Fig. 4.29a and b show the tracer spreading at different packing heights for RALU-PAK and MONTZ-PAK respectively (CO$_2$ was injected in compartment 8). From the shape of the curves we can conclude that under these experimental conditions (no horizontal pressure gradients) the gas tracer distribution has the appearance of a normal distribution.

From the fact that we found an almost uniform outlet flow distribution we may conclude that there is no wall effect. This is confirmed by the tracer distributions shown in Fig. 4.30a and b, resulting from a tracer injection in compartment 15.

We repeated these tracer distribution tests for different liquid and gas loads to investigate if there is any load influence on the spreading behaviour of tracer gas. Fig. 4.31a shows the lateral spreading coefficient as function of the gas velocity for both RALU-PAK and MONTZ-PAK. From this figure it is clear that above a superficial gas velocity of $\pm 1 \text{ m/s}$, which corresponds with Reynolds number (based on the packing channels) of about 1100, the lateral spreading coefficient is practically
Fig. 4.30 Measured gas tracer distribution after 4 m of packing for RALU-PAK (a) and MONTZ-PAK (b) obtained with a uniform initial gas distribution and a tracer injection in bottom compartment 15
(a) \( u_g = 4 \, \text{m/s} \) (b) \( u_g = 3 \, \text{m/s} \)

independent of the gas velocity. At very low gas velocities (below 1 m/s) the lateral spreading coefficient increases. Due to the decrease of turbulence intensity there is less turbulent mixing between crossing gas flows which results in a higher lateral gas transport. This will increase the lateral spreading coefficient.

For MONTZ-PAK we did not find any influence of the liquid load on the lateral spreading coefficient, which is in agreement with the results found in Section 2.3. For RALU-PAK we found an increase of the lateral spreading coefficient with an

Fig. 4.31a Measured dry lateral gas spreading coefficient \( D_r \) [m] after 4 m of packing for RALU-PAK and MONTZ-PAK at different gas rates

Fig. 4.31b Measured lateral gas spreading coefficient \( D_r \) [m] after 2 and 4 m of packing for RALU-PAK at different liquid rates \( (u_g = 2 \, \text{m/s}) \)
increase of the liquid load as shown in Fig.4.31b. In Section 2.3 we found that, for a dry packing, the gas will flow through the slits. However a counter current liquid flow will close the slits, which means that the gas has to follow the channel direction. This is the explanation for the increase of the lateral spreading with an increase of the liquid load.

Striking point in Fig.4.31b is the large difference between the lateral spreading coefficient obtained after 2 and 4 m in case of a counter current liquid flow. This can be explained by an increasing liquid wall flow with an increase of packing height as earlier mentioned in Section 4.2. In the first 2 m the liquid is still well distributed, which means that the liquid will close the slits. However, due to a developing liquid wall flow, in the lower part of the 4 m bed most liquid flows downwards through the wall zone. So in the lower part of the packed bed there is hardly any liquid in the bulk of the packing, which means that most of the slits are open and the gas will flow more vertically. In other words gas-liquid segregation will promote a more vertically oriented gas flow through the slits. The development of the gas-liquid segregation over the packing height in our column simulator explains why the lateral spreading coefficient for the gas phase will decrease with increasing bed height.

**Fig.4.32** Measured gas distribution after 4 (above) and 2 m (below) of packing for RALU-PAK (a) and MONTZ-PAK (b) originating from a initial point source in bottom compartment 8 (u_g=±1.2m/s)
Gas point source distributions

Until now we only investigated the flow behaviour in structured packing under uniform initial conditions. This also means that there are almost no horizontal pressured gradients in the packed bed that can affect the gas flow direction. To investigate the effect we introduced a gas point source which will cause very large horizontal pressure gradients. The gas distribution after different packing heights for the case that the gas was introduced only in compartment 8 is shown in Fig.4.32a and b for RALU-PAK and MONTZ-PAK respectively. It is clear that, due to the horizontal pressure gradients, the gas will spread much faster than in case of the tracer spreading without horizontal pressure gradients. This becomes more obvious if we further look at the outlet gas flow distribution in the case that we introduce gas in compartment 15 only. From these gas distributions we can conclude that, for both packings, the average direction of the gas flow makes a larger angle with the vertical axis than expected from the channel direction. In the case of channel flow in our column simulator, with a maximum bed height of 4.4 m and a width of 3 m, the gas will never reach the left side wall when it is introduced in compartment 15. The fact

![Graphs showing gas distribution](image)

**Fig.4.33** Measured gas distribution after 4 (above) and 2 m (below) of packing for RALU-PAK (a) and MONTZ-PAK (b) originating from a initial gas point source in bottom compartment 15 ($u_o=2.12 \text{ m/s}$)
that we found a positive gas velocity in compartment 1 already after 2 m of packing (Fig.4.33a and b) indicates that, there is a horizontal gas transport in the column, due to the large horizontal pressure gradients. How large these horizontal pressure gradients are, is shown in Fig.4.34 for the case that the gas is introduced in compartment 8 and in Fig.4.35 for the case that we only introduced gas in compartment 15. Interesting point here is that the absolute values of the pressures measured in the bottom compartments are independent of the packing height, which means that for this case the vertical pressure gradient is reversely proportional to the bed height. The explanation can be found in a faster gas spreading, due to a relative increase in horizontal pressures gradient, compared to the vertical pressure gradient for an increasing bed height. This becomes clear if we look at Fig.4.35 where we may consider the pressure in compartment 1 equal to the outlet pressure. This means that the ratio of the horizontal and vertical pressure gradient is equal to the ratio of the bed height and the width of the column simulator. This ratio increases with an increase of the bed height. So the horizontal pressure gradient becomes more important compared to the vertical pressure gradient with an increase of the bed height. This means a faster equalization of the gas profile for deep beds,

Fig.4.34 Measured bottom pressure distribution for a bed height of 4 (above) and 2 m (below) for RALU-PAK (a) and MONTZ-PAK (b) originating from a initial gas point source in bottom compartment 8 \( (u_\text{g} = \pm 1.2 \text{ m/s}) \)
Fig. 4.35 Measured bottom pressure distribution for a bed height of 4 (above) and 2 m (below) for RALU-PAK (a) and MONTZ-PAK (b) originating from a initial gas point source in bottom compartment 15 \((u_p = 21.2 \text{ m/s})\)

which results in a more uniform gas distribution profile. This more uniform profile will result in a lower average vertical pressure gradient. It also means that shallow beds are much more prone to initial gas maldistribution than deep beds.

4.3.4.2 RESULTS OF THE DESIGN SCALE GAS MODEL

We will illustrate for MONTZ-PAK that we can simulate the gas tracer distribution in a uniform gas distribution with the same approach as presented for the liquid phase, if we take a splitting factor equal to 1 and calculate the cell size with Eq.(4.4) or Eq.(4.5). Further we will show that in case of channel flow (splitting factor equal to 1), we can simulate the gas flow distribution of a non uniform initial distribution by solving the balance equations of Section 2.3 over the cells of the design scale gas distribution model.

If we use Eq.(4.5) with the experimental determined spreading coefficient, we will find a cell size of approximately 0.07 which means 3 vertical placed cells per
Fig. 4.36 Calculated gas tracer distribution after 4.2 m (above) and 2.2 m (below) for 3 vertically placed cells per packing layer and a splitting factor equal to 1, in case the tracer was injected in bottom compartment 8.

Fig. 4.37 Calculated gas tracer distribution after 4.2 m (above) and 2.2 m (below) for 3 vertically placed cells per packing layer and a wall splitting factor equal to 0.5, in case the tracer was injected in bottom compartment 15.

packing layer. The simulated tracer distribution for MONTZ-PAK after 4.2 and 2.2 m calculated with this cell size are shown in Fig. 4.36. The tracer was introduced in compartment 8. From the comparison with the experimental determined tracer profile (Fig. 4.36 vs Fig. 4.29b) we see that even for the extreme case of a splitting factor equal to 1, Eq. (4.4) still holds.

According to the definition of the wall splitting factor (Fig. 4.5) we have to take the splitting factor of the wall cell half of the splitting factor of a bulk cell to avoid accumulation near the wall. The simulated tracer distribution in case the tracer was injected in compartment 15 is shown in Fig. 4.37 which is in good agreement with the experimentally obtained tracer distribution in Fig. 4.30b.

Fig. 4.38a shows the outlet flow profile after 2.2 m of packing originating from an initial gas point source in bottom compartment 8. The corresponding pressure distribution in the bottom compartment is shown in Fig. 4.38b. Fig. 4.39 shows the results for a gas point source introduced in compartment 15. If we compare the
Fig. 4.38 Simulated gas flow distribution after 2.2 m (a) and corresponding bottom pressure distribution (b) originating from an initial gas point source in bottom compartment 8 ($u_g=1.2$ m/s)

Fig. 4.39 Simulated gas flow distribution after 2.2 m (a) and corresponding bottom pressure distribution (b) originating from an initial gas point source in bottom compartment 15 ($u_g=1.3$ m/s)

Simulated results in Fig. 4.38 and 4.39 with the corresponding measured results of MONTZ-PAK for a bed height of 2.2 m (Fig. 4.32 to 4.35) we see that there is a good agreement for both the outlet flow distribution and the inlet pressure distribution.

4.3.5 CONCLUDING REMARKS

From the measurements of the large scale tracer and gas flow distributions we saw that there is a large difference between a tracer distribution in a uniform flow field and distribution resulting from a gas point source causing a horizontal pressure gradient. Further we saw that the gas spreading coefficient is hardly dependent on the superficial gas velocity. Only at very low channel Reynolds numbers ($<1100$), the spreading coefficient increases. For MONTZ-PAK the gas spreading coefficient was found to be independent of a counter current liquid flow. However for RALU-PAK we found an increase of the spreading coefficient with the liquid load, because the
liquid blocks the gas flow paths through the slits which means that more gas is forced to flow in the channel direction.

To combine the flow and mixing phenomena we developed a model in which the cell size is determined by the tracer spreading coefficient and the gas flow distribution by the balance equations (mass, momentum and energy) from Section 2.3 applied to larger cells. The cell pressure loss coefficients can be taken proportional to the cell size, which gives good results. A disadvantage of the developed model, which is a problem for all flow simulators, is the convergence in case of complex flow fields.

According to Süss [6], the initial gas profile can be estimated using common flow simulation programs.

In Chapter 8 we will show that the model can be used for simulations of practical columns.

NOTATION

SYMBOLS

\( D_r \)  Lateral spreading coefficient  \( m \)
\( h_{\text{cell}} \)  Cell height (dimension) of square cell  \( m \)
\( L \)  Length over which liquid is collected  \( m \)
\( s \)  Standard deviation  \( m \)
\( z \)  Bed height  \( m \)
\( S_t \)  Liquid splitting factor  -
REFERENCES


CHAPTER 5

WETTING CHARACTERISTICS OF STRUCTURED PACKINGS

5.1 INTRODUCTION

For the mass transfer it is extremely important to have a large interfacial area between the different phases because the mass transfer is proportional to this area. In columns containing structured packings the interfacial area is created by the wetted surface of the packing. To know how this wetted surface is created we first looked at the wetted surface on a flat plate with respectively the surface design of RALU-PAK 250 YC and MONTZ-PAK B1 250. These results are presented in Section 5.2.

To compare the wetting characteristics of flat plates with the wetting characteristics of columns containing structured packings, interfacial areas were measured in a column (ID 0.45 m) for both packings. The interfacial area was determined with the well known method of chemical absorption of CO₂ in a sodium hydroxide solution. In Section 5.3 a simple wetting model is presented which can predicts interfacial area in combination with the detailed liquid distribution model. The simulated results are compared with the measured ones.
5.2 WETTED SURFACE ON A FLAT PLATE

5.2.1 EXPERIMENTAL SET UP OF FLAT PLATE WETTING TESTS

To measure the wetted surface on a flat plate a liquid rivulet was supplied on a flat tilted plate. We measured the rivulet width on the plate for different liquid loads and plate slopes. The rivulet width was measured by eye using a ruler.

The influence of a counter current gas flow was investigated in a similar device which had a fixed inclination of 45°. Here we used a closed configuration to introduce a counter current gas flow.

We carried out the tests with a smooth plate and a flat RALU-PAK sheet both made of stainless steel 316. For MONTZ-PAK we only carried out the test with the regular MONTZ surface design (shallow embossed). This plate was fabricated out of stainless steel 304.

To see how the liquid properties will influence the wetting characteristics we used different liquids; demineralized water, water-glycol mixtures to study the influence of the viscosity and a water-ethanol mixture to study the influence of the

<table>
<thead>
<tr>
<th></th>
<th>Density [kg/m³]</th>
<th>Surface-tension [N/m]</th>
<th>Viscosity [Pa.s]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Demi water</td>
<td>1000</td>
<td>0.068</td>
<td>0.001</td>
</tr>
<tr>
<td>Water-50% ethanol (by vol.)</td>
<td>900</td>
<td>0.032</td>
<td>0.002</td>
</tr>
<tr>
<td>Water-50% glycol (by vol.)</td>
<td>1060</td>
<td>0.052</td>
<td>0.003</td>
</tr>
<tr>
<td>Water-90% glycol (by vol.)</td>
<td>1100</td>
<td>44</td>
<td>0.011</td>
</tr>
</tbody>
</table>

*Table 5.1* Liquid properties

<table>
<thead>
<tr>
<th></th>
<th>smooth plate ss 304</th>
<th>shallow embossed ss 316</th>
</tr>
</thead>
<tbody>
<tr>
<td>Demi water</td>
<td>~58</td>
<td>~43</td>
</tr>
<tr>
<td>Water-50% ethanol (by vol.)</td>
<td>~30</td>
<td>~30</td>
</tr>
<tr>
<td>Water-50% glycol (by vol.)</td>
<td>~55</td>
<td>~53</td>
</tr>
<tr>
<td>Water-90% glycol (by vol.)</td>
<td>-</td>
<td>~55</td>
</tr>
</tbody>
</table>

*Table 5.2* Measured contact angles
surface tension. The different liquid properties are given in Table 5.1. Table 5.2 gives the measured (by photographic method) contact angles for the different liquid properties and packing surfaces.

Before their use, the plates were cleaned with ethanol and thoroughly washed with demineralized water. Between all tests the plates were pre-wetted to avoid any history effects and to ensure comparable experimental conditions.

Fig. 5.1 shows that for MONTZ-PAK, in the case we do not pre-wet between tests, we will find a hysteresis effect.

\[ \text{Hysteresis effect} \]

\[ \begin{array}{c}
\text{rivulet width [mm]} \\
\text{liquid load [ml/s]} \\
\end{array} \]

\[ \text{fig.5.1 Hysteresis effect for a tilted plate under 45° a shallow embossed surface design (MONTZ-PAK)} \]

\[ \text{obtained with water} \]

5.2.2 EXPERIMENTAL RESULTS OF FLAT PLATE WETTING TESTS

In the literature [1..6] different models can be found that can predict the wetting characteristics of a single rivulet on a flat plate. However after evaluation of these models, they turned out to predict quite different relations between the rivulet width and the liquid load. Some of these models take the surface roughness into account, but none of them can predict the influence of the surface design. These shortcomings make existing models less useful for a study of the influence of the surface design on the wetting characteristics of a single rivulet. Because of this only a qualitative interpretation of our test results is possible.

Surface tension

Fig. 5.2 shows the rivulet width as function of the liquid load for water and a water-50%ethanol mixture for the three different surface designs all measured for tilted plates under 45°. The surface tensions and contact angles are given in Table 5.1 and 5.2 respectively. From these figures it is clear that, for all three designs, the film
Fig. 5.2 Rivulet width as function of the liquid load for tilted plates under 45° with a) smooth surface design b) a surface design with slits c) shallow embossed surface design obtained with water and a water-50%ethanol mixture.

Fig. 5.3 Rivulet width as function of the liquid load for tilted plates under 45° with a) smooth surface design b) a surface design with slits c) shallow embossed surface design obtained with water, a water-50%glycol and a water-90%glycol mixture.
width will increase with a decrease of the surface tension. This is conform the expectation, because a decrease of the surface tension will cause a less curved film surface due to a decrease of the contact angle.

Viscosity

The influence of the viscosity on the film width is given in Fig.5.3 for tilted plates under 45°. The properties of the water-glycol mixtures are given in Table 5.1. Table 5.2 gives the measured contact angles. The increase in rivulet width for the two designs with a closed surface can be explained from the fact that the film velocity will decrease with an increasing viscosity. For a certain flow rate this will result in an increase in rivulet cross section and so in an increase of the rivulet width. For regular RALU-PAK surface design with slits, it seems that viscosity has hardly any influence on the rivulet width. However, the higher viscosity will reduce the (film) velocity, which means that the amount of liquid that can pass through the fixed slit cross section will decrease. So for a certain flow rate not all the liquid can change side through the same slit, which means that both sides of the plate are wetted over a certain trajectory. Fig.5.4 gives a schematic flow trajectory for the regular RALU-PAK surface design with slits for 'low' and 'high' viscosity. This effect is incorporated in the detailed distribution model discussed in Section 2.2.

![Fig.5.4](image)

**Fig.5.4** Schematic trajectory of liquid flow for a surface design with slits: 'low' (a) and 'high' (b) viscosity

**Counter current gas flow**

Fig.5.5, shows the film width for two liquid loads as function of the gas velocity for flat plates with closed surface designs. As liquid we used the water-50%ethanol mixture and the plate was tilted under 45°. From this figure we may conclude that a
counter current gas flow will not influence the film width in the operation range of distillation columns containing structured packings with a specific surface of 250 m2/m3. Only at very high gas velocities we see that the film width will increase. Here the gas flow decelerates the rivulets, which reduces average film velocity. For a certain flow rate this will result in a larger film cross section and so to a larger film width.

**Packing configuration**

To investigate how the packing configuration will influence the wetting we did some tests with water, for different angles with the vertical from +45° to -45°.

**Fig.5.6** Rivulet width as function of the liquid load for plates under different slopes for a) smooth surface design b) shallow embossed surface design obtained with water
for the two designs with a closed surface. The results in Fig.5.6 show that in case that the film flows over flat tilted plate, the film width will increase with an increase of angle with the vertical. The explanation can be found from the fact that for less inclined plate the component of the gravitational force, that pulls the film downwards, will decrease, which results in a lower film velocity. On the other hand the component of the gravitational force perpendicular to the plate will increase, which results in a more flat cross section. Both effects will increase the film width. For a negative angle with the vertical the film flows downwards under the plate. In this case the two gravitational force components are equal to the situation described above. However, for a less inclined plate the increase of the gravitational force component perpendicular the plate will now increase the film thickness, which will result in a lower film width. So for negative angles with the vertical both effects work in opposite direction. From the fact that we see no influence of a negative inclination on the film width (Fig.5.6), we may conclude that both effects compensate each other.

Surface design

From Fig.5.2 to 5.5 we know that the packing surface design can influence the wetting characteristics of a single rivulet. This was also concluded by Mc Glamery [4]. However the differences in wetting behaviour are not always significant, which becomes clear if we compare the results in Fig.5.2a-c obtained with water for the three different surface designs.

5.3 INTERFACIAL AREA IN COLUMNS CONTAINING STRUCTURED PACKINGS

In the Section 5.2 we saw that many parameters can influence the rivulet width. Here we will answer the question whether all these parameters play a role in the prediction of the wetting characteristics of a column containing structured packing.

5.3.1 EXPERIMENTAL SET UP AND MEASURING METHOD

To measure the effective surface in a column containing structured packing, we used the well known method of chemical absorption of CO₂ in a sodium hydroxide solution. The assumption for the determination of the interfacial area is that the mass transfer rate between the gas phase and the liquid phase is totally determined by the reaction rate of the chemical reaction, which takes place in the liquid phase at the gas liquid interface. This reaction rate can be measured and modelled, which enables the determination of the gas liquid interfacial area. The reaction rate can be modelled by a second order reaction:
where $r$ is the reaction rate and $k_r$ the reaction rate constant of the second order reaction.

A detailed description of the calculation of the reaction rate $k_r$, taking into account the ion strength of the solution, is given by Pohorecki and Muniuk [7]. When the CO$_2$ transfer rate from gas phase to liquid phase is totally reaction controlled (Hatta number > 3), the mass transfer rate can be calculated from the reaction rate of the chemical reaction. The requirement that the Hatta number is above 3 is met when the pH is above 13. For these conditions the mass transfer coefficient can be calculated from:

$$k_i = \sqrt{k_r \, c_{i}^{OH^{-}} \, D_i^{CO_2}}$$  \hspace{1cm} (5.2)

where $k_i$ is the mass transfer coefficient, $D_i^{CO_2}$ is the diffusion coefficient for CO$_2$ and $c_{i}^{OH^{-}}$ the concentration OH$^-$ all in the liquid phase.

With the assumption that the CO$_2$ concentration in the bulk of the liquid is equal to zero (CO$_2$ will react with 2NaOH to Na$_2$CO$_3$ and H$_2$O) and the assumption of no gas phase resistance, we can calculate the average driving force (in the liquid phase) from the measured top and bottom CO$_2$ concentration (in the gas phase) with:

$$\Delta c_{\text{in}} = \frac{P}{H e} \frac{(c_{g, \text{bottom}} - c_{g, \text{top}})}{\ln\left(\frac{c_{g, \text{bottom}}}{c_{g, \text{top}}}\right)}$$  \hspace{1cm} (5.3)

The Henry coefficient ($H e$) is corrected for ion strengths of the sodium hydroxide carbonate mixture according to [8]. The total CO$_2$ mass transfer from the gas phase to the liquid phase can be calculated with:

$$N_{CO_2, \text{total}} = a_{\text{eff}} \, k_i \, \Delta c_{\text{in}}$$  \hspace{1cm} (5.4)

The total CO$_2$ mass transfer is equal to the amount of CO$_2$ which disappears out of the gas flow and follows from the CO$_2$ bottom- and top concentrations and the
total gas flow ($\Phi_g$) through the column. The interfacial surface ($a_{ef}$) can now be calculated with:

$$
a_{ef} = \frac{\Phi_g (c_g^{\text{bottom}} - c_g^{\text{top}})}{k_i \Delta c_{ie}}
$$

(5.5)

To avoid influence of increasing Na$_2$CO$_3$ concentration and decreasing pH on the reaction rate and the Henry coefficient, we started each series of tests with a fresh sodium hydroxide solution and a well defined pH.

The measurements presented here were carried out in a perspex column with a diameter of 0.45 m and a packing height of ± 0.8 m. The liquid was supplied by a gravity type narrow trough distributor with 100 drip pipes per square meter. The tests were carried out for different liquid loads to investigate the effects on the wetted surface. Before each test the column was flooded to avoid any hysteresis effect.

5.3.2 PROPOSED MODEL FOR EFFECTIVE WETTED AREA

The effective wetted area in a column containing structured packing depends on two parameters:

- The width of each rivulet
- The number of rivulets

If we assume that the relation between local liquid load and the rivulet width in a column is identical to that of a flat plate we can translate the data from Section 5.2 to a column containing structured packing, if we know the number of parallel rivulets and the liquid distribution over those rivulets. First we will assume that the number of parallel rivulets is equal to the number of packing channels and that we have an ideal liquid distribution. According to Section 3.2 the liquid distribution for MONTZ-PAK approaches this last assumption. Fig.5.7 shows the translation for the results obtained with water and the shallow embossed surface of MONTZ-PAK to the effective surface in a column.

The obtained effective wetted area seems to be much too small, especially if we realize that this surface will further reduce if some kind of (small scale) liquid maldistribution occurs. We can increase the calculated surface by assuming more than one rivulet per packing channel as is done by [9], or we can assume another relation for the rivulet width as function of the (local) liquid load. For reasons of simplification we have chosen for the latter option and assumed that a packing channel is totally wetted when liquid is present. The extra surface of the column wall
was neglected. This means that in case of an ideal liquid distribution, we will have a wetted area equal to the packing area, and that the decrease in wetted area is only a consequence of the absence of liquid in certain packing channels due to small scale liquid maldistribution. In our model therefore the wetted area is calculated with:

$$a_{\text{eff}} = n_{\text{cell, wet}} \cdot a_{\text{cell}}$$  \hspace{1cm} (5.6)

where $a_{\text{eff}}$ is the effective wetted area, $a_{\text{cell}}$ the packing area in a cell and $n_{\text{cell, wet}}$ is the number of cells were liquid is present which can be calculated with the detailed liquid distribution model. In Section 5.3.3 we will show that this assumption gives reasonable results for the experimental conditions we investigated.

5.3.3 EXPERIMENTAL RESULTS AND COMPARISON WITH MODEL CALCULATIONS

The experimental results for both RALU-PAK 250 YC and MONTZ-PAK B1 250 as function of the liquid loads are given in Fig.5.8. For both packings the actual measured area was about 25% larger than the packing area. Bornhütter and Mersmann [10] determined with the same measuring method also a larger interfacial area than expected from the packing area for random packing. According to these authors this is caused by the presence of drops in the packing. From observation we know that this explanation can not be true for structured packing operating below the loading point, because there are hardly any drops in the packing. From the fact that we measured a specific area which was independent of the packing height, the larger

![Graph showing percentage wetted area of packing area as function of the superficial liquid velocity obtained with water for a column packed with MONTZ-PAK translated from a flat plate experiment](image)

**Fig.5.7** Percentage wetted area of packing area as function of the superficial liquid velocity obtained with water for a column packed with MONTZ-PAK translated from a flat plate experiment

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specific area can not be explained by in- and outflow phenomena, because they should have a larger influence for short beds. The sodium hydroxide solution was frequently refreshed, so the extra measured surface can not be explained form a large accumulation of Na₂CO₃ or a large decrease of the pH. Also the fact that we did not totally meet the requirement of a negligible gas side mass transfer resistance can not explain the larger area, because additional gas side mass transfer resistance will lower the overall mass transfer rate, which will lead to even larger (calculated) area. So it is not totally clear whether the extra measured area is a consequence of a systematical deviation in the measuring method or that the extra area is really available in the column. That is the reason why we have set the maximum wetted area (measured for MONTZ-PAK B1 250) to 100%. From the graph in Fig.5.8 it is clear that at low liquid loads the wetted area for MONTZ-PAK is significantly larger than for RALU-PAK. If the wetted area for both packings is calculated assuming that a channel is totally wetted when liquid is present and using the detailed liquid distribution model, we find (Fig.5.9) that the large difference in wetted area comes from the difference in the extent of small scale liquid maldistribution. The decrease of the wetted area, as simulated for MONTZ-PAK at low liquid loads, is a consequence of the absence of the flow path via the crossings for this condition as earlier discussed in Section 2.2. After the redistribution between the packing layers the liquid enters the next packing element. At low liquid loads no liquid will flow via the crossings to the adjacent sheet in this packing element, which will cause more dry packing channels.

Fig.5.8 Percentage wetted area of maximum wetted area as function of the liquid load for both RALU-PAK and MONTZ-PAK measured by chemical absorption of CO₂ in a sodium hydroxide solution in a 0.45 m ID column and a bed height of ±0.8 m

Fig.5.9 Percentage wetted area of maximum wetted area as function of the liquid load for both RALU-PAK and MONTZ-PAK simulated for a 0.45 m ID column and a bed height of ±0.8 m and the model parameters for water
5.3.4 CONCLUDING REMARKS

Although there are a number of uncertainties in the experimental data we may conclude from the comparison of the wetting data obtained with water for both RALU-PAK and MONTZ-PAK for the same experimental conditions, that there is a significant difference in wetting behaviour.

The fact that the assumption of total wetting of a packing channel if liquid is available fits the experimental data, indicates that the wetting in structured packing is much better than we should expect from the wetting data of a single rivulet. Whether this is a consequence of a large number of small rivulets within a packing channel or due to a positive effect of the packing corrugations is not totally clear. It seems that at moderate to high liquid loads, the small scale liquid distribution and the film spreading within a packing channel are of equal importance for the creation of the gas-liquid interfacial area.

NOTATION

SYMBOLS

\begin{align*}
\text{a}_{\text{eff}} & : \text{Effective interfacial area} & \text{m}^2 \\
\text{a}_{\text{cell}} & : \text{Interfacial packing area in a cell} & \text{m}^2 \\
\text{c}_{\text{g,\text{top}}} & : \text{CO}_2 \text{ concentration in the gas phase at the top} & \text{mol/m}^3 \\
\text{c}_{\text{g,\text{bottom}}} & : \text{CO}_2 \text{ concentration in the gas phase at the bottom} & \text{mol/m}^3 \\
\text{c}_{\text{i}}^{\text{OH}^-} & : \text{OH}^- \text{ concentration in the liquid phase} & \text{mol/m}^3 \\
\text{D}_i^{\text{CO}_2} & : \text{Diffusion coefficient of CO}_2 \text{ in the liquid phase} & \text{m}^2/\text{s} \\
\Phi_g & : \text{Total gas flow through the column} & \text{m}^3/\text{s} \\
\text{He} & : \text{Henry coefficient} & \text{mol/m}^3 \text{.Pa} \\
\text{k}_r & : \text{Reaction rate constant of second order reaction} & \text{m}^3/\text{mol.s} \\
\text{k}_{\text{i}} & : \text{Mass transfer rate in the liquid phase} & \text{mol/s} \\
\text{n}_{\text{cell, wet}} & : \text{Number of cells in which liquid is present} & - \\
\text{r} & : \text{Reaction rate of second order reaction} & \text{mol/s} \\
\text{p} & : \text{Pressure} & \text{Pa}
\end{align*}
REFERENCES


CHAPTER 6

LIQUID DISTRIBUTOR PERFORMANCE MODELLING AND EVALUATION

6.1 INTRODUCTION

The malperformances of large diameter packed columns experienced in practice during the last decade are generally considered to be a consequence of a liquid maldistribution originating from an inadequate initial liquid distribution or redistribution [1]. The industrial scale tests discussed in Section 4.2 have clearly demonstrated that large diameter beds consisting of structured packings are practically incapable to restore initial liquid maldistribution.

Recognizing the importance of initial liquid distribution, the major packing manufacturers as well as some large production companies have put considerable effort in development and testing of liquid distributors. They also have developed their own methods for quantifying maldistribution, which is usually measured by taking samples from drip points. From time to time some of this knowledge is released to open literature, however, most of it is still strictly proprietary. For most
design and production engineers a liquid distributor seems to remain a black-box item.

Having in mind the importance of matching the distribution quality (number and lay-out of drip points) with the redistribution properties of different types and sizes of structured packings for obtaining optimal performances [2], we also have devoted some experimental and simulation effort to liquid distributors. Here a model is presented which enables the prediction of the initial liquid flow pattern, which can be expected from gravity type distributors disturbed in some way in their function.

6.2 LIQUID DISTRIBUTOR MODEL

6.2.1 EXPERIMENTAL SET-UP AND PROCEDURE

The experimental set-up consisted of a liquid circulation loop with a bypass comprising a pump, rotameters, and valves. The experimental procedure consisted of visual observation and simple but accurate time-volume measurement.

Schematic representation of the prototype of a narrow trough distributor with exchangeable drip pipes with different type of perforations is given in Fig.6.1. The distributor was made of plexiglass and the drip pipes of stainless steel. The holes have been made by means of a special (high voltage current utilising) tool and the slots and triangles by using common mechanical tools. Demineralized water, an aqueous solution of T-pol, a liquid detergent from SHELL which reduces the surface tension (and the contact-angle) of water with about 60%, were used as test liquid for single drip pipe tests. The measurements have been performed with increasing and decreasing level, to observe hysteresis effect, and they have been repeated several times to test reproducibility.

Distributor maldistribution tests have been carried out with four industrially fabricated distributors, with tap water as test liquid. The overall design was equal to that of the prototype shown in Fig.6.1, which means that each of the distributors

![Fig.6.1 Schematic representation of the distributor used for experimental purposes, with discharge systems tested](image-url)
Fig. 6.2 Details of trough and pipe design of distributors used in liquid distribution tests

comprised ten drip pipes with a square pitch (an equivalent to 100 drip points/m²). Fig. 6.2 shows in detail the dimensions of narrow trough and drip pipes with holes. As a basis for evaluation the so called maldistribution factor was used. The maldistribution factor ($M_f$) is defined as:

$$M_f = \frac{1}{n} \sum \left( \frac{Q_{\text{drip point}}^2 - Q_{\text{average}}^2}{Q_{\text{average}}^2} \right)$$  \hspace{1cm} (6.1)$$

where $n$ denotes the number of drip pipes, and $Q$ the liquid flow rate.

Imperfections built in during fabrication have been evaluated by a thorough inspection of fifteen available distributors of the same design manufactured by J. Montz. The following values are thought to be the maximum expectable deviations:

- Hole diameter $\pm 0.1 \ %$
- Distance between holes $\pm 1 \ mm$
- Distance from reference level $\pm 4 \ mm$

The maldistribution was measured simply by taking a number of samples (time/volume) from each of the drip points (pipes).
6.2.2 MODELLING OF THE PERFORMANCE OF GRAVITY DISTRIBUTORS

A survey of standard distributors and a detailed discussion of all aspects of distributor design and operation can be found in the book of Kister [3]. From the most recent brochures of packing manufacturers it can be seen that gravity distributors are preferred in conjunction with structured packings.

The principle of a gravity type distributor is fairly simple. The distributor consists of troughs or pans with perforations in the bottom or walls. The flow rate of discharging liquid is determined by the liquid velocity, form and the cross section area of openings. The velocity is a function of hydrostatic pressure of the liquid above the opening.

Gravity type distributors are commonly classified as orifice-type and weir-type. In the case of the orifice-type the holes placed in bottom plates are always totally submerged and the liquid level is related to the discharge velocity of liquid. The weir-type distributor with discharge openings in the sides of troughs or drip pipes works at the same principle but its operation can be adversely affected by the location of liquid level, i.e. insufficient submergence of openings.

The relation between velocity \( u \) and pressure for incompressible fluids is described by the reduced form of Bernoulli equation:

\[
g h + \frac{1}{2} u^2 = \text{constant}
\]  \hspace{1cm} (6.2)

where \( g \) denotes the gravity acceleration and \( h \) the liquid level.

From this equation we can calculate the liquid velocity in the perforation at a certain distance beneath the liquid surface (position \( x \)).

\[
u(x) = \sqrt{2 \ g \ h(x)}
\]  \hspace{1cm} (6.3)

By integration of this formula over the submerged cross section area \( A \) the liquid flow \( Q_{\text{sub}} \) through a perforation can be obtained.

\[
Q_{\text{sub}} = \int \sqrt{2 \ g \ h(x)} \ dA(x)
\]  \hspace{1cm} (6.4)

This will be discussed in more detail for three most common designs of discharge openings: holes, slots and triangles (V-notches), respectively.
Holes

If the holes are placed in the bottom then the distributor is classified as a standard orifice type design. Because of its sensitivity to plugging, this type of distributor is less used in conjunction with structured packings. Preferable is the design with narrow troughs, with holes placed in the walls on both sides of a trough, or in the walls of drip pipes. This is actually a combination of a weir type and a orifice type distributor, depending on the location of liquid level, which is usually above the holes, but can also be at the same height as holes. Here we will consider the second more troublesome option.

For holes the effective cross section at the level (height) \( x \) is described by:

\[
\frac{dA(x)}{dx} = 2 \sqrt{R_{\text{eff}}^2 - (R_{\text{eff}} - x)^2} \quad (6.5)
\]

where \( R_{\text{eff}} \) denotes the effective radius.

There is always a certain cross section contraction, depending on the form of the opening and fabrication tolerances. This is described by so called contraction factor \( Cf \). If we take this into account, the effective radius of a hole can be expressed as:

\[
R_{\text{eff}} = \sqrt{Cf} \cdot R_{\text{hole}} \quad (6.6)
\]

The distance beneath the surface \( h(x) \) is described by:

\[
h(x) = h_{\text{level}} - x \quad (6.7)
\]

If we combine Eqs (6.4) to (6.7) we can calculate the liquid flow rate at position \( x \) of the hole cross section. The integration over the submerged cross section gives:

\[
Q_{\text{hole}} = \int 2g \left( h_{\text{level}} - x \right) 2 \sqrt{R_{\text{eff}}^2 - (R_{\text{eff}} - x)^2} \ dx \quad (6.8)
\]

As shown in Fig.6.3 the integration boundaries are: 0 to 2 \( R_{\text{eff}} \), in the case that the hole is totally submerged, and 0 to \( h_{\text{level}} \) in the case of a partly submerged hole. This integral was solved numerically.
Fig. 6.3 Parameters used in the calculation of the outlet liquid flow through a partly and totally submerged hole.

Fig. 6.4 Parameters used in the force balance between the pressure, as consequence of the liquid head, and surface tension.

In the case that we have small holes and a liquid with a high surface tension, the pressure caused by the liquid level has to overcome the surface tension to enable the liquid to flow. In other words, the following condition must be satisfied:
\[ F_{\text{pressure}} \implies F_{\text{surface}} \quad (6.9) \]

where

\[ F_{\text{pressure}} = \rho \ g \ h(x) \sqrt{R_{\text{hole}}^2 - (R_{\text{hole}} - x)^2} \ dx \quad (6.10) \]

and

\[ F_{\text{surface}} = \sigma \cos \theta \ dx \frac{R_{\text{hole}}}{\sqrt{R_{\text{hole}}^2 - (R_{\text{hole}} - x)^2}} \quad (6.11) \]

This is shown in Fig.6.4.

**Slots**

Discharge through slots can be classified as a weir type system. The flow through a slot is simulated in the same way as for holes. Only the relationship between the discharge cross section area and the level position \( x \) is less complex.

The rounding at the lower end of the slot is a half circle with a radius which is one half of the effective width of the slot. So, we can use here the formula for a hole and integrate over the lower half of the hole only.

The flow through the straight part of the slot (\( Q_{\text{straight}} \)) has a constant width and is easily described by:

\[ Q_{\text{straight}} = \int \sqrt{2 \ g \ h(x)} \ W_{\text{effective}} \ dx \quad (6.12) \]

where \( W_{\text{effective}} \) denotes the effective width of a slot.

Fig.6.5 shows that the bounds of the integral are: 0 to \( h_{\text{level}} \).

Here we have also the influence of the surface tension, which is described in the same way as for a hole.
V-notches

A V-notch (triangle) is the same as a slot with an increasing width. The rounding at the lower part is described by the formula of a half hole. The liquid flow
rate through the part above the perforation can be described as:

\[ Q_{\text{triangle}} = \int \sqrt{2 \ g \ h(x)} \ (W_{\text{eff,triangle}}(1-\frac{x}{H_{\text{triangle}}})+W_{\text{eff,rounding}}) \ dx \]  \hspace{1cm} (6.13)

where \( H_{\text{triangle}} \) denotes the height of the V-notch, and \( W_{\text{eff,triangle}} \) and \( W_{\text{eff,rounding}} \) the effective top and bottom width of the V-notch respectively. The triangle is shown in Fig.6.6, which also show the bounds of the integral: 0 to \( h_{\text{level}} \). The influence of the surface tension is taken into account as in the case of holes.

**Sources of unlevelness**

To use these liquid discharge models to simulate real liquid distributors we have to know the effective liquid level for each drip point. The effective liquid level depends on: tolerances in distance from apertures in drip pipe to reference level due to fabrication irregularities, unlevelness of distributor itself, and the unlevelness caused by wave forming in troughs. The latter one can be avoided by baffling or by some other mechanical means.

### 6.2.3 RESULTS AND DISCUSSION

Fig.6.7 shows the measured and calculated liquid flow rate plotted against the liquid level for a drip pipe with holes. For the simulation, the corresponding

![Fig.6.7 Relationship between liquid level and liquid load for one drip pipe with holes](image)
Table 6.1  Used orifice contraction factors

<table>
<thead>
<tr>
<th>Cf hole</th>
<th>Cf slot</th>
<th>Cf V-notch</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.75</td>
<td>0.8</td>
<td>0.67</td>
</tr>
</tbody>
</table>

Fig. 6.8  Relationship between liquid level and liquid load for one drip pipe with a slot

Fig. 6.9  Relationship between liquid level and liquid load for one drip pipe with a V-notch (triangle)

experimentally found contraction factors (Table 6.1), a contact-angle of 30 degrees and the physical properties of water were used. The drip pipe was provided with three holes, the lowest one with a diameter of 4 and other two of 6 mm, respectively. Fig. 6.7 shows also the results for water with T-pol. It can be seen that in the case of
T-pol at low liquid levels we have more liquid discharge through the small hole.

Fig.6.8 shows the relationship between liquid flow rate and the liquid level as measured and calculated for water and a drip pipe with a slot with a width of 5 millimetre.

Fig.6.9 shows this relationship for a V-notch (triangle) with a base angle of 20 degrees.

From the comparison of measurements and simulations we can conclude that a weir type discharge requires less liquid height for the same liquid load than the orifice type discharge.

According to the achieved agreement between measurements and calculations we can conclude that good results can be obtained with theoretical relation between liquid level and discharge flow rate if we use contraction-factor as fit factor. The influence of the surface tension and contact angle is relatively small.

The simulated and measured maldistribution factors for the four test distributors with ten drip pipes are compared in Fig.6.10. From the simulation we know that the extreme peaks in the Mf value appear when the liquid level is within the diameter of a hole. As a consequence of the unlevelness at low liquid loads, the lowest holes of some drip pipes can be partly submerged or even above the liquid level. This can cause large differences in flow rates across the distributor. This effect decreases with an increasing number of submerged holes, and practically disappears when all available holes are completely submerged. In other words, when the liquid level is somewhere in between two vertically placed holes, the Mf value becomes negligibly small. The most sensitive situation for this kind of unlevelness occurs at low liquid loads, i.e. when only the holes closest to the bottom of the trough are active. Such a situation (only one row of holes or notches in the walls of troughs or drip pipes) commonly appears in practice.

![Graph showing maldistribution factors comparison](image)

**Fig.6.10** Measured and calculated effect of maldistribution due to an uneveness in the drip pipe (hole) level for a distributor with ten drip pipes for a trough with 10 drip pipes
**Fig. 6.11** Relationship between liquid level and liquid load (left) and the percentage of the deviation in outlet flow corresponding to the position of liquid level (right), for 2 drip pipes (with holes) and an out-of-levelness of 2 mm (a) and 5 mm (b).

**Fig. 6.12** Relative deviation of outlet flow as a function of liquid level, according to Sulzer [4,5], for two drip pipes with an out-of-levelness of 2 mm and 1 discharge hole (a), 4 discharge holes (b), and a slot (c).
In order to get more insight into the effect of unlevelness we used 2 drip pipes with an out-of-levelness of 2 and 5 millimetre, respectively. The measured and simulated results are given in Fig. 6.11a and b. These graphs show also the relative deviation of the outlet flow rate as function of the absolute liquid level. The conditions (drip pipe design and liquid properties) were the same as during tests with single drip pipes.

The researchers from Sulzer [4,5] achieved similar results (Fig. 6.12). Although not all of the parameters were known to us, we made a simulation which gave a fair reproduction of Sulzer's results (see Fig. 6.13). Obviously, a weir type liquid discharge is more sensitive to inclination than the orifice type. According to our experience, the weir type discharges are more suitable for achieving higher turndown ratios. Slotted drip pipes are the best solution for such purposes, because they are not easily plugged by fouling and enable high turndowns without using deeper troughs.

![Fig. 6.13 Reproduction of Sulzer's results (Fig. 6.12) by simulation](image)

### 6.3 PRACTICAL APPLICABILITY OF THE DISTRIBUTOR MODEL

The problems with initial maldistribution occur mostly in conjunction with large column diameters and/or low liquid loads. Special solutions are needed to handle very low liquid loads (below 0.5 mm/s or 2 m$^3$/m$^3$/h).

The model developed enables quantification of liquid flow rates through drip pipes for a given distributor. In other words, if we know the effective liquid level for each drip point, we can calculate the liquid throughput for each drip pipe (point). The differences in the individual flows determine the level of uniformity of liquid distribution, which can be expected in given situation. This means that the sensitivity to unlevelness of a design can be evaluated during design work, as well as the possible consequences of an unlevelness resulting from installation.

The model is flexible and can calculate the cross section area of discharge
openings in such a way that the total inlet flow of a trough is proportional to the number of drip points taking the number of feeds into account. This means that one or more stages of predistribution (parting boxes or troughs) can be included. The flow scheme of the model is given in Fig.6.14.

To illustrate how useful the proposed simulation model can be, we compared four different two-stage designs: narrow trough distributor with a parting trough, narrow trough distributor with a split parting trough (two in one length), a distributor with troughs in form of concentric rings with a crossed parting box, and a distributor with five pans and a central parting box. These hypothetical designs with a diameter of 2 m and 100 drip pipes/m² are schematically shown in Fig.6.15. In each drip pipe there are three holes of 4.5 mm placed vertically at a distance of 60 mm from each other. The drip pipes in the first (predistribution) stage are of the same design as the drip pipes in the second stage, only the diameters of the holes and pipes are larger. To simulate the effect of the fabrication tolerances, the maximum deviations listed in the table given earlier in the text were used.
Fig. 6.15 Schematic representation of four distributors designs evaluated by simulation: a) narrow trough distributor with a parting trough, b) narrow trough distributor with a split parting trough, c) a ring distributor with a crossed parting box and a multi-pan distributor with the square parting box.

Fig. 6.16a to d show the liquid profiles leaving different distributor types, all inclined 1 degree to horizontal axis, seen from the left front side to the right back side. The exhibited sensitivity to inclination of the ring distributor (Fig. 6.16c) is mostly due to the large length of the flow path (equal to column diameter) of the liquid. The large liquid concentration in the left back corner is caused by liquid discharging through the third vertically placed hole. For other drip pipes the liquid level is just below the third hole.

For the narrow trough distributor (Fig. 6.16a), only the maximum flow path length in the left to right direction is limited. In the back front direction the maximum travel distance is still the same as the column diameter. Therefore, a similar result is obtained as with the ring distributor.

For the split narrow trough distributor (Fig. 6.16b) the maximum liquid flow path length corresponds to one half of the column diameter. We see that this results in a substantially improved distribution, which means a better failing resistance with respect to unlevelness of distributor.

According to Fig. 6.16d a similar behaviour can be expected from a multi-pan distributor. However, such designs cope with the problem of limiting space for gas
Fig. 6.16 3-dimensional representation of the effect of distributor unlevelness (D = 2 m, incl. = 1°, \( u_a = 3 \) mm/s) on the outlet profile for the distributor designs shown in Fig. 6.15

Fig. 6.17 Effect of the superficial liquid velocity on the liquid outlet profile for an unlevelled multi-pan distributor (D = 2 m, incl. = 1°)
flow. A promising design has been introduced most recently by Nutter [6].

The effect of the liquid load on the distribution resulting from given inclination is illustrated in Fig.6.17a to d, for the multi-pan distributor. The three-dimensional graphs represent profiles corresponding to superficial liquid velocities of 0.5, 1, 2, and 4 mm/s, respectively. Fig.6.16d belongs also to this series; it represents in somewhat larger scale a profile corresponding to liquid velocity of 3 mm/s.

From these figures we may conclude that the sensitivity to unlevelness increases with an decreasing liquid load and an increasing length of liquid flow path, and depends on the position of the liquid level with respect to vertically placed holes.

Similar simulations can be performed for any type of liquid discharge system and lay-out of distributor. Calculations are iterative and become more tedious if predistribution is included. Increasing the diameter or the number of drip points (pipes) results in an increased calculation time, depending on the type (speed) of the PC used for this purpose.

6.4 CONCLUDING REMARKS

If properly fabricated and installed (levelled) the narrow trough distributors with drip pipes and holes can ensure uniform distribution over a wide range of liquid loads. Malperformance will be reduced to a minimum if holes are large enough and the liquid level is adjusted to be in between two rows of holes. However, for higher turn downs considerably high troughs (liquid levels) are needed. The drip pipes with slot enabled achieving equal liquid loads with approximately four time lower liquid level in troughs, but they proved to be more sensitive to unlevelness.

The capability to simulate distributors can result in reduced experimental effort and increased confidence. The developed software can be applied to the study and design of liquid distributors.

The tests with water are highly recommended to verify distributor performance before installation.
NOTATION

SYMBOLS

\[ \begin{align*}
\theta & \quad \text{Inclination angle} \\
\rho & \quad \text{Density} \quad \text{kg/m}^3 \\
\sigma & \quad \text{Surface tension} \quad \text{N/m} \\
A & \quad \text{Cross section area} \quad \text{m}^2 \\
C_f & \quad \text{Orifice contraction factor} \\
F & \quad \text{Force} \quad \text{N} \\
g & \quad \text{Gravity acceleration} \quad \text{m/s}^2 \\
H & \quad \text{Height of a V-notch (triangle)} \quad \text{m} \\
h & \quad \text{Liquid level} \quad \text{m} \\
M_f & \quad \text{Maldistribution factor} \\
n & \quad \text{Number of drip points (pipes)} \\
Q_{ab} & \quad \text{Liquid flow through submerged cross section} \quad \text{m}^3/\text{s} \\
R & \quad \text{Radius} \quad \text{m} \\
u & \quad \text{Superficial liquid velocity} \quad \text{m/s} \\
W & \quad \text{Width of slot or the top of a V-notch} \quad \text{m} \\
x & \quad \text{Liquid level position} \quad \text{m}
\end{align*} \]

INDICES

\[ \text{eff} \quad \text{Effective} \]

REFERENCES


7.1 INTRODUCTION

Until now we studied the hydraulics of beds consisting of structured packings. In this chapter we will see that liquid distribution and mixing can influence the separation efficiency of columns containing structured packing. The experimental data were obtained in a 0.45 m ID distillation column which is described in Section 7.2.

In Chapter 2 to 4 we discussed two different distribution models. One based on detailed information, the so called detailed model, and one based on overall information, the so called design scale model. In Section 7.3 we will discuss how these models can be used to calculate the separation performance of columns containing structured packings.

In Section 7.4 we will compare the calculated separation efficiencies with the
experimental data for both the detailed and design scale model. This gives us information whether knowledge about the flow behaviour is needed to predict the separation performance of a packed column.

7.2 EXPERIMENTAL SET UP

The experiments were carried out in 0.45 m ID distillation column. All experiments were carried out under total reflux conditions and atmospheric pressure with the test system methanol-ethanol, which has a relative volatility of about 1.7. Because the operation range of the column is limited to a gas load factor of 2 vPa we only investigated the separation performance below the loading point. The distributor used for all experiments was a Montz narrow trough gravity distributor with 16 drip pipes (100 drip pipes per square meter). The drip pipes were provided with 3 small vertically placed holes to ensure a good liquid distribution over the drip pipes for the total flow range. For both packings RALU-PAK and MONTZ-PAK we did a number of tests with different initial (mal)distributions:
- Uniform initial distribution
- Plugged drip pipes in the periphery zone
- Plugged drip pipes in the centre
- One side of the distributor plugged

The top and bottom compositions were measured from which the overall HETP was calculated with the Fenske equation. Besides the measurements to determine the HETP, we also measured the liquid composition at the bottom of the column for 14 points over the cross section; 7 points in the channel orientation and 7 points perpendicular to the channel orientation. This gave us information about the composition profiles over the cross section. The compositions were measured with a refractometer which was calibrated with samples analyzed with gas chromatography.

7.3 PACKING EFFICIENCY MODEL

The most simple approach to calculate the separation performance of a packed column is based on an equilibrium stage model. The height of a column needed for a certain separation can then be calculated by multiplying the required number of theoretical stages with the HETP (Height Equivalent to a Theoretical Plate).

Zuiderweg and Hock [1] introduced the basic HETP i.e. the minimum value of the HETP which can be obtained from a uniform gas and liquid distribution and which is in distillation relatively insensitive to the liquid-gas ratio. In the model we assume that in the different cells we have such an ideal situation, which means that in the case that the cell has the same height as the basic HETP, we can calculate the composition change over a single cell. However the cell height is normally smaller than the basic HETP. We assume that for these smaller cells the composition change
is a certain fraction of the composition change calculated by an equilibrium stage calculation. This fraction can be calculated from the ratio of cell height and basic HETP. This ratio can be seen as an overall efficiency ($E_{ov}$).

$$E_{ov} = \frac{N_{eb}}{n_{cell}} = \frac{h_{cell}}{HETP_{basic}}$$ (7.1)

where $N_{eb}$ is the number of equilibrium stages, $n_{cell}$ is the number of cell layers, $h_{cell}$ is the cell height (dimension) and $HETP_{basic}$ height of a theoretical stage in case of a uniform gas and liquid distribution.

As shown in Fig.7.1 the composition change over a layer with height $h_{cell}$ can be described with a cell ($E_{cell}$) or Murphree ($E_{mv}$, efficiency;

$$E_{cell} = E_{mv} = \frac{y_n - y_{n+1}}{y^* - y_{n+1}} \quad n=1 \text{ for top and } n_{cell} \text{ for bottom}$$ (7.2)

where $y$ is the vapour composition (superscript * means equilibrium composition) and $n$ is the cell layer number. In the case the operating line and the equilibrium line are straight lines the cell efficiency can be related to the overall efficiency by the relation of Lewis [2].

$$E_{cell} = E_{mv} = \frac{S^{E_{x}} - 1}{S - 1}$$ (7.3)
The stripping factor $S$ can be expressed as:

$$S = \frac{G}{L} K \quad (7.4)$$

where $G$ the gas (vapour) flow and $L$ the liquid flow (in moles) and $K$ the slope of the equilibrium curve.

![Diagram](image)

**Fig. 7.2** Iteration scheme of separation efficiency calculation with cell model
Knowing the relation between in- and outlet compositions for each cell we can calculate the composition change over the column by connecting the cells by the design scale or detailed distribution model as described before. For the detailed model we modeled the wall flow as a bypass flow, because we assume, as in Section 5.3, that the interfacial surface created by wall flow is negligible. The calculations are carried out from top to bottom. Because we do not know the inlet gas composition this composition has to be estimated from the previous calculation. The calculation is repeated until the calculated top and bottom compositions reach a constant value. The iteration scheme is shown in Fig. 7.2.

![Diagram](image)

**Fig. 7.2** Different initial liquid (mal)distributions used in distillation experiments a) uniform initial distribution b) periphery drip pipes plugged c) centre drip pipes plugged d) drip pipes at one half of the distributor plugged
7.4 RESULTS AND COMPARISONS

7.4.1 EXPERIMENTAL RESULTS

In our distillation column we did a number of tests for both RALU-PAK and MONTZ-PAK to compare the separation efficiency for a uniform initial distribution and three different non-uniform initial distributions. The different initial distributions are shown in Fig. 7.3. First the experimental results are discussed.

![graph](image)

**Fig.7.4** Experimental HETP as function of the F-factor for RALU-PAK and MONTZ-PAK with a uniform initial liquid distribution

**Uniform initial distribution**

Fig. 7.4 shows the results for both RALU-PAK and MONTZ-PAK for a uniform initial distribution. We repeated the experiments for MONTZ-PAK with a shorter packed bed to check the reproducibility of the low HETP value. From these experimental results it is clear that the HETP measured for MONTZ-PAK is about 50% lower than the one measured for RALU-PAK in case we have a uniform initial distribution. If we look at the outlet liquid composition profile taken over the column cross section obtained for RALU-PAK (Fig. 7.5a), we see that a higher methanol composition is measured in the wall zone. This indicates that there is relatively less mass transfer in the wall zone, which can be explained from a high liquid wall flow as it was found in our liquid distribution experiments. The outlet composition profile for MONTZ-PAK is given in Fig. 7.5b. This profile may be seen as flat.
Fig. 7.5 Experimental composition profiles measured at the bottom parallel (above) and perpendicular (below) to the sheet orientation with a uniform initial distribution for RALU-PAK (a) and MONTZ-PAK (b).

Fig. 7.6 Experimental HETP as function of the F-factor for RALU-PAK and MONTZ-PAK with an initial liquid maldistribution introduced by plugging the periphery drip pipes.
Plugged drip pipes in the periphery zone

Fig.7.6 shows that for both packings the HETP will not significantly change for the situation where the periphery zone is plugged. This indicates that the spreading of both packings is sufficiently large to restore from this initial liquid distribution. This is confirmed by the outlet liquid composition profiles measured over the column cross section, which have the same shape as the one measured with the uniform initial liquid distribution. The improvement of the HETP measured for RALU-PAK indicates that the initially under-irrigated periphery zone will postpone wall flow tendency, which results for a short trajectory in a better distributed liquid phase over the column cross section.

Plugged drip pipes in the centre

Fig.7.7 shows the HETPs for RALU-PAK and MONTZ-PAK measured for the situation with plugged drip pipes in the centre. From the comparison between Fig.7.5 and Fig.7.7 we can conclude that there is no increase of the HETPs due to this typical initial liquid maldistribution. We did also not find any significant difference in the outlet composition profile, in comparison with the one obtained with a uniform initial distribution. Both facts indicate that the packings investigated can restore from this initial liquid maldistribution.

![plugged centre](image.png)

**Fig.7.7** Experimental HETP as function of the F-factor for RALU-PAK and MONTZ-PAK with an initial liquid maldistribution introduced by plugging the centre drip pipes

One half of the distributor plugged

From the HETP values presented in Fig.7.8 we see a significant increase of the HETP value in the case that one half of the distributor is plugged. The difference between this initial liquid maldistribution and both initial maldistributions presented
**Fig. 7.8** Experimental HETP as a function of the F-factor for RALU-PAK and MONTZ-PAK with an initial liquid maldistribution introduced by plugging the drip pipes at one half of the distributor.

**Fig. 7.9** Experimental composition profiles measured at the bottom parallel (above) and perpendicular (below) to the sheet orientation with an initial maldistribution introduced by plugging the drip pipes at one half of the distributor for RALU-PAK (a) and MONTZ-PAK (b).
before is the distance over which the liquid has to be transported to recover from the initial liquid maldistribution. In case of a plugged centre or periphery the liquid has to be transported over only 0.10 m to restore to a uniform profile. In case that one half of the distributor is plugged the liquid has to be transported (in lateral direction) over more than 0.25 m. From the results in Fig.7.8 it is clear that both packings need a significant packing height to repair the initial liquid profile. The relatively smaller increase of the HETP measured for RALU-PAK indicates that this packing needs less height to establish it naturale flow pattern than MONTZ-PAK. The explanation for this difference in sensitivity for severe initial maldistribution can be found in the difference of lateral spreading behaviour as discussed in Section 2.2. That RALU-PAK can better recover from initial maldistribution than MONTZ-PAK is confirmed by the difference in the inclination of the outlet liquid composition profile presented for both packings in Fig.7.9a and b.

![Diagram](image)

**Fig.7.10** Different initial liquid (mal)distribution used for simulations with the detailed model a) uniform initial distribution b) periphery drip pipes plugged c) centre drip pipes plugged d) plugged drip pipes at one half of the distributor plugged
7.4.2 RESULTS OF SIMULATION WITH DETAILED MODEL

From Chapter 2 and 4 we know that the liquid flow behaviour for RALU-PAK and MONTZ-PAK is totally different. To check the influence of the small scale liquid spreading (mixing) mechanisms on the HETP we did some simulations with the detailed model under the same conditions as the experiments in Section 7.4.1. The different initial distributions are given in Fig.7.10. We determined the basic HETP on the basis of the experimental results obtained with a uniform initial liquid distribution. We further assumed that the liquid will close the slits in RALU-PAK as we showed in Section 2.3 and 4.3. This means that RALU-PAK will react as a totally closed packing with a lateral gas mixing behaviour similar to the lateral gas mixing behaviour of MONTZ-PAK. The gas distribution was assumed to be uniform.

For the liquid distribution calculation we used the splitting factors and minimum and maximum limitations obtained with the water-6%ethanol mixture giving in Table 2.1. Table 7.1 shows the used flow parameters and basic HETP for both RALU-PAK and MONTZ-PAK.

<table>
<thead>
<tr>
<th>Packing:</th>
<th>RALU-PAK</th>
<th>MONTZ-PAK</th>
</tr>
</thead>
<tbody>
<tr>
<td>$S_{l\text{ perforation}}$</td>
<td>0.95</td>
<td>-</td>
</tr>
<tr>
<td>$S_{l\text{ crossing}}$</td>
<td>-</td>
<td>0.65</td>
</tr>
<tr>
<td>$S_{l\text{ ridge}}$</td>
<td>-</td>
<td>0.15</td>
</tr>
<tr>
<td>$\text{Min}_{\text{perforation}}$</td>
<td>1 l/h</td>
<td>-</td>
</tr>
<tr>
<td>$\text{Min}_{\text{crossing}}$</td>
<td>-</td>
<td>0.6 l/h</td>
</tr>
<tr>
<td>$\text{Max}_{\text{perforation}}$</td>
<td>32 l/h</td>
<td>-</td>
</tr>
<tr>
<td>Periphery return factor</td>
<td>0.05</td>
<td>0.00</td>
</tr>
<tr>
<td>$\text{Max}_{\text{periphery}}$</td>
<td>24 l/h</td>
<td>17 l/h</td>
</tr>
<tr>
<td>Wiper leakage factor</td>
<td>0.6</td>
<td>0.6</td>
</tr>
<tr>
<td>$S_g$</td>
<td>0.15</td>
<td>0.15</td>
</tr>
<tr>
<td>Basic HETP</td>
<td>0.115 m</td>
<td>0.115 m</td>
</tr>
</tbody>
</table>

Table 7.1 Parameters detailed model
CHAPTER 7

Uniform initial distribution

Table 7.2 shows the experimental and simulated results for both RALU-PAK and MONTZ-PAK for the same conditions (F-factor=1.2). The basic HETP value was determined by fitting the model-HETP to the experimental HETP of MONTZ-PAK obtained with a uniform initial distribution. Interesting point here is that this basic HETP is much lower than the overall HETPs determined form top and bottom compositions, which indicates that the small scale liquid distribution in the packing is far from ideal. The fact that calculated overall HETP for RALU-PAK is much higher than the one obtained for MONTZ-PAK using the same basic HETP value, indicates that the small scale distribution, calculated for RALU-PAK, is less uniform.

<table>
<thead>
<tr>
<th>HETP for:</th>
<th>SIMULATION</th>
<th>EXPERIMENTAL</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>RALU-PAK</td>
<td>MONTZ-PAK</td>
</tr>
<tr>
<td>uniform initial distribution</td>
<td>0.29 m</td>
<td>0.20 m</td>
</tr>
<tr>
<td></td>
<td>RALU-PAK</td>
<td>MONTZ-PAK</td>
</tr>
<tr>
<td>periphery drip pipes plugged</td>
<td>0.31 m</td>
<td>0.20 m</td>
</tr>
<tr>
<td></td>
<td>RALU-PAK</td>
<td>MONTZ-PAK</td>
</tr>
<tr>
<td>centre drip pipes plugged</td>
<td>0.31 m</td>
<td>0.22 m</td>
</tr>
<tr>
<td></td>
<td>RALU-PAK</td>
<td>MONTZ-PAK</td>
</tr>
<tr>
<td>half of distributor plugged</td>
<td>0.40-0.41 m</td>
<td>0.35-0.36 m</td>
</tr>
<tr>
<td></td>
<td>RALU-PAK</td>
<td>MONTZ-PAK</td>
</tr>
</tbody>
</table>

Table 7.2 Calculated and measured HETP

Plugged drip pipes in the periphery zone

Table 7.2 shows that for the situation where the drip pipes in the periphery zone were plugged, the results of the simulation are in good agreement with the experiments. It is clear that the results for MONTZ-PAK are a slightly worse in comparison with the results of the uniform initial distribution. This can be explained because of an under-irrigated wall zone. The experimental results of RALU-PAK indicate a minor opposite effect. The simulated overall HETP for the case of plugged drip pipes in the periphery zone is a little bit higher than the case with the uniform initial distribution. In the simulation the positive effect of less wall flow is partly compensated by the under-irrigated periphery zone at the top of the column.

Plugged drip pipes in the centre

Table 7.2 shows that for MONTZ-PAK we find for both experiments and simulations a small increase of the HETP value in comparison with the HETP deter-
mined with the uniform initial distribution. This is explained by the over-irrigation of the wall zone, which affect gradually disappears in the packing. Although RALU-PAK has a much higher spreading we also see some effect of this initial maldistribution.

One half of the distributor plugged

A severe form of initial maldistribution is introduced by blanking one half of the distributor. For both packings we carried out three simulations with different redistribution patterns between the packing layer which is random to some extent. From the experimental results in Table 7.2 we see that for MONTZ-PAK the simulated results are a little bit too optimistic which probably means that the assumed spreading is too large. For RALU-PAK the calculated HETPs are in good agreement with the experimental results.

The variations in experimental HETPs at different loads shown in Fig.7.8 may be explained by different redistribution patterns between packing layers. However from the differences between the three simulated results we can conclude that the influence of the redistribution patterns is relatively small.

7.4.3 Results simulated with the design scale model

Here we will investigate whether we can predict the experimental data with the design scale model using rather small cells. The splitting factors for both the bulk and the wall are calculated from the experimentally determined lateral spreading coefficient, obtained in our large column simulator presented in Chapter 4. Because these spreading coefficients are only known for MONTZ-PAK, the simulations with the design scale model are limited to MONTZ-PAK. For the (pseudo) basic HETP we used the HETP value determined with the small scale model for a uniform initial liquid distribution. We will look whether the design scale model can predict the influence of an under-irrigated wall zone, an under-irrigated bulk zone and a half blanked distributor.

In order to simulate the separation performance of a column with 0.45 m ID using the design scale model the smallest allowable cell size of 0.067 m was chosen.

As mentioned before, we determined the splitting factors for the bulk for both gas and liquid phase from the lateral spreading coefficients as presented in Chapter 4. The splitting factors for the wall zone for both gas and liquid were taken half of the splitting factors of the bulk zone, which means that there are no wall effects. Table 7.3 shows the used splitting factors and (pseudo) basic HETP. Fig.7.11 shows the initial distribution used in the model calculation. From this figure it appears that we used in our model calculations a worse initial distribution than the one used in the experiments (Fig.7.3). The used drip point configuration was chosen to obtain the same lateral distance over which the liquid has to be transported to restore the initial distribution of a distributor with plugged drip pipes at one half. Fig.7.11d shows the
### Table 7.3 Parameters design scale model

<table>
<thead>
<tr>
<th>Packing:</th>
<th>MONTZ-PAK</th>
</tr>
</thead>
<tbody>
<tr>
<td>$S_{t \text{bulk}}$</td>
<td>0.87</td>
</tr>
<tr>
<td>$S_{t \text{wall}}$</td>
<td>0.43</td>
</tr>
<tr>
<td>$S_{g \text{bulk}}$</td>
<td>1</td>
</tr>
<tr>
<td>$S_{g \text{wall}}$</td>
<td>0.5</td>
</tr>
<tr>
<td>Pseudo Basic HETP</td>
<td>0.20 m</td>
</tr>
<tr>
<td>Cell size</td>
<td>0.067 m</td>
</tr>
<tr>
<td>Number of cells per layer</td>
<td>3</td>
</tr>
</tbody>
</table>

### Fig. 7.11 Different initial liquid (mat)distribution used for simulations with the design-scale model

- **a)** uniform initial distribution
- **b)** periphery drip pipes plugged
- **c)** centre drip pipes plugged
- **d)** plugged drip pipes at one half of the distributor plugged
model situation for a distributor with plugged drip points at one half which is comparable to the experimental situation shown in Fig.7.3d.

<table>
<thead>
<tr>
<th></th>
<th>MONTZ-PAK</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>SIMULATION</td>
</tr>
<tr>
<td>uniform</td>
<td>0.21 m</td>
</tr>
<tr>
<td>outside drip pipes</td>
<td>0.21 m</td>
</tr>
<tr>
<td>plugged</td>
<td></td>
</tr>
<tr>
<td>inside drip pipes</td>
<td>0.21 m</td>
</tr>
<tr>
<td>plugged</td>
<td></td>
</tr>
<tr>
<td>half of distributor</td>
<td>0.35 m</td>
</tr>
<tr>
<td>plugged</td>
<td></td>
</tr>
</tbody>
</table>

Table 7.4 Simulated and measured HETP with design scale model

Table 7.4 shows the results of the simulated HETPs for MONTZ-PAK for the four different initial distributions: uniform, plugged periphery, plugged centre and one half of the distributor plugged. For comparison we also gave the experimental results obtained under comparable conditions (F-factor=1.2). Although the model underpredicts the influence of the initial distribution on the HETP we may conclude in general that the results obtained for MONTZ-PAK with the design scale model are rather satisfactory. A possible explanation for the small under prediction of the HETP can be found in the applied splitting factors which are derived from spreading coefficients with water. Liquids with a lower surface tension exhibit a lower spreading at low liquid loads.

7.5 CONCLUDING REMARKS

If we compare the experimental results with the results predicted by the detailed model we may conclude that the model predicts the right trend. However for MONTZ-PAK with a poor initial distribution the simulated HETPs are a little bit too optimistic.

From the fact that there is a good agreement between the experimental and simulated HETPs, using the same basic HETP we can conclude that the big difference between RALU-PAK and MONTZ-PAK in overall HETP is mainly caused by the (small scale) liquid maldistribution. The large HETP found for RALU-PAK in comparison with MONTZ-PAK for the situation of a uniform initial distribution can be explained by a pronounced liquid wall flow and the absence of an intensive small scale mixing effect. The fact that RALU-PAK is relatively insensitive to the initial maldistribution can be explained by the large lateral spreading, as earlier concluded by Potthoff and Stichlmair [3].

From the comparison between the experimental results and the results obtained with the design scale model we can conclude that for packings that obey the
dispersion theory, we can predict the influence of an initial distribution from a spreading coefficient and a pseudo basic HETP. In this pseudo basic HETP the effects of the small scale maldistribution are inherently included, which means that this HETP can not be calculated directly from the mass transfer coefficient and the available interfacial area. For MONTZ-PAK the pseudo basic HETP is almost twice the basic HETP. So even for MONTZ-PAK with a relatively strong small scale mixing of liquid, the small scale liquid maldistribution decreases the packing efficiency.
NOTATION

SYMBOLS

$E_{\text{cell}}$  Cell efficiency  -
$E_{\text{m}}$  Murphree vapour efficiency  -
$E_{\text{ov}}$  Overall vapour efficiency  -
h_{\text{cel}}  Cell size (height)  m
HETP_{\text{basic}}  Basic HETP  m
G  Gas (vapour) flow  mol/s
K  Slope equilibrium curve  -
L  Liquid flow  mol/s
Min  Minimum liquid flow limit  l/h
Max  Maximum liquid flow limit  l/h
n_{\text{cel}}  Number of cell layers  -
N_{\text{th}}  Number of theoretical stages  -
S  Stripping factor  -
S_{\text{g}}  Gas splitting factor  -
S_{\text{l}}  Liquid splitting factor  -
y  Equilibrium vapour composition  mol/mol
y  Actual vapour composition  mol/mol

INDEX

n  Cell layer number ($n_{\text{top}}=1$, $n_{\text{bottom}}=n_{\text{cell}}$)

REFERENCES


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8.1 INTRODUCTION

In Chapter 2 to 7 we presented two models, the detailed and design scale distribution model. From the efficiency calculations with the detailed model we know that small scale liquid maldistribution will increase the HETP. However, for MONTZ-PAK we showed that the effect of small scale maldistribution can be incorporated in a pseudo basic HETP. This means that we can calculate the effects of large scale maldistribution with a design scale model without worrying about the small scale effects. For the calculation of industrial columns a simulation model with relatively large cells is preferred in order to achieve acceptable computer run time. In this chapter we will show some examples of realistic distribution problems in distillation columns. The separation efficiencies were calculated with the design scale model for a 1.4 ID column and a bed height of 2, 4 or 8 m. If not mentioned otherwise the simulations are carried out for total reflux conditions, a binary mixture
with a relative volatility of 1.5 and a uniform initial gas and liquid distribution.

For large columns, where the packing bed is built of packing blocks, the gaps between these packing blocks will introduce maldistribution. The simulated influence of liquid accumulation around gaps on the separation efficiency is presented in Section 8.2. We will also show that the simulated results depend on the cell size used in the design scale model.

In Section 8.3 we will show how the liquid accumulation around gaps can influence the separation efficiency for long packed beds and how we can reduce this influence by redistribution.

In Section 8.4 the influence of realistic initial gas and liquid maldistributions is discussed.

Finally in Section 8.5 we will investigate to which extent the relative volatility and reflux ratio will influence the separation efficiency.

### 8.2 EFFECT OF BED LAY-OUT ON THE SEPARATION EFFICIENCY

As discussed in Chapter 4 the spreading behaviour, calculated with the design scale model, is independent of the cell size if we adapt the splitting factors for the different cell sizes. However, because the liquid accumulation around a gap is very local, it is expected the cell size will influence the predicted separation efficiency. Here we will show the influence of the cell size on the separation efficiency as consequence of liquid accumulation around gaps.

![Diagram](image)

**Fig. 8.1** Gaps between packing blocks which are in line for layers with the same orientation
### Table 8.1 Distribution and mass transfer parameters for design scale model with two cell sizes

We assume that the packed bed is built of cells with a maximum length of 0.7 m which result in a gap, perpendicular to the channel orientation, in the middle of the column as shown in Fig. 8.1. The column was simulated with large cell dimensions (0.2 m) and with smaller cells (0.067). The simulation was carried out for 'normal operating conditions' and 'loading conditions'. The splitting factors for these conditions are given in Table 8.1. For all simulations we used a (pseudo) basic HETP of 0.2 m. The calculated HETP values are given in Table 8.2. From the simulation for the 'normal operating conditions' we see only a small influence of the gap for both cell sizes. Although the predicted HETP of the small cells is about 5% higher than the predicted HETP for the large cells we do not need to worry about the influence of the cell size.

Table 8.1 shows that under loading conditions the model predicts a significant increase of the HETP due to the large liquid accumulation around the gaps. Further we see that the predicted HETP for the small cells is higher than the HETP predicted for the large cells for both bed heights. In the model the accumulation of liquid around a gap is equally distributed over the cell cross section. In reality the

<table>
<thead>
<tr>
<th>HETP at:</th>
<th>normal conditions</th>
<th>loading conditions</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cell size</td>
<td>0.2 m</td>
<td>0.067 m</td>
</tr>
<tr>
<td>$S_t_{bulk}$</td>
<td>0.29</td>
<td>0.87</td>
</tr>
<tr>
<td>$S_t_{gap}$</td>
<td>0.24</td>
<td>0.42</td>
</tr>
<tr>
<td>$S_t_{wall}$</td>
<td>0.18</td>
<td>0.72</td>
</tr>
<tr>
<td>$S_g_{bulk}$</td>
<td>0.34</td>
<td>1.0</td>
</tr>
<tr>
<td>$S_g_{wall}$</td>
<td>0.17</td>
<td>0.5</td>
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<tr>
<td>pseudo basic HETP</td>
<td>0.2 m</td>
<td>0.2 m</td>
</tr>
</tbody>
</table>

### Table 8.2 Effect of gaps in line for two cell sizes and two bed heights

<table>
<thead>
<tr>
<th>HETP at:</th>
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<th>loading conditions</th>
</tr>
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<tbody>
<tr>
<td>Cell size</td>
<td>0.20 m</td>
<td>0.067 m</td>
</tr>
<tr>
<td>$z = 2$ m</td>
<td>0.204 m</td>
<td>0.211 m</td>
</tr>
<tr>
<td>$z = 4$ m</td>
<td>0.211 m</td>
<td>0.219 m</td>
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</table>

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accumulation around a gap is very local, so we may expect that the HETP predicted with the small cells is more representative for the HETP of a real column with a bed lay-out given in Fig.8.1.

As designers will not place all the gaps in line the bed lay-out from Fig.8.1. is not very realistic. Fig.8.2 shows a more realistic bed lay-out where the gaps stagger every four packing layers. In this case the accumulation of liquid takes place in a larger area. The predicted HETP is given in Table 8.3 for two bed heights and both the 'normal operating conditions' and the 'loading conditions' and the two cell sizes. When comparing the 'small cell' results of Table 8.3 with the corresponding result in Table 8.2 it can be concluded that the HETP can be improved approximately 15% by choosing a good bed lay-out. That no improvement is found for the large cells is a consequence of the fact that the staggering gap remains within the dimensions of the large cells. The comparison of the small and large cell results in Table 8.3 shows that we can use the large cells for the prediction of the HETP increase, due to liquid accumulation around gaps, when the accumulation takes place over a larger area, for example, due to gaps that stagger every four packing layers.

<table>
<thead>
<tr>
<th>HETP at:</th>
<th>normal conditions</th>
<th>loading conditions</th>
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<tbody>
<tr>
<td><strong>Cell size</strong></td>
<td><strong>0.20 m</strong></td>
<td><strong>0.067 m</strong></td>
</tr>
<tr>
<td><strong>z = 2 m</strong></td>
<td><strong>0.204 m</strong></td>
<td><strong>0.206 m</strong></td>
</tr>
<tr>
<td><strong>z = 4 m</strong></td>
<td><strong>0.211 m</strong></td>
<td><strong>0.212 m</strong></td>
</tr>
</tbody>
</table>

*Table 8.3  Effect of gaps that stagger every 4 packing layers for two cell sizes and two bed heights*
8.3 EFFECT OF LIQUID AND GAS REDISTRIBUTION

In practice long beds are mostly divided in a number of shorter beds in series with re-distributors between the beds. Here we will investigate whether and when re-distributors are necessary. For the simulations we used a total bed height of 8 m with the lay-out of Fig.8.2. In Section 8.2 we showed that, in the case we use a gap that stagger every four packing layers and a rather long bed, it is allowed to use large cell sizes (0.2 m). This will reduce the calculation time significantly. We simulated a bed of 8 m and two beds of 4 m in series with a redistributor between the beds. Here re-distribution means that all flow and concentration profiles are flat after a re-distribution step. The results for 'normal' and 'loading conditions' are given in Table 8.4. This table shows that for MONTZ-PAK at normal operating conditions, the long bed will not significantly influence the separation efficiency. However, under loading conditions i.e. a developing maldistribution over the bed height, a significant increase of the HETP is predicted by the model.

<table>
<thead>
<tr>
<th>Number of packed beds in series</th>
<th>1 x 8 m</th>
<th>2 x 4 m</th>
</tr>
</thead>
<tbody>
<tr>
<td>HETP for 'normal' conditions</td>
<td>0.217 m</td>
<td>0.209 m</td>
</tr>
<tr>
<td>HETP for 'loading' conditions</td>
<td>0.306 m</td>
<td>0.245 m</td>
</tr>
</tbody>
</table>

*Table 8.4 Effect of re-distribution for 8 m packed bed*

8.4 EFFECT OF INITIAL MALDISTRIBUTION ON SEPARATION EFFICIENCY

To study the influence of the initial gas and liquid maldistribution on the separation efficiency we used the column configuration of Section 8.2, with the small cells, a lay out of the gap as shown in Fig.8.2 and a bed height of 2 meter. To simulate the gas flow distribution with the design scale model we need channel flow, which means that we have to use the small cell configuration (cell size=0.067) as discussed in Chapter 4.

As initial gas maldistribution the profile shown in Fig.8.3a was used. The superficial gas velocity is 2.25 m/s. This initial gas distribution may be seen as an initial gas maldistribution as consequence of a poor design of the gas inlet pipe. Due to the high momentum of the gas a low velocity will exist directly above the gas inlet, and a high velocity near the opposite column wall of the gas inlet pipe. According to Süss [1] realistic estimations of the inlet gas profile can be made with computational flow simulation packages.

Fig.8.3 shows that, according to our design scale gas simulation model, the initial gas profile will equalize very fast. In Section 4.3 we saw that the gas velocity
Fig. 8.3 Initial gas maldistribution (a) and the corresponding gas distribution profiles after one (b) and two (c) packing layers

for the channels with a left side orientation and the gas velocity for the channels with a right side orientation can be different in one cell. For the calculation of the HETP we used the resulting total flow through a cell.

The calculated HETP for 'normal operating conditions' and 'loading conditions' are given in Table 8.5 for both the uniform and non uniform initial gas distribution.

<table>
<thead>
<tr>
<th>(Initial) gas distribution</th>
<th>ideal</th>
<th>nonideal</th>
</tr>
</thead>
<tbody>
<tr>
<td>HETP for 'normal' conditions</td>
<td>0.206 m</td>
<td>0.209 m</td>
</tr>
<tr>
<td>HETP for 'loading' conditions</td>
<td>0.225 m</td>
<td>0.227 m</td>
</tr>
</tbody>
</table>

Table 8.5 Effect of initial gas distribution
From the comparison of the HETP obtained for both initial gas distributions we may conclude that, according to the model, the initial gas distribution hardly influence the separation efficiency.

To study the influence of the initial liquid distribution we used the distributor simulation program presented in Chapter 6 to estimate the initial liquid profile. We simulated two gravity distributors with 100 and 60 drip pipes per square meter. Both distributors have two stages. The first stage consists of one pan which feeds the four pans in the second stage (Fig. 8.4). The distributor simulations were carried out with

![Two stage gravity distributor with four pans in the second stage (diameter = 1.4 m)](image)

**Fig. 8.4** Two stage gravity distributor with four pans in the second stage (diameter = 1.4 m)

![Simulated initial distribution for the distributor in Fig. 8.4 with 100 drip point per square meter placed under an inclination of 1 degree in the case of: a) drip pipes with two vertically placed holes b) drip pipes with slots](image)

**Fig. 8.5** Simulated initial distribution for the distributor in Fig. 8.4 with 100 drip point per square meter placed under an inclination of 1 degree in the case of: a) drip pipes with two vertically placed holes b) drip pipes with slots

![Simulated initial distribution for the distributor in Fig. 8.4 with 60 drip point per square meter placed under an inclination of 1 degree in the case of: a) drip pipes with two vertically placed holes b) drip pipes with slots](image)

**Fig. 8.6** Simulated initial distribution for the distributor in Fig. 8.4 with 60 drip point per square meter placed under an inclination of 1 degree in the case of: a) drip pipes with two vertically placed holes b) drip pipes with slots
two types of drip pipes with different discharge systems. One type with two vertical placed holes (d=4.5 mm) and one with slots of 4.5 mm. As we showed in Chapter 6, the drip pipes with holes are not very sensitive to unlevelness but need a rather high liquid head. The drip pipes with slots need a smaller liquid head, which means that they can be used over a larger liquid loading range within the same level variation. However, they are more sensitive to unlevelness especially at low liquid loads. All distributors were simulated under an inclination of one degree and a superficial liquid load of 1 mm/s, which is typical for vacuum distillation. The influence of the surface tension was neglected. Fig.8.5a and b show the results of the distributor simulations with 100 drip points per square meter for respectively drip pipes with holes and drip pipes with slots. Fig.8.6a and b shows the same situation only now we used 60 drip points per square meter.

<table>
<thead>
<tr>
<th>Drip points per square meter</th>
<th>±100/m²</th>
<th>±60/m²</th>
</tr>
</thead>
<tbody>
<tr>
<td>HETP for 'normal' conditions</td>
<td>'Ideal drip pipe'</td>
<td>0.216 m</td>
</tr>
<tr>
<td>drip pipes with 2 holes</td>
<td>0.226 m</td>
<td>0.254 m</td>
</tr>
<tr>
<td>drip pipes with slots</td>
<td>0.833 m</td>
<td>0.530 m</td>
</tr>
<tr>
<td>HETP for 'loading' conditions</td>
<td>'Ideal drip pipe'</td>
<td>0.234 m</td>
</tr>
<tr>
<td>drip pipes with 2 holes</td>
<td>0.245 m</td>
<td>0.272 m</td>
</tr>
<tr>
<td>drip pipes with slots</td>
<td>0.934 m</td>
<td>0.583 m</td>
</tr>
</tbody>
</table>

Table 8.6 Effect of initial liquid distribution

The simulated results for 'normal conditions' and 'loading conditions' are given in Table 8.6. The 'ideal drip pipe' gives the situation where the liquid was ideally distributed over the drip pipes. If we compare the results of HETPs obtained with a good initial liquid distribution, as simulated with the drip pipes with holes and a poor initial distribution simulated with the drip pipes with slots we see a significant increase of the calculated HETP due to poor initial liquid distribution. Especially for the loading conditions, where the packing is more sensitive to initial distribution due to the smaller liquid spreading behaviour. If we reduce the number of drip pipes from 100 to 60 drip pipes per square meter we see that in the case of a configuration with the drip pipes with holes we will loss separation efficiency. For the configuration of the drip pipes with slots the separation efficiency increases. In the latter case the negative effect of the reduction of the number of drip pipes is over-compensated through a better performance of the liquid distributor as consequence of the increase of liquid head (compare Fig.8.5b with Fig.8.6b).

For the same reason, an increase of liquid head as consequence of an increase in liquid load will increase the performance of the liquid distributor with slots.
8.5 EFFECT OF RELATIVE VOLATILITY AND REFLUX RATIO ON SEPARATION EFFICIENCY

Simulations were carried out to investigate the influence of maldistribution on the separation efficiency at different relative volatilities. The column configuration of Section 8.2 was used with small cells (cell size=0.067) and the bed lay-out of Fig.8.2. The splitting factors for 'loading conditions' were used to create a significant maldistribution. The results obtained for 2 m bed height are presented in Table 8.7, which shows that maldistribution has more influence on the separation efficiency at large volatilities.

<table>
<thead>
<tr>
<th>Relative volatility</th>
<th>1.2</th>
<th>1.5</th>
<th>2</th>
<th>4</th>
</tr>
</thead>
<tbody>
<tr>
<td>HETP 'loading' conditions</td>
<td>0.225m</td>
<td>0.225m</td>
<td>0.226m</td>
<td>0.230m</td>
</tr>
</tbody>
</table>

*Table 8.7 Effect of relative volatility on the sensitivity of the separation performance on maldistribution (loading condition cell size=0.067m)*

To investigate the influence of the reflux ratio on the separation performance we used the column configuration of two beds of 4 meter in series. A feed flow was introduced in the middle of the column. For the simulation a feed composition of 50 mol %, and two feed conditions (saturated liquid, q=1, and equimolar two phase mixture, q=0.5) were used. The top product (distillate) and bottom product (bottoms) were equal (in moles) and half of the feed flow. The results are given in Table 8.8. From these results we may conclude that in the case that the (mal)distribution and the basic HETP are independent of the load, the reflux ratio does not influence the overall HETP.

<table>
<thead>
<tr>
<th>REFLUX RATIO</th>
<th>3</th>
<th>6</th>
<th>∞</th>
</tr>
</thead>
<tbody>
<tr>
<td>q feed = 0.5</td>
<td>0.245m</td>
<td>0.246m</td>
<td>0.245m</td>
</tr>
<tr>
<td>q feed = 1</td>
<td>0.245m</td>
<td>0.245m</td>
<td>0.245m</td>
</tr>
</tbody>
</table>

*Table 8.8 Effect of reflux ratio on the sensitivity of the separation performance on maldistribution (loading condition cell size=0.2 m)*

8.6 CONCLUDING REMARKS

From calculations with the design scale model and different cell sizes we have seen that the negative effect of large local liquid accumulation on the HETP increases with a decrease of the cell size. Because in practice it is possible to avoid large local liquid accumulation (no gaps in line) we can use the design scale model
with large cells for the prediction of the HETP of large diameter columns.

According to the model it is not necessary to redistribute the liquid in the case we use MONTZ-PAK or any other packing with the similar distribution behaviour as long as the packing is operated below the loading range and no very low or high liquid loads are used provided that the bed height is below 8 m. In the case that the column is designed to operated in the loading range (i.e. strong deviation from uniform distribution) shorter beds should be considered.

Due to fast equalization of the initial gas distribution no large increase is predicted due to realistic initial gas(mal)distribution.

For gravity distributors with a large operating range used for low liquid loads a small installation error (1 degree inclination) results in a substantial increase of the HETP.

The sensitivity of the separation performance to maldistribution increases with the relative volatility. As earlier concluded by Yuan and Spiegel [2], in general the sensitivity of the separation performance for maldistribution decreases or remains the same at partial reflux conditions. Of course this does not mean that the reflux ratio does not influence the separation efficiency (HETP). An increase of the reflux ratio increases the liquid and gas load in the column which affect the hydraulics conditions in the column and thus the maldistribution.

NOTATION

SYMBOLS

S_l Liquid splitting factor -
S_g Gas splitting factor -
z Bed height m

REFERENCES


SAMENVATTING

Het doel van dit promotiewerk is enerzijds om de oorzaak en de gevolgen van klein- en grootschalige maldistributie voor gestructureerde golfplaat pakking te onderzoeken en anderzijds om een volledig voorspellend model voor de hydraulica en scheidingssefficiëntie voor industriële kolommen te ontwikkelen. Hiervoor hebben we twee gestructureerde pakkingen gekozen (MONTZ-PAK en RALU-PAK) die representatief zijn voor de twee extremen met betrekking tot het ontwerp van het pakkingsoppervlak. MONTZ-PAK heeft een geheel gesloten tamelijk ruw pakkingsoppervlak, terwijl het pakkingsoppervlak van RALU-PAK is voorzien van een open jaloeziesachtige structuur die makkelijk doorgang biedt voor de beide fasen.

Voor het modelleren van de gas- en vloeistofspreding hebben we een model met discrete cellen gebruikt. Bij het voorspellen van de kleinschalige maldistributie zijn de celafmetingen gekozen overeenkomstig met de hydraulische diameter van de geteste pakkingen. Met de berekende lokale vloeistof- en gasbelastingen kan de scheidingssefficiëntie worden voorspeld door middel van het berekenen van de lokale scheiding aan de hand van evenwicht.

De onderlinge uitwisseling tussen de cellen in het detaillereerde model bleek sterk afhankelijk te zijn van het ontwerp van het pakkingsoppervlak en werd onderzocht aan de hand van testen met punctbronnen die vloeistof introduceerden tussen twee platen gestructureerde pakking. Uit deze testen bleek dat voor RALU-PAK de vloeistof op de plaat waarop het geïntroduceerd was bleef en dat er een sterke uitwisseling was tussen de twee kanten van de plaat. Ten gevolge van het gesloten
pakkingsoppervlak van MONTZ-PAK is er geen uitwisseling mogelijk tussen de twee zijden van de plaat. Hier stroomt een deel van de vloeistof via de contactpunten naar de naastegelegen plaat met een tegenovergestelde oriëntatie, wat resulteert in een gelijkmatige verdeling over beide platen en een kleinere spreiding dan voor RALU-PAK.

Uit de testen met water (hoge oppervlakte spanning) hebben we kunnen vaststellen dat de spreiding toeneemt bij een afnemende vloeistofbelasting door de vorming van stroompjes die de kanaal richting volgen. Voor lage oppervlakte spanning bleek de spreiding minder afhankelijk te zijn van de vloeistofbelasting.

Voor beide pakkingen bleek de gas-verdeling uniform te blijven in geval van een uniforme beginverdeling. Voor het normale werkgebied waren resultaten van de spreidingstesten met een tracer vergelijkbaar voor beide pakkingen.

Het detaileerde verdelingsmodel kan gebruikt worden voor het voorspellen van de gas- en vloeistofverdeling in kleine kolommen en de invloed hiervan op het bevochtigd oppervlak en voor de scheidings efficiëntie.

De voorspelde vloeistofverdelingen zijn gecontroleerd aan de hand van experimentele data verkregen in een 0,5 m ID kolom met water. De vloeistof verdeling werd gemeten met behulp van 332 opvangbakjes (doorsnede = 2,5 x 2,5 cm) in de bodemsectie. De vloeistofverdeling die voor MONTZ-PAK werd gevonden bleek veel uniformer te zijn dan die gevonden voor RALU-PAK, vooral voor de lage vloeistofbelasting. Uit berekeningen met het gedetailleerde model bleek dat de niet uniforme vloeistofverdeling gevonden voor RALU-PAK, direct gerelateerd is aan grote vloeistof spreiding in deze pakking. In het algemeen kunnen we concluderen dat in het geval van een uniforme beginverdeling, de grotere spreiding een minder uniforme vloeistofstroom veroorzaakt.

Met een chemische absorptiemenode hebben we het bevochtigde pakkingsoppervlak voor water gemeten in een 0,45 m ID kolom. Het bevochtigde oppervlak bleek voor RALU-PAK ongeveer 0,7 keer dat van MONTZ-PAK te zijn, beiden gemeten bij een lage vloeistofbelasting (1 mm/s). Bij hoge vloeistofbelastingen (10 mm/s) was het bevochtigde oppervlak voor beide pakkingen ongeveer even groot. Uit berekeningen met het gedetailleerde model bleek dat het verschil in bevochtigingsgedrag kan worden voorspeld uit het verschil in de vloeistofverdeling.

De HETP (Hoogte Equivalent aan een Theoretische Plaat) is gemeten in een 0,45 m ID destillatiekolom, werkend onder totale reflux met het systeem methanol-ethanol. De HETP voor RALU-PAK (0,31 m) was ongeveer 1,5 keer zo groot als die van MONTZ-PAK (0,19 m), gemeten voor het normale werkgebied met een uniforme beginverdeling. Een half afgesloten verdeler levert een HETP van ongeveer 0,4 m voor beide pakkingen op. Het gedetailleerde model toont aan dat de scheidings efficiëntie voor beide pakkingen kan worden voorspeld op basis van de vloeistofverdeling en een gelijke waarde van de basis HETP.

De uitwisseling tussen de cellen in een model op ontwerp schaal hangt alleen af van de oriëntatie van de platen. De verdelingsparameters werden bepaald aan de hand
SAMENVATTING

van grootschalige spreidingsdata, gemeten voor een rechthoekige kolomsimulator (3 x 0.5 x 4.4 m³).

De gas- en vloeistofverdeling werd gemeten in 15 compartimenten, respectievelijk aan de boven- en onderkant van de kolom. Voor MONTZ-PAK vonden we een tamelijk egale, grootschalige vloeistofverdeling. Ten gevolge van de ophoping van vloeistof bij de spleten tussen de pakkingsblokken, werden kleine vloeistofpieken gevonden in het midden van de kolom. In het 'loading' gebied (maximale belasting) ontwikkelden deze pieken zich tot grote pieken van ongeveer 2 keer de gemiddelde vloeistofbelasting. Uit tracerverdelingsmetingen verkregen we een spreidingscoëfficiënt van ongeveer 0.012 m. voor het normale werkgebied. In het 'loading' gebied nam de spreidingscoëfficiënt af tot 0.006 m. De resultaten van RALU-PAK werden beïnvloed door een buitengewoon grote wandstroom ten gevolge van de kleine diepte (0.5 m) van de kolomsimulator, wat interpretatie van de spreidingsresultaten voor deze pakking moeilijk maakt.

Voor beide pakkingen bleef de gasverdeling uniform in geval van een uniforme gasbeginverdeling. Uit tracer(puntbron)testen werd voor het gas een spreidingscoëfficiënt gevonden van ongeveer 0.008 m voor RALU-PAK, die toeneemt met de vloeistofbelasting. Voor MONTZ-PAK hebben we een spreidingscoëfficiënt voor het gas van 0.017 m gevonden die onafhankelijk van de vloeistofbelasting bleek te zijn. Voor beide pakkingen neemt de gasspreiding toe bij gassenheden onder de 1 m/s.

Het effect van vloeistof-maldistributie op het scheidingsresultaat is voor het model op ontwerpschaal op een analoge wijze gemodelleerd als in het gedetailleerdemodel. De voorspellingen met beide modellen over de invloed van maldistributie op de scheidingscoëfficiënt komen vrij goed met elkaar en met de metingen overeen, wat wijst op de bruikbaarheid van het model op ontwerpschaal voor de simulatie van grootschalige industriële kolommen.

Rekening houdend met het belang van de beginverdeling van de vloeistof werd een verdelingsmodel ontwikkeld om realistische schattingen voor de beginverdeling te verkrijgen. Voor een enkel drippunt werd de voortgebrachte vloeistof als functie van het vloeistofniveau en het type opening bestudeerd en gemodelleerd. Op basis van het model voor een driphuis werd een model voor een twee lagen multi-compartiment verdeler ontwikkeld. Simulaties laten duidelijk de invloed van uitstroomenopeningen en de opdeling in compartimenten zien in geval van een niet 100% horizontale verdeler. Het model voor de verdeler kan worden samengevoegd met het model op ontwerpschaal om de invloed van de werking van verdeler op de scheidingscoëfficiënt te evalueren.