

MODELING AND MEASUREMENT OF MACROSCOPIC FLOW FIELDS IN STRUCTURED PACKINGS

B. Mahr, D. Mewes
University of Hanover, Germany Institute of Process Engineering Callinstrasse 36, D-30167 Hannover; E-mail: dms@ifv.uni-hannover.de

Corrugated structured packings feature strong preferential flow directions due to their structure of crossing triangular channels. This leads to good radial spreading but makes modelling a challenge. Detailed CFD calculations reflecting the exact packing structure are only feasible for small sections of packing. However, effects like large scale maldistribution and instabilities in the flow field can only be modelled, if the hydrodynamics of the entire column are taken into account. In the present study, the macroscopic flow field of an entire column is modelled and numerically calculated. The model is based on the elementary cell model by Mewes et al. (1999). It is extended to be used on anisotropic porous structures like corrugated structured packings in counter-current operation. For the elementary cell model, flow field variables (velocity, phase volume fraction, pressure, etc.) and packing properties (void fraction, pressure drop, momentum exerted on fluids, etc.) are averaged over the volume of a representative elementary cell. For the gas phase, measurements of the directional pressure drop are conducted and used to model the anisotropic gas flow resistance tensor. To model the liquid flow field, two liquid phases are modelled each representing laminar film flow along one preferential flow direction. The elementary cell model allows to determine the macroscopic flow field in columns of technical dimensions. In the present study, it is used to calculate the two-phase flow field under stationary, counter-current operating conditions below the loading point. The results are tested against X-ray radiographic measurements on a quasi two-dimensional segment of structured packing.

KEYWORDS: CFD, structured packing, modelling, two-phase flow, anisotropic pressure drop, liquid distribution, elementary cell model

INTRODUCTION

Structured packings have become increasingly popular in heat and mass transfer applications. Compared to conventional dumped packings, they feature lower pressure drop, higher separation efficiency, higher capacity and better radial mixing. Due to their structure of crossing channels, they show strong preferential flow directions for both liquid and gas flow. When operated in two-phase flow, these preferential flow directions and the low pressure drop cause structured packings to be more susceptible to maldistribution of the phases and instabilities in the flow field – which in turn leads to poor column performance. For equipment design, scale-up and operation it is therefore desirable to gain better understanding of the hydrodynamics in structured packings, and to be able to predict flow fields and operational ranges in those columns (Dudukovic et al., 1999; Spiegel and Meier, 2003).

There have been many approaches to modelling multiphase flow in packed columns. Jiang et al. (2002) calculate the stationary flow field based on the minimization of energy dissipation. A neural network approach was used by Iliuta et al. (1999) to model hydrodynamics and gas-liquid mass transfer in trickle bed reactors based on extensive measurement data from the open literature. Mewes et al. (1999) adapted the elementary cell model by Arbogast et al. (1990) to model two-phase flow in monoliths consisting of parallel channels. They solve volume-averaged mass and momentum equations to calculate the flow field. Recently, CFD has been used to calculate the local flow field through the exact geometry of triangular channels of corrugated structured packing (Petre et al., 2002) and packed beds of spheres (Nijemeisland and Dixon, 2004). Currently, this approach is only feasible for single phase flow through small segments of packed beds, though. Modelling two-phase flow at the scale of the exact packing geometry is not yet possible for columns of technical dimensions with today's computing power.

Effects like large scale maldistribution and instabilities in the flow field can only be modelled, if the hydrodynamics of the entire column are taken into account. In the present study, the macroscopic flow field of an entire column is modelled and numerically calculated. The model is based on the elementary cell model by Mewes et al. (1999), but extended to be used on anisotropic porous structures like corrugated structured packings in counter-current operation. It allows to calculate the two-phase flow field under stationary, counter-current operating conditions below the loading point. The results are tested against X-ray radiographic measurements on a quasi two-dimensional segment of structured packing.

ELEMENTARY CELL MODEL APPLIED TO STRUCTURED PACKINGS

Arbogast et al. (1990) developed a mathematical model to calculate flow in porous media. It is based on the idea that the entire porous media can be subdivided into elementary cells, over which properties like flow resistance and pressure drop are homogenized. Conservation equations for mass and momentum are solved for the elementary cells and the entire media. Thus, the macroscopic flow field can be described with values for pressure, velocities and phase fractions each averaged over the elementary cell volume.

Analogously applied to corrugated structured packings, an elementary cell is its smallest periodically repeating structure, which is shown schematically in Figure 1. Its interaction with the fluid can be characterized by the volume-averaged momentum that the packing exerts on the fluids. Assuming that the velocity varies much on a local scale, but is very similar for adjacent elementary cells, periodic boundary conditions can be used. If the flow field is furthermore stationary, and gravity can be neglected for the gas phase, the differential form of the moment conservation equations for volume-averaged values can be simplified and written for the gas phase

$$\overline{\alpha}_g \nabla \overline{p} = \underline{\underline{\pi}} \vec{j}_g + \vec{D}_{gl} \tag{1}$$

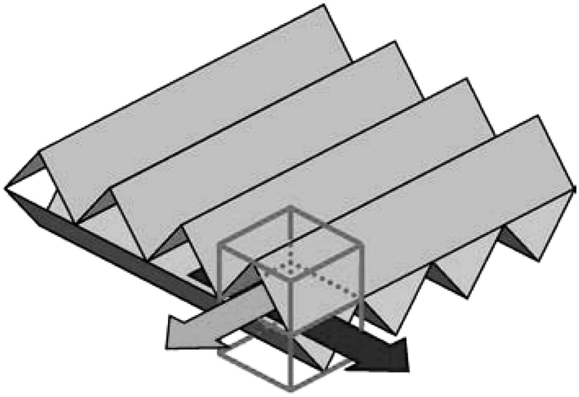

Figure 1. Schematic drawing of a segment of structured packing. The boundaries of an exemplary elementary cell and the preferential flow directions through it are indicated

and for the liquid phase

$$\overline{\alpha}_l \nabla \overline{p} = \underline{\underline{\xi}}\, \vec{j}_l\, \vec{D}_{gl} + \overline{\alpha}_l \rho_l \vec{g}, \tag{2}$$

where the indices l and g refer to the liquid and the gas phase, respectively, $\bar{\alpha}$ is the volume-averaged phase volume fraction, $\overline{p}$ the pressure averaged over the cell boundary, $\vec{j}$ is the superficial velocity, $\underline{\underline{\pi}}$ the gas flow resistance tensor, $\underline{\underline{\xi}}$ the liquid flow resistance tensor, $\vec{D}_{gl}$ the momentum transport between gas and liquid phase, ρ_l the liquid density and $\vec{g}$ the acceleration due to gravity. The conservation of mass for stationary conditions is given by

$$\nabla \vec{j}_g = 0 \text{ and } \nabla \vec{j}_l = 0. \tag{3}$$

A detailed derivation can be found in Mewes et al. (1999). The equations describe the macroscopic flow field with volume-averaged flow field variables. The model's advantage is that, since the packing's exact surface geometry does not have to be modelled, a relatively coarse mesh can be used, thus making calculation of technical scale columns feasible.

ANISOTROPIC GAS FLOW RESISTANCE

The momentum transport between the packing and the gas phase is described by the gas flow resistance tensor $\underline{\underline{\pi}}$ in equation (1). If no liquid is present, this is the dry pressure drop averaged over one elementary cell. It should be noted in Figure 1, that one elementary cell includes two adjacent sheets of packing, i.e. one pair of crossing channels. Due to this channel structure, the pressure drop within the packing is highly anisotropic: it depends not only on the gas velocity, but also on the flow direction of the gas. Along the channels, pressure drop is expected to be lower than along the vertical column axis, since the gas flow path is much more tortuous then. In order to measure the directional pressure drop within the packing sheet plane, narrow sections of packing material (Sulzer Mellapak 250.Y PP), several sheets of packing thick, are cut out in various angles, like indicated in Figure 2. These cut-outs are fit tightly into flow channel, and the pressure drop over the packing cut-out is measured at various air flow rates. Also, the same cut-outs are scanned in three-dimensionally in an X-ray CT, and a numerical calculation of single phase air flow through the three-dimensional structure is conducted. This approach opens up the possibility to obtain directional pressure drop data without actually having to cut packings manually. The results are shown in Figure 3. Both numerical results and measurements show excellent agreement. An angle of $\gamma = 0^\circ$ refers to the vertical column axis. As expected, the specific pressure drop along the column axis is higher than in direction of the channels. The corrugation angle of the packing and therefore the angle of the channels is 41° to the vertical. At an angle of around 90°, the pressure drop is even higher than at 0°. This is reasonable, since the channels form a narrower angle with each other, i.e. causing lower flow resistance, when gas is flowing in a vertical ($\gamma = 0^\circ$) direction. For all angles, pressure drop is found to grow with velocity to the power of 1.8. From these measurements, $\underline{\underline{\pi}}$ is interpolated. Calculation results reproduce spreading of point source gas flow very well.

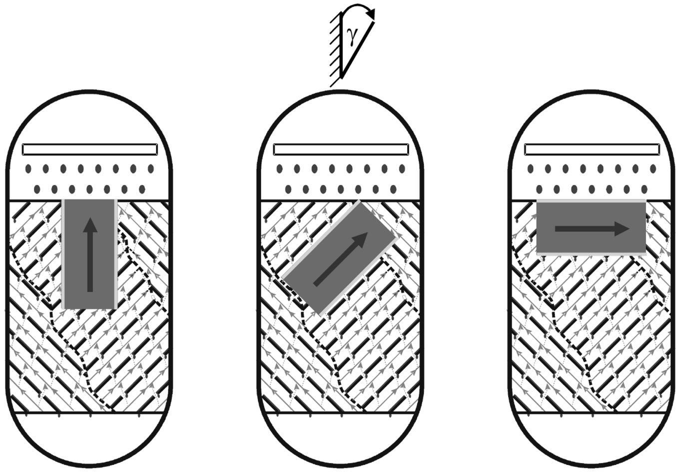

Figure 2. Orientation of cut-out packing sections

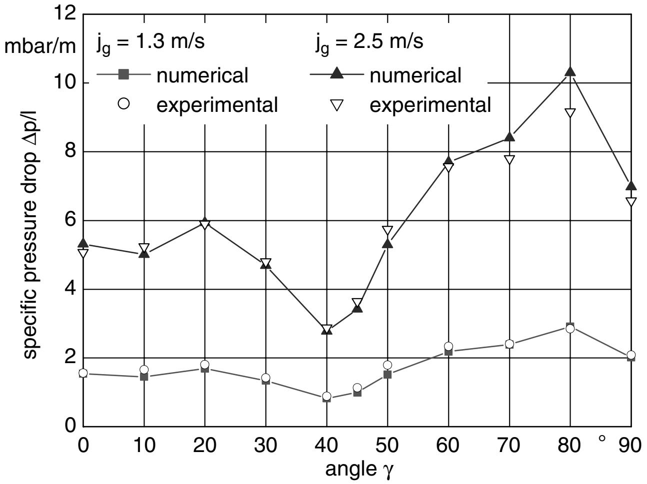

Figure 3. Anisotropic pressure drop of Mellapak250.Y

ANISOTROPIC LIQUID FILM FLOW RESISTANCE

For operating conditions below the loading point, gas-liquid momentum transfer is negligible, and for moderate liquid flow rates laminar film flow can be assumed. As mentioned before, one elementary cell includes gas and liquid flowing along both preferential flow directions. Assuming that liquid flow in both channels has about the same velocity under the same angle, i.e. symmetric with respect to the vertical column axis, averaging the liquid velocity over the elementary cell volume would always lead to liquid velocity vector pointing straight downwards. Liquid flow is only driven by gravity, which points in the same direction. Thus, the model would not be able to describe any radial liquid spreading. It should be noted that this is different for the gas phase, where the pressure gradient is the driving force, which can build up in any direction. To overcome the dilemma for the liquid phase, it is modelled by two liquid phases with mass transfer between them. Each liquid phase represents liquid flowing along one of the two preferential flow directions. The liquid momentum conservation equation (2) has to be written and solved twice with different flow resistance tensors $\underline{\underline{\xi}}$ then. The flow resistance tensor is derived from the laminar film flow model on an inclined plate (Figure 4). This allows to correlate the average film velocity v_l, film thickness δ, inclination angle γ, gravity g with the forces $F_{x,l}$ parallel and $F_{y,l}$ perpendicular to the plate: $F_{y,l}$ is expressed as a function of the weight of the liquid film, and $F_{x,l}$ is a function mainly of the average film velocity. From these equations, the flow resistance tensor is written.

Net mass exchange between the liquid phases is assumed to depend linearly on their difference in superficial velocities. The linear coefficient is chosen as 0.5 at the column

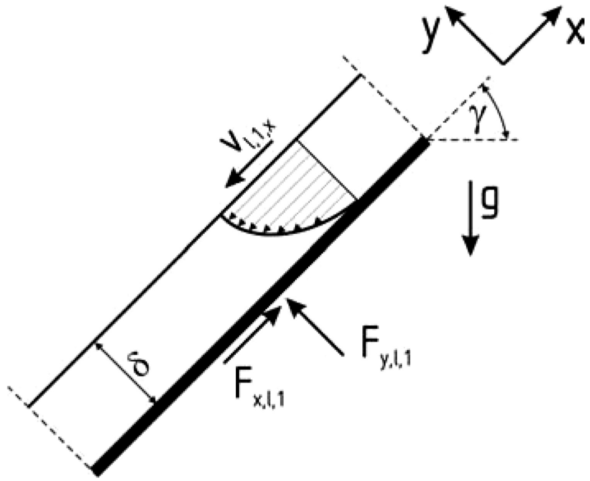

Figure 4. Laminar film flow model on an inclined plate

wall, so that the liquid that is transported towards the wall in one phase is forced to transfer to the other phase. This reflects the behaviour that occurs when the packing fits tightly into the column. For the rest of the packing, the coefficient is chosen to be around 0.05, which is a rough estimate and has to be verified by measurements. It is found, though, that the coefficient within the packing influences the flow field only moderately.

IMPLEMENTATION

The model is implemented in the commercial software package CFX 10.0. Three continuous phases are used to model one gas and two liquid phases. The corrugation angle is assumed to be 45°, and the specific packing surface is 250 m^2/m^3. One quasi two-dimensional mesh of 30 by 140 volume elements (representing a column of 288 mm width and 1030 mm height including 840 mm of packing) and one three-dimensional mesh of 67,000 elements (representing a column of 960 mm height with 288 mm inner diameter, containing 4 elements of packing, rotated against each other by 90°) have been used so far. Gas and liquid are introduced as mass sources within the column. They leave the column through outlet boundary conditions at the top, respectively bottom face of the column. Calculations are performed on a Dell PWS 650 PC with dual Xeon 2.8 GHz processors and on an IBM pSeries 690 supercomputer at the HLRN (High Performance Computing Network of Northern Germany), Hanover.

RESULTS AND DISCUSSION

Figure 5 shows numerical results of the two-dimensional column model. Liquid is fed as a point source just above the region that is filled with packing. The liquid superficial velocity

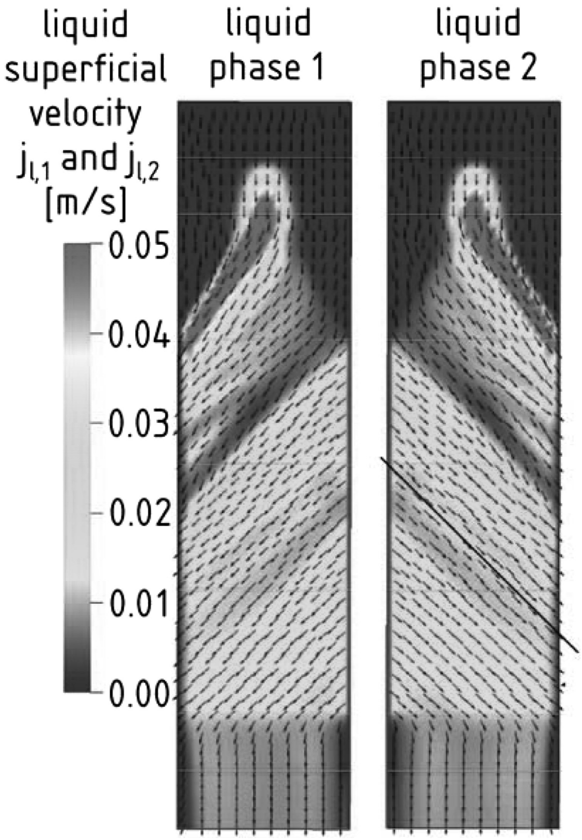

Figure 5. Superficial velocities of both liquid phases shown separately

is shown for both modelled fluid phases separately. It can be seen how the liquid of phase 1 is guided along the preferential flow directions to the left. When reaching the wall, all the liquid transfers to liquid phase 2, thus being led downwards to the right away from the wall. Towards the bottom of the packing, the liquid has spreaded evenly over the crosssection. After leaving the bottom end of the section filled with packing, the liquid return to a vertical flow direction, since it is accelerated by gravitational forces.

Figure 6 shows the liquid hold-up distribution of both liquid phases added up. As expected, the distribution is symmetrical to the vertical column axis. Two distribution patterns at the column head are compared: pointwise feeding in the left image and even irrigation in the right image. For both cases, the liquid superficial velocity is 100 m^3/m^2 and the counter-current gas flow rate is 0.45 m/s. In the lower part of the packing, roughly the

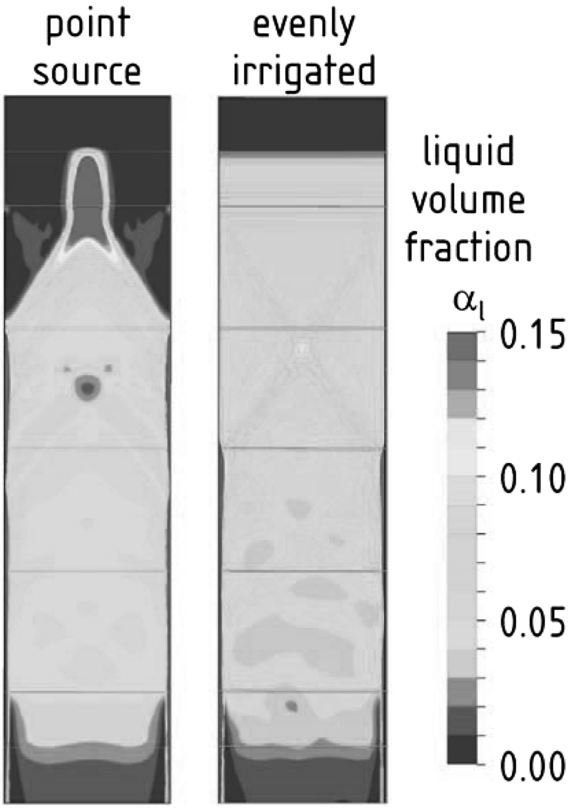

Figure 6. Distribution of total liquid hold-up for liquid fed from a point source (left) and evenly distributed (right)

same constant hold-up profiles are reached. In case of the point source feeding, at first more liquid is guided to the sides than straight down into the packing. This pattern is verified with experimental results obtained on a quasi two-dimensional test rig (Figure 7). Two sheets of packing material with sealed holes are irrigated in between on the inward facing surfaces. Thus, two liquid films form on both inner sides of the packing, presenting good comparability to the numerical model. The liquid film thickness is measured by X-ray radiography. It should be noted that the liquid superficial velocity in the experiment is lower than in the numerical calculation, since it could not be introduced as a liquid film at higher flow rates. There was no gas flow in the experiment. The conditions of the numerical calculation presented in Figure 7 is the same as for the results presented in Figure 6. The experiment shows how the liquid is spreading more to the sides

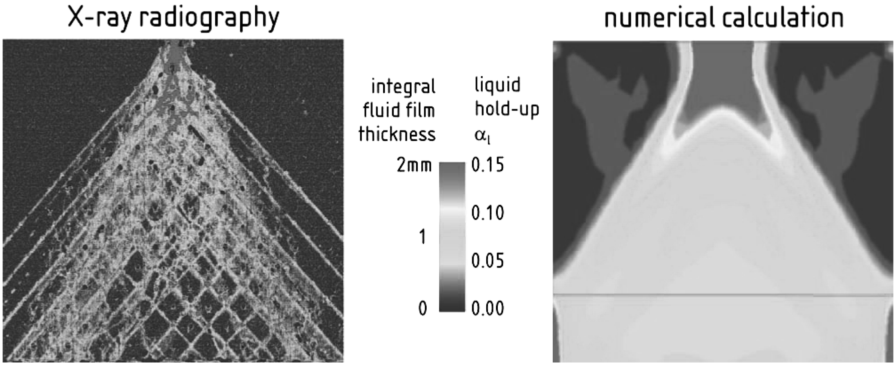

Figure 7. Comparison of X-ray radiographic film thickness measurement on a quasi two-dimensional segment of structured packing (left) and liquid hold-up distribution from numerical calculation. In both cases, liquid was fed from a point source

than straight down from the injection point. This trend is also reflected in the numerical results. The angle under which most of the liquid flows down the packing is steeper in the experiment than in the calculation. This is due to the assumption in the two-dimensional model, that the liquid is guided along the corrugation angle. In reality, the packing material, on which the liquid film forms, is inclined both along the corrugation angle and towards the adjacent sheet of packing. The effective three-dimensional inclination is therefore less than the assumed 45°. This should be incorporated in future three-dimensional column models.

CONCLUSIONS

The elementary cell model is extended to model two-phase flow through structured packings. The anisotropic flow resistance is modelled by a directional pressure drop for the gas phase. Two liquid phases are used to model radial liquid spreading based on a laminar film flow assumption. The liquid spreading pattern from a point source reflects experimental results well, although the assumption of a channel inclination angle of 45° leads to wider spreading than what experiments show. So far numerical results are very promising for reliably calculating macroscopic flow fields in structured packings. The model will be tested against more experimental data, liquid spreading perpendicular to the packing sheet plane has to be adequately accounted for, and flow fields in larger columns will be calculated. In the future, an extension of the model to operating conditions in the loading regime and transient flow fields is planned.

ACKNOWLEDGEMENTS

The authors would like to thank the HLRN (High Performance Computing Network of Northern Germany) in Hanover for their kind support and making the computational facilities available (project code nii00023). The packing for experimental work has been kindly provided by Sulzer Chemtech (Winterthur, Switzerland).

REFERENCES

Arbogast, T., Douglas, J., and Hornung, U., 1990, Derivation of the double porosity model of single phase flow via homogenization theory, SIAM J. Math. Anal., *21*, 823–836.

Dudukovic, M. P., Larachi, F., and Mills, P. L., 1999, Multiphase reactors – revisited, Chem. Eng. Sci., *54*, 1975–1995.

Iliuta, I., Ortiz-Arroyo, A., Larachi, F., Grandjean, B. P. A., and Wild, G., 1999, Hydrodynamics and mass transfer in trickle-bed reactors: an overview, Chem. Eng. Sci., *54*, 5329–5337.

Jiang, Y., Khadilkar, M. R., Al-Dahhan, M. H., and Dudukovic, M. P., 2002, CFD modeling of multiphase flow in packed-bed reactors: I. k-fluid modeling issue II. Results and applications, AIChE J., *48*, 701–730.

Mewes, D., Loser, T., and Millies, M., 1999, Modelling of two-phase flow in packings and monoliths, Chem. Eng. Sci., *54*, 4729–4747.

Nijemeisland, M. and Dixon, A. G., 2004, CFD Study of Fluid Flow and Wall Heat Transfer in a Fixed Bed of Spheres, AIChE Journal, *50*, 906–921.

Petre, C. F., Larachi, F., Iliuta, I., and Grandjean, B. P. A., 2002, Pressure drop through structured packings: Breakdown into the contributing mechanisms by CFD modeling, Chem. Eng. Sci., *58*, 163–177.

Spiegel, L. and Meier, W., 2003, Distillation columns with structured packings in the next decade, Chem. Eng. Res. Des., *81*, 39–47.