Performance Map for the Design of Liquid-Liquid Extraction Columns

Benedikt Weber, Christian Meyer, and Andreas Jupke

DOI: 10.1002/cite.201900057

This is an open access article under the terms of the Creative Commons Attribution License, which permits use, distribution and reproduction in any medium, provided the original work is properly cited.

Supporting Information available online

Dedicated to Prof. Dr. techn. Hans-Jörg Bart on the occasion of his 65th birthday

For the fast approximation of liquid-liquid extraction column designs via HETS (height equivalent of a theoretical stage) values and flooding points, a population balance model was developed. The model requires only one pilot-scale column experiment to fit a coalescence parameter. The comparison of experimental data with simulations shows good agreement for a stirred extraction column. With the model, the operation domain of the column is simulated and visualized in a performance map, showing the holdup, the distance to the flooding point and HETS values.

Keywords: Kühni column, Liquid-liquid extraction, Operation domain, Population balances, Stirred column

Received: March 14, 2019; revised: July 27, 2019; accepted: August 01, 2019

1 Introduction

Liquid-liquid extraction columns are often designed by applying the HTU-NTU (height of transfer units – number of transfer units) concept or by stage-to-stage models in combination with pilot-plant experiments [1]. In the past years, single drop based methods were developed. They reduce the number of expensive pilot-plant experiments needed for the design of extraction columns by single drop experiments [2]. The single drop rise velocity and mass transfer are determined separately and model parameters can be derived for each chemical system. All significant phenomena that influence the performance of liquid-liquid extraction columns like coalescence, drop breakage due to energy dissipation, mass transfer and the drop movement can be calculated with independent submodels [3 – 6]. Drop movement results from the interaction of the single drop rise velocity, swarm effects and drop backmixing induced by vortexes from internals like stirrers. The combination of these submodels for the above-mentioned phenomena allows the simulative evaluation of the separation efficiency for different column types, geometries and operation parameters. Consequently, the optimal column design and operation parameters can be determined [1, 2].

For the drop-based modeling of extraction columns, population balances are applied [3, 6, 7]. In the present study, a Monte-Carlo method is used to solve the population balance [6, 8]. With the Monte-Carlo method a representative number of drops is simulated [3, 6]. For each drop its velocity as well as mass transfer are calculated. Additionally, the probabilities for drop breakage and coalescence are determined for each drop [3].

The parameters for the submodels for the rise velocity and the mass transfer between the continuous and the disperse phase are typically fitted to single drop experiments [3]. Therefore, standardized single drop measuring cells were developed [3, 4]. Drop breakage, swarm effects and drop backmixing can be modeled without further parameter fitting. Only coalescence requires additional experiments [9].

In many cases, an approximation of HETS (height equivalent of a theoretical stage) values and flooding points of liquid-liquid extraction columns is desirable only on the basis of physico-chemical properties of the chemical system. Therefore, based on the modeling framework “ReDrop” [3, 6], a one-dimensional population balance model was developed, which does not require single drop experiments for different chemical systems. Instead, universal models and correlations are applied, reducing the experimental effort significantly. A generalized model that approximates the rise velocity of single drops is developed. This model is evaluated with experimental data from typical extraction systems from literature. Mass transfer is considered by

Benedikt Weber, Christian Meyer, Prof. Dr.-Ing. Andreas Jupke
anders.jupke@avt.rwth-aachen.de
RWTH Aachen University, AVT – Fluid Process Engineering,
Forckenbeckstraße 51, 52074 Aachen, Germany.

The authors contributed equally to this manuscript.
applying the model of Kumar and Hartland [10], which is based on 491 measurements. However, up to now coalescence phenomena are difficult to predict since they are influenced by numerous factors. For example, small amounts of surface-active components can greatly affect the coalescence efficiency [11]. Consequently, one pilot-plant experiment is required to fit the coalescence parameter [7,9].

The population balance model is used to predict the operation domain of a stirred extraction column. The resulting performance map (see Fig.5 and 6), depicts HETS values, the holdup and the distance to the flooding point depending on the specific loads of the continuous and disperse phase. Performance maps allow a quick and easy approximation of the column design (diameter and height) for a given separation task (volume flow ratio and number of necessary equilibrium stages). An application of the performance map is shown in this publication.

First, the extraction column and chemical system that are used in this study for model validation are presented (Sect. 2.1 and 2.2), followed by a detailed introduction of the applied submodels (Sect. 2.3) and the simulation settings (Sect. 2.4). Afterwards, the results of the developed single drop rise velocity model are compared with published experiments (Sect. 3.1). In Sect. 3.2, the coalescence parameter is presented. Subsequently, the simulation results are compared with column experiments (Sect. 3.3) and the performance maps for two operation points are described and compared (Sect. 3.4). The application of the performance map for the column design is presented in Sect. 3.5. Finally, a conclusion is given in Sect. 4.

### 2 Materials and Methods

#### 2.1 Equipment

Garthe [5] conducted several liquid-liquid extraction column experiments in pilot-plant scale. The investigated DN 80 column (type Kühni) is simulated in this study. Tab. 1 summarizes the geometry of the extraction column.

| Variable                  | Value |
|---------------------------|-------|
| Column diameter $d_{col}$ [mm] | 80    |
| Column height $h_{col}$ [m]       | 2.95  |
| Compartment height $h_{comp}$ [mm] | 50    |
| Rotor diameter $d_r$ [mm]  | 45    |
| Rotor height $h_r$ [mm]    | 7     |
| Number of rotor blades $n_{rd}$ [-] | 6     |
| Shaft diameter $d_h$ [mm]  | 10    |
| Free cross section of stator $\phi$ [%] | 40    |

#### 2.2 Chemicals and Operation Parameters

In this study, a biphasic standard test system [12] is used with water as continuous phase, toluene as dispersed solvent and acetone as dissolved transfer component. In Tab. 2, the chemical properties are listed, which were measured by Garthe [5] at two different concentrations. In the simulations, a linear interpolation between these concentrations is used for all physical properties of Tab. 2. For the distribution coefficient ($K = \gamma_{\text{Acetone}}/\gamma_{\text{Acetone}}$), the following interpolation fitted to the experimental data of Garthe [5] is applied: $K = 0.6442 + 2.6814x_{\text{Acetone}} - 5.9026x_{\text{Acetone}}^2$ with $x_{\text{Acetone}}$ and $\gamma_{\text{Acetone}}$ being the mass concentration of the transfer component in the continuous and disperse phase. The diffusion coefficient $D$ of acetone is $2.788 \cdot 10^{-9}$ m$^2$s$^{-1}$ in toluene and $1.152 \cdot 10^{-9}$ m$^2$s$^{-1}$ in water [12].

| Property                  | $x_{\text{Acetone}}$ [wt %] |
|---------------------------|-------------------------------|
| $\rho_{\text{aq}}$ [kg m$^{-3}$] | 998.8                        |
| $\rho_{\text{org}}$ [kg m$^{-3}$] | 867.5                        |
| $\eta_{\text{aq}}$ [Pa s]   | 0.001029                     |
| $\eta_{\text{org}}$ [Pa s]  | 0.000596                     |
| $\sigma$ [N m$^{-1}$]      | 0.03431                      |

Garthe [5] investigated six different operation points. Those operation points are listed in Tab. S1 in the Supporting Information (SI) and are used to evaluate the accuracy of the simulation model.

#### 2.3 Submodels for Drop Motion, Mass Transfer, Coalescence and Breakage

The velocity of rising drops strongly depends on the mobility of the drop interface and its shape. Drops are slower when the interface is rigid or the drop shape is deformed. When the interface is mobile, the drops rise faster. The interface mobility depends on the size of the drop, its physico-chemical properties, the accumulation of surface-active components at the interface and mass transfer [3, 6, 13, 14]. Henschke developed a drop rise velocity model that accounts for mobile, deformed and rigid drop interfaces, but requires single drop experiments to fit three model parameters [6]. In this publication, a generalized model approach is presented in the following paragraphs. This model only requires one universal fitting parameter, which is fitted to 491 single drop measurements.

Drops in extraction columns often have a rigid or partially mobile interface under mass transfer conditions [13]. Moreover, the mobility of the drop interface is partially...
reduced or completely blocked when impurities or surfactants are present. However, the degree of contamination is often unknown [13]. To model the different degrees of mobility the boundary cases of a rigid drop \((v_{t,rigid})\) and an ideally mobile drop interface \((v_{t,mobile})\) are linearly interpolated

\[
v_t = \gamma v_{t,mobile} + (1 - \gamma)v_{t,rigid}
\]

where \(\gamma\) is a linear weighting parameter and \(v_t\) represents the single drop terminal velocity. In case of \(\gamma = 1\) the drop interface is ideally mobile and in case of \(\gamma = 0\) it is rigid.

The boundary case for ideally mobile drop interfaces consists of two models. For small spherical drops with ideally mobile interface the model of Thorsen et al. [15] (Eq. (S1) in the SI) is used \((v_{t,rigid})\). For larger deformed drops, still ideally mobile interface, the model of Hamiec and Johnson [16] (Eqs. (S2) and (S3)) is applied \((v_{t,HJ})\).

The model choice for the case of a mobile drop interface is given in Eq. (2).

\[
v_{t,mobile} = \begin{cases} v_{t,T} & \text{if } v_{t,T} \leq v_{t,HJ} \\ v_{t,HJ} & \text{if } v_{t,HJ} < v_{t,T} \end{cases}
\]

The model of Grace et al. [17] is applied for rigid spherical and deformed drops (Eqs. (S4) and (S5), \(v_{t,rigid}\)).

Fig. 1 depicts the dependency of the drop velocity from the parameter \(\gamma\) with the physical properties of the pure biphasic system \((\text{water(c)/toluene(d)}\) without transfer component) given in Tab. 2.

![Figure 1. Single drop velocity as function of the weighting parameter \(\gamma\).](image)

The absolute drop velocity \(v_d\) is calculated with Eqs. (3) and (4).

\[
v_d = v_t k_x (1 - \varepsilon)^{n-1} - \frac{\dot{V}_c}{A_{\text{red}}}
\]

\[
A_{\text{red}} = A (1 - \varepsilon)
\]

The first term considers the terminal single drop rise velocity \(v_t\). However, the drop movement is slowed down by swarm effects. \(n\) represents the swarm exponent, which is chosen to be 2 according to Kalem et al. [9]. Additionally, drop backmixing due to vortexes induced by internals is considered by the slowing factor \(k_x\) calculated with a model from Garthe [5] (Eq. (S6)). The second term describes the countercurrent velocity of the continuous phase, where \(V_c\) is the continuous phase volume flow and \(A_{\text{red}}\) the free column cross-sectional area. The local accumulation of drops (holdup) leads to less space for the continuous phase to pass the drops. Therefore, the cross-sectional area \(A\) is reduced according to the holdup.

The mass transfer into the drop is calculated with

\[
\Delta m = \Delta t \pi d^2 \frac{1}{\rho_d \beta_d + \rho_c \beta_c} (K x - y)
\]

where \(\Delta t\) is the differential time step size in the simulation, \(\beta_d\) and \(\beta_c\) the mass transfer coefficients of the disperse and continuous phase, \(K\) the distribution coefficient and \(y\) and \(x\) the concentrations in the disperse and continuous phase. The mass transfer resistance in the continuous phase can be neglected when drops have a residence time above 5 s [4], as it is the case in liquid-liquid extraction columns. The mass transfer coefficient for the disperse phase is modeled with the Sherwood correlation given by [10] (Eq. (S7) in SI). Continuous phase backmixing is considered by the model of Breysse and Bühlmann [18].

For the probability of drop breakage, the model of Garthe [5] is applied (Eqs. (S8)–(S10)). The model parameters \((c_1 = 0.0012, c_2 = 0.476, c_3 = 2.58, c_4 = 0.136, c_5 = 0.0285)\) were determined for different chemical systems in the AIF project report “Vom Einzeltropfenexperiment zur Extraktionskolonne” (From single drop experiment to the extraction column) in 2005 [19]. The model of Henschke [6] is used for the coalescence probability (Eqs. (S11) and (S12)).

### 2.4 Simulation Settings and Procedures

In the simulation framework, the differential time-step size is set to 100 ms, as recommended by Kalem et al. [9]. The column is simulated for 15 000 s for each operation point to assure steady-state operation. Furthermore, more than 1000 representative drops per meter extraction column are considered as suggested by Kalem et al. [9]. The drop size distribution at the inlet of the disperse phase is initialized with the measured drop size distribution at the disperser in the DN 80 column as measured by Garthe [5].

In addition, a toolbox is developed, which automatically simulates the operation window, varying volume flow ratios, stirrer speeds and loads of the disperse and continuous phase until in one height element a holdup of > 50 % occurs. This is considered as phase inversion and therefore flooding. The simulation results for each simulated operation point are used to generate a performance map of the simulated extraction column.
3 Results and Discussion

3.1 Single Drop Rise Velocity

The drop rise velocity model was validated with experimental data of single drop rise experiments. The experiments were performed at conditions similar to technical extraction columns, where mass transfer is a necessity and small contaminations can often not be excluded. Therefore, single drop experiments without mass transfer and very pure extraction systems are neglected. As a result, the drop interface in liquid-liquid extraction columns is often only partially mobile or rigid under mass transfer conditions [13].

In total 491 single drop experiments from Haverland [20], Kalem et al. [21], Altunok [22] and Henschke [6] are used to fit the weighting parameter of Eq. (1) to \( g = 0.2 \), which is equivalent to a slightly mobile interface (Fig. 1). The average deviation is 10.4 % and the maximum deviation is below 30 %. The corresponding simulations and experiments are compared in the parity plot in Fig. 2.

![Figure 2. Parity plot comparing the rise velocity model \( (\nu = 0.2) \) with experimental data [6, 20–22].](image)

3.2 Coalescence

Since coalescence is difficult to predict and no universal coalescence model without adjustable parameters is available, it is proposed to use the results of one pilot-plant experiment (per chemical system) to determine the coalescence parameter of the applied model [9, 23, 24]. The derived parameter is then used in the subsequent simulations for the calculation of the whole operation window. Two quantities can be used to specify the coalescence parameter: holdup or drop size. Usually the coalescence parameter is fitted to the drop size distribution of pilot-plant experiments, since the combination of coalescence and breakage probabilities determines the drop size. However, for a fast approximation of the column performance, the simulation is fitted to the holdup. The experimental investigation of the drop size is considerably more difficult than measuring the holdup. For determining the coalescence parameter, the holdup measurement of Garthe [5] in the middle section of the column was chosen at the operation point of 150 rpm and a continuous phase volume flow of \( V_c = 60 \text{ L h}^{-1} \) and disperse flow of \( V_d = 72 \text{ L h}^{-1} \). The coalescence parameter \( C_{\text{coa}} \) of Eq. (S11) is 192. This value is used for the further simulations to generate the performance map.

3.3 Comparison of the Simulation Results with Experiments

The calculated holdup, drop Sauter diameter and concentration profiles are compared with experimental data of Garthe [5]. The simulated drop Sauter diameter has an averaged deviation of 27 % from the experimentally measured diameters. This deviation can be expected since the coalescence parameter was only fitted to the holdup of one column experiment. In Fig. 3 the measured holdups of different operation points (Tab. S1) are compared with simulation results. The graph differentiates between three column sections (lower, middle, upper) where the experimental holdup was measured. The symbol in gray indicates the operating point and section (150 rpm and middle section), which was used to fit the coalescence parameter, and therefore, has a very good agreement. The error bars show the fluctuation in holdup of the two pilot-plant experiments performed by Garthe [5].

![Figure 3. Parity plot of the measured [5] and simulated holdup (fitted operation point: gray symbol).](image)
reported only one drop size distribution at the disperser for all operation points, independent of solvent flow. This can lead to different holdups in the lower section of the column. However, the inlet drop size distribution influences the fluid dynamics only in the inflow region of the disperse phase.

The simulated and experimental concentration profiles are exemplarily shown for the operation point at 200 rpm and \( V_c = 50 \, \text{L/h} \) and \( V_d = 60 \, \text{L/h} \) in Fig. 4, which was not used for the parameter fitting. The calculated concentration profile, therefore, represents an extrapolated operation point. The simulation agrees well with the experimentally measured concentration profiles of the continuous and disperse phase.

3.4 Performance Map

The performance maps for the investigated extraction column at the stirrer speeds of 150 rpm and 200 rpm are presented in Fig. 5 and Fig. 6. The axes represent the specific volume flows of the disperse and continuous phase. The operation domain is spread out between the volume flow ratios 1/8 and 4/1 (o/w) and the flooding boundary (dashed line 100 % flooding). In this domain, the dotted lines represent the holdup, the dashed lines 40 %, 60 %, and 80 % distance from the flooding point (= load/flooding load). The logarithmic grayscale contours show the HETS values. The experimental operation points of Garthe [5] are marked with X. The operation point that was used for the fit of the coalescence parameter is shown in gray.

Garthe reported that the experiments at the highest phase loads are close to the flooding point. The simulated and experimental flooding points agree well at both stirrer speeds (150 rpm and 200 rpm) in the performance maps. The first accessible information in the performance maps is the holdup (dotted lines) and the distance of an arbitrary operation point from the flooding point (dashed lines). For a given column geometry the stable operation window is shown. The holdup depends on three operation parameters, firstly the hydraulic load, secondly the volume flow ratios and thirdly the stirrer speed. The highest holdup is found in the bottom right corner of the operation window. The HETS values vary in the operation window in a broad domain from 0.8 to 0.1. The lowest HETS domain is located in the area of the highest holdup in the bottom right corner of the operation window at high volume flow ratio and at high phase load. Therefore, the preferred operation point is close to the flooding point in both performance maps, where the holdup is high and HETS is low.

The main difference between the performance maps at 150 rpm and 200 rpm is the smaller operation window at 200 rpm. Since the operation window is larger at 150 rpm, the column can be operated at higher specific volume flow rates. As a result, the HETS value close to the flooding point at 150 rpm is smaller (min. \(~0.13 – 0.16\) compared to an operation at 200 rpm (min. \(~0.16 – 0.2\) ). Assuming a column with given height and diameter and required HETS of 0.16 m, the optimal stirrer speed depends on the feed flow rate \( B_c \). At high feed flows (e.g., \( B_c = 15 \, \text{m}^2/\text{m}^3 \cdot \text{h}^{-1} \) ) an opera-
tion at 150 rpm requires more solvent ($B_d = 9 \text{ m}^3\text{m}^{-2}\text{h}^{-1}$) compared to 200 rpm ($B_d = 8 \text{ m}^3\text{m}^{-2}\text{h}^{-1}$). At lower feed flows (e.g., $B_c = 5 \text{ m}^3\text{m}^{-2}\text{h}^{-1}$) it is the opposite.

### 3.5 Performance Map for the Design of Liquid-Liquid Extraction Columns

The proposed application procedure of the performance map for a column design is shown in Fig. 7. For the calculation of flooding points, HETS values and for the generation of a performance map only the physico-chemical properties (density, viscosity, interfacial tension, distribution coefficient, diffusion coefficient) and one pilot-plant for the experimental investigation of the fluid-dynamics are required. The measured holdup is used for the determination of the coalescence parameter. The population balance model can be used to generate and plot the performance map or to directly calculate the column dimensions for a given separation task. For the direct calculation additional information is required: feed composition, volume flow ratio and number of required equilibrium stages. This information can be calculated in a flow sheet model. For given volume flow ratio the phase load is increased until the flooding point is reached. Industrial extraction columns are usually operated at approximately 80% flooding load [1]. For chosen volume flow ratio and operation at 80% flooding load, the specific volume flows (phase loads) and the corresponding HETS values can be calculated with the population balance model. On the basis of HETS and required equilibrium stages the column height is calculated. The column diameter is calculated with the required feed volume flow.

Summing up, the HETS values and flooding points are approximated in the operation window. For a given separation task the approximate column design is easily accessible based on physico-chemical properties and one pilot-plant experiment.

### 4 Summary and Conclusion

A one-dimensional population-balance model for liquid-liquid extraction columns was developed, which requires only physico-chemical properties and one pilot-plant experiment for the approximation of the column performance. For the rise velocity, three models from literature were combined. The resulting model shows good agreement with experimental data with an averaged deviation of 10.4%. Mass transfer is modeled with a universal correlation by Kumar et. al [10].

The population-balance model was successfully applied for a Künni liquid-liquid extraction column, with a diameter of 80 mm and a height of 2.95 m. The coalescence parameter in the derived population-balance model was fitted to one column experiment. A comparison of the simulation results with pilot-plant experiments shows good agreement concerning holdup and mass transfer. The column performance was determined for two different stirrer speeds and the resulting performance maps were presented, showing holdup, HETS values and flooding points as a function of the load. These performance maps can be used to determine column heights and diameters for a given separation task.

The method will be tested for different extraction columns with different internals and chemical systems. Validation of the performance map limits will need further experiments since the available experimental data for Künni columns only cover the domain of a volume flow ratio close to one. A further reduction of experimental effort may be possible by downsizing experiments to mini-plant scale.
Symbols used

- $A$ [m$^2$] column cross section area
- $A_{\text{red}}$ [m$^2$] free column cross section area
- $B$ [m$^3$m$^{-1}$h$^{-1}$] phase load
- $c_{1-5}$ [-] breakage parameter
- $d$ [m] drop diameter
- $k_v$ [-] slowing factor
- $\Delta m$ [kg] transferred mass
- $n$ [-] swarm exponent
- $\Delta t$ [s] time step
- $\bar{v}$ [m s$^{-1}$] velocity
- $\bar{V}$ [L h$^{-1}$] volume flow
- $x$ [-] mass concentration in continuous phase
- $y$ [-] mass concentration in disperse phase

Greek letters

- $\beta$ [m s$^{-1}$] mass transfer coefficient
- $\gamma$ [-] weighting parameter
- $\epsilon$ [-] hold-up
- $\rho$ [kg m$^{-3}$] density

Sub- and superscript

- aqu aqueous
- c continuous phase
- d disperse phase
- HJ Hamilec and Johnson
- org organic
- T Thorsen
- t terminal

References

[1] R. Goedecke, *Fluideverfahrenstechnik*, Wiley-VCH Verlag, Weinheim 2006.
[2] H.-J. Bart, D. Garthe, T. Groemping, A. Pfennig, S. Schmidt, J. Stichlmair, *Chem. Ing. Tech.*, 2006, 78 (5), 543 – 547. DOI: https://doi.org/10.1002/cite.200500146
[3] J. Ayestaran, N. Kopriva, F. Buchbender, M. Kalem, A. Pfennig, *Chem. Eng. Technol.*, 2015, 38 (10), 1894 – 1900. DOI: https://doi.org/10.1002/cate.201500097
[4] M. Henschke, A. Pfennig, *AIChE J.*, 1999, 45 (10), 2079 – 2086.
[5] D. Garhe, *Fluidodynamics and Mass Transfer of Single Particles and Swarms of Particles in Extraction Columns*, Ph.D. Thesis, Technische Universität München 2006.
[6] M. Henschke, *Auslegung pulsierter Siebboden-Extraktionskolonnen*, Berichte aus der Verfahrenstechnik, Shaker, Aachen 2004.
[7] M. Attarakih, M. W. Hlawitschka, M. Abu-Khader, S. Al-Zyod, H.-J. Bart, *Appl. Math. Model.*, 2015, 39 (17), 5105 – 5120. DOI: https://doi.org/10.1016/j.apm.2015.04.006
[8] D. Ramkrishna, *Population balances: Theory and applications to particulate systems in engineering*, Academic Press, San Diego, CA 2000.
[9] M. Kalem, F. Buchbender, A. Pfennig, *Chem. Eng. Res. Des.*, 2011, 89 (1), 1 – 9. DOI: https://doi.org/10.1016/j.cher.2010.05.001
[10] A. Kumar, S. Hartland, *Chem. Eng. Res. Des.*, 1999, 77 (5), 372 – 384. DOI: https://doi.org/10.1016/026387699526359
[11] J. Kamp, M. Kraume, *Chem. Eng. Sci.*, 2016, 156, 162 – 177. DOI: https://doi.org/10.1016/j.ces.2016.08.028
[12] T. Misek, R. Berger, J. Schröter, Standard test systems for liquid extraction, 2nd ed., European Federation of Chemical Engineering by Institution of Chemical Engineers, Rugby 1985.
[13] M. Wegener, N. Paul, M. Kraume, *Int. J. Heat Mass Transfer*, 2014, 71, 475 – 495. DOI: https://doi.org/10.1016/j.ijheatmasstransfer.2013.12.024
[14] J. Villwock, F. Gebauer, J. Kamp, H.-J. Bart, M. Kraume, *Chem. Eng. Technol.*, 2014, 37 (7), 1103 – 1111. DOI: https://doi.org/10.1002/ceat.201400180
[15] G. Thorsen, R. M. Stordalen, S. G. Terjesen, *Chem. Eng. Sci.*, 1968, 23 (5), 413 – 426. DOI: https://doi.org/10.1016/0009-2509(68)87017-4
[16] A. E. Hamielec, A. I. Johnson, *Can. J. Chem. Eng.*, 1962, 40 (2), 41 – 45. DOI: https://doi.org/10.1002/cjce.5450400202
[17] J. Grace, T. Wairegi, T. H. Nguyen, *Trans. Inst. Chem. Eng.*, 1976, 54 (3), 167 – 173.
[18] J. Breyssse, U. Bühlmann, *AIChE Sym. Ser.*, 1983, 80, 94 – 101.
[19] H.-J. Bart, T. C. Grömping, A. Pfennig, M. Schmidt, J. Stichlmair, *Vom Einzeltröpfchenexperiment zur Extraktionskolonne AIF*, Schlussbericht, Aachen, Februar 2005.
[20] H. Haverland, Untersuchungen zur Tropfendispergierung in flüssigkeitspulsierten Siebboden-Extraktionskolonnen, Ph.D. Thesis, Technische Universität Clausthal 1988.
[21] M. Kalem, M. Y. Altunok, A. Pfennig, *AIChE J.*, 2009, 56 (1), 160 – 167. DOI: https://doi.org/10.1002/aic.11976
[22] M. Y. Altunok, Zur Auslegung von Siebbodenextraktionskolonnen für die Reaktivextraktion barierend auf Versuchen im Labormaßstab, Ph.D. Thesis, Rheinisch-Westfälische Technische Hochschule Aachen 2009.
[23] F. Buchbender, F. Onink, W. Meindersma, A. de Haan, A. Pfennig, *Chem. Eng. Sci.*, 2012, 82, 167 – 176. DOI: https://doi.org/10.1016/j.ces.2012.07.035
[24] S. A. Schmidt, M. Simon, M. M. Attarakih, L. Lagar G., H.-J. Bart, *Chem. Eng. Sci.*, 2006, 61 (1), 246 – 256. DOI: https://doi.org/10.1016/j.ces.2005.02.075
Performance Map for the Design of Liquid-Liquid Extraction Columns

B. Weber, C. Meyer, A. Jupke*

Research Article: A population-balance model is applied for the calculation of the operation window of liquid-liquid extraction columns, showing holdup, flooding point and HETS, which can be used for the estimation of the column design. This model only requires the physico-chemical properties and one pilot-plant experiment to determine the coalescence parameter.