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Review

Research on Mass Transfer Columns: passé?

Since university research on mass transfer columns has decreased significantly in recent years, it is discussed whether the current knowledge of mass transfer technology for trays, packing materials, and column internals is sufficient to design energy- and cost-efficient systems. Points of view are described that have not yet been clarified from an industrial position and should play an important role in future research. In addition to the thermodynamics of industrially relevant systems, an improved description of the fluid dynamics and mass transfer efficiency in trayed and packed columns is included. Research should focus beyond ideal test mixtures which employ standardized test equipments and test systems. Industrial parameters such as foams and solids as well as chemical and catalytic mass transfer are also of great interest.

Keywords: Mass transfer column, Packings, Standard testing procedure, Trays

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1 Introduction to Mass Transfer Column Design

Process simulators, such as Aspen, ChemCad, Hysys, and Pro/II, have become standard design tools to simulate complete processes from their origins in feedstock through the final products. Various unit operations can be selected to create a complete flow diagram with straightforward and/or recycle flow streams. In such a flow diagram, the composition of streams can change significantly during the various unit operations. For example, different solvents can be used to treat gas flow streams. Several thermodynamic models can be selected to simulate the physical and thermodynamic properties of the process, such as Peng Robinson (PR), NRTL, Wilson, Van Laar, Soave-Redlich-Kwong (SRK), Chao-Seader, Hayden O'Connell, Flory-Huggins, Grayson Streed, Margules, Uniquac, and Raoult's Law, to name but a few. Selecting the correct thermodynamic model with appropriate interaction parameters at the individual process steps is essential for suitably simulating a process. The simulation results are generally reliable when the correct thermodynamic model has been used for the individual process steps.

Following this design step, the individual unit operations have to be dimensioned in detail. The following sections focus on mass transfer columns as an individual unit operation. The design process for a mass transfer column starts with a technical specification that is elaborated by the process simulation.

This specification includes a list of feed components and flow rates as well as target specifications for the column overhead and bottom products. Pressures and temperatures of the various feed and product flows, column internal flow rates, and physical and kinetic properties of the phases can be derived from the process simulation. Furthermore, information about the number of theoretical stages or mass transfer segments is also available from the process simulation.

The next design step is the proper selection of *efficient* mass transfer column internals. Efficient mass transfer column internals are items supporting the separation process such as mass transfer trays and structured or random packing materials. Column diameter, pressure drop, liquid holdup, and tray efficiency have to be estimated to start sizing the individual columns. The height equivalent to a theoretical plate (HETP) or the height of a transfer unit (HTU) can also be used as an alternative to the tray efficiency. The following is a list of options to determine this information.

- Experimental data from universities and test institutes can provide information about capacities, pressure drops, and liquid holdup. However, this approach primarily yields information on standard test mixtures. This limitation persists for information regarding the separation efficiency. Thus, the experimental data from standard test mixtures need to be transferred to industrial fluids and conditions, using either experience or rule-of-thumb approaches. This approach is valid if the physical properties for the industrial applications are close to the test conditions. In industrial applications, the component mixtures and phase flow ratios often differ significantly from these standard conditions. That is why it can be difficult to evaluate the fluid dynamics and separation efficiency for industrial columns. If the industrial conditions differ significantly from the ideal test

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systems, lab or mini-scale experiments that use industrial fluids and, if possible, industrial flow conditions can be conducted as an alternative. Nevertheless, this approach may not answer all questions regarding industrial scale-up.

- State-of-the-art calculations can also be employed to predict flooding capacity, pressure drop, liquid holdup, and separation efficiencies. Various published correlations are available for this purpose. However, these correlations only offer reliable results for the test mixtures and efficient column internals from which they were derived. Consequently, this approach may not be reliable for extrapolations.
- To avoid the above-mentioned uncertainties, the operating experience of existing (older) plants is often considered during industrial practice. This approach often avoids modern efficient column internals due to a lack of experience and does not consider any resultant potential benefits.

Additional hardware items beyond the use of efficient column internals are required to design a mass transfer tower. These items consist of liquid and gas distributors, droplet separators, liquid collectors, support and hold-down devices, two-phase flashing feed devices, internal piping, and downcomers, to name but a few. These hardware units are not supporting the mass transfer efficiency but are essential for proper column operation and are defined as *non-separating* column internals.

Fig. 1 presents a typical mass transfer column that consists of several sections. The packed bed accounts for only a fraction of the column volume. Column internals, such as droplet separators, liquid distributors, support and hold-down constructions, annular channels, and gas distributors, comprise a significant additional share of the column volume. To minimize

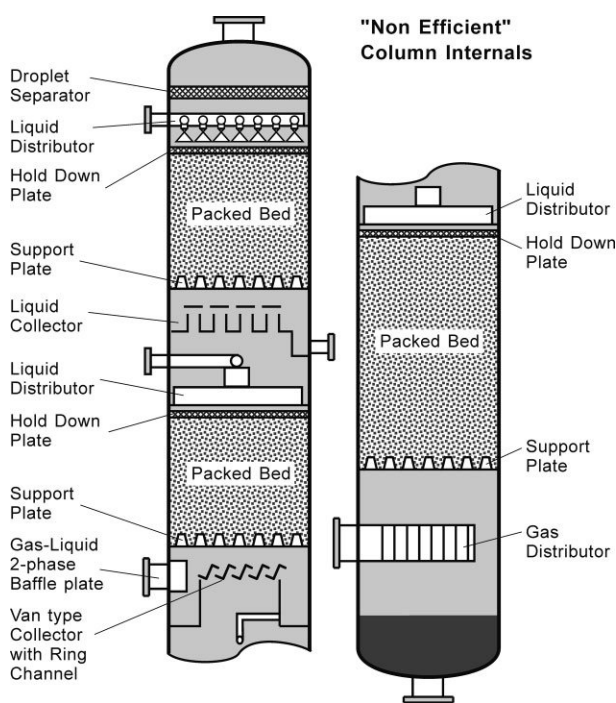


Figure 1. Layout of a mass transfer column with several feeds and bed sections, as well as various non-separating tower internals.

the column volume, the non-separating column internals have to be sized properly not only with respect to mechanical aspects but also with respect to fluid-dynamic principles. For example, liquid distributors should not overflow at the upper liquid loading conditions. Conversely, they should keep a minimum liquid level at low loading conditions. Liquid feed pipes should direct the liquid to the distributor without a strong impulse. A gas distributor should direct the gas with a low pressure drop into the column by ensuring a homogeneous gas distribution. The same is given for gas-liquid two-phase feed entries (flashing feeds). These are only a few parameters that have to be taken into account for a proper column design. The influence of non-separating column internals on the function of a mass transfer column was frequently underestimated in the past, as reported by Kister [1].

The next section discusses the use of experimental data versus the application of state-of-the-art equations to obtain reliable fluid dynamic and mass transfer information for industrial column design. The following section will then focus on non-separating column internals.

2 Experiments and Calculation Modes for Fluid Dynamics and Separation Efficiency of Efficient Column Internals

2.1 Reliability of Experimental Data

Experimental data are typically obtained from lab-scale columns with diameters of maximum 600 mm. Few industrially scaled test facilities with column diameters >800 mm exist, such as the example provided by Fractionation Research Inc. (FRI) in Stillwater, OK, USA. The following must be first considered in this approach: generating reliable and valuable test data and transferring small-scale test results to industrial column designs. For example, the performance data for mass transfer trays are noticeably influenced by the tray deck and downcomer layout, e.g., active area, flow path length, open deck area, inlet and outlet weir design, valve geometry, tray spacing, etc. The test system and system pressure also affect the test results [2]. Fig. 2 illustrates the various performance efficiencies of sieve trays with different hole diameters tested at FRI [3]. This example demonstrates that the tray performance data vary noticeably with geometric design parameters. Therefore, using tray test data for industrial design requires detailed information about the test setup.

One would expect that test data for structured or random packing materials are easier to use for industrial column design because the geometric data of structured or random packing materials are almost independent of the column diameter (neglecting wall effects). Fig. 3 displays the pressure drop that occurs when the metal Raschig Super-Pak 250 material is tested in a standard air/water system under ambient conditions. The pressure drop was measured for a constant liquid rate of $10 \text{ m}^3 \text{ m}^{-2} \text{ h}^{-1}$ and various F -factors ($F_V = u_V \sqrt{\rho_V}$). The tests were performed at the Ruhr University of Bochum, the Separation Research Program (SRP) in Austin (TX, USA), BASF SE in Ludwigshafen, and the Bulgarian Academy of

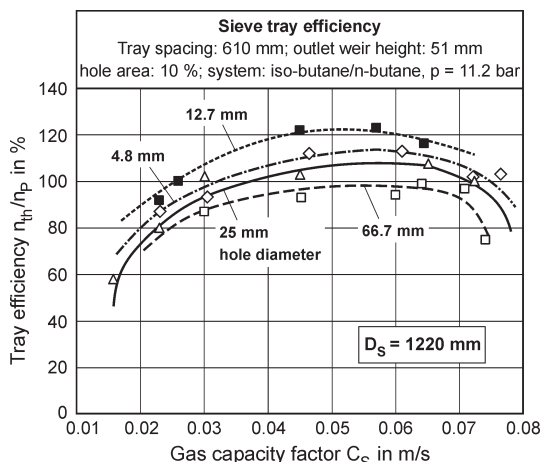


Figure 2. Hole diameter effect on tray efficiency for sieve trays [3].

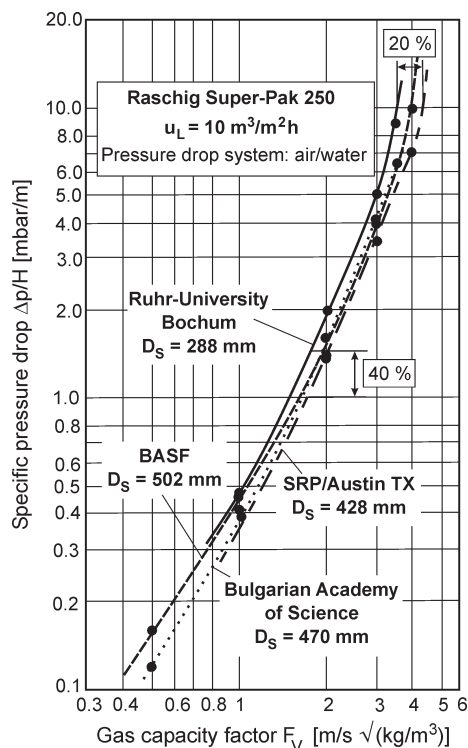


Figure 3. Specific pressure drop versus F -factor at a constant liquid load for the Raschig Super-Pak 250 measured at different test institutes.

Science in Sofia. The results indicate a 40% and 20% difference in the pressure drop and capacity limit, respectively. It is interesting to note that even with this simple air/water test system, the fluid-dynamic test results vary widely among institutes for the same product.

A report by Olujic [4] argues that the pressure drop of a structured packing material depends on the column diameter for small column diameters of up to 1 m (wall wiper effect). Olujic concluded that the pressure drop decreases and that the

capacity becomes higher with increasing column diameter. A similar trend can be observed in Fig. 3 in the mid and upper gas capacity range but not at low gas flow rates.

Conversely, Fig. 4 and Fig. 5 demonstrate pressure drop and mass transfer efficiency measurements for distillation tests from the following three test institutes: Fractionation Research Inc. (FRI, column diameter 1.22 m), Separation Research Program (SRP, column diameter 0.43 m), and Prague Institute of Chemical Technology (ICT Prague, column diameter 0.15 m). The test system, system pressure, and mid-bed compositions were very similar. Comparing the pressure drop results in Fig. 4 does not show a clear effect of column diameter (wall wiper effect) on packing performance.

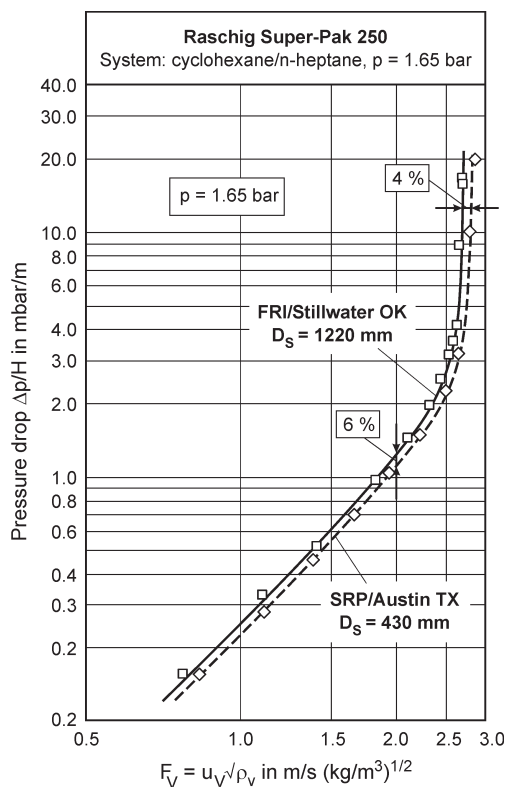


Figure 4. Specific pressure drop of the Raschig Super-Pak 250 tested at various test institutes with the same test system under rectification conditions but with different column diameters.

Fig. 5 provides information on the efficiency of the Raschig Super-Pak 250 material in terms of an HETP value for various F_V -factors under total reflux distillation conditions. The test system was cyclohexane/*n*-heptane at 1.00–1.65 bar column top pressure. The HETP value differs by 20% depending on the test institute selected.

Fig. 6 presents measurements for mass transfer efficiency in terms of the specific liquid side mass transfer coefficient, $\beta_{L, a_{ph}}$. The tests were performed using the standard test system that strips CO_2 from water into air under ambient conditions at a constant gas rate but varying liquid loads. The 25- and 50-mm Pall-Ring products were tested. Billet and co-workers published data for 50-mm metal Pall-Rings in 1985 and

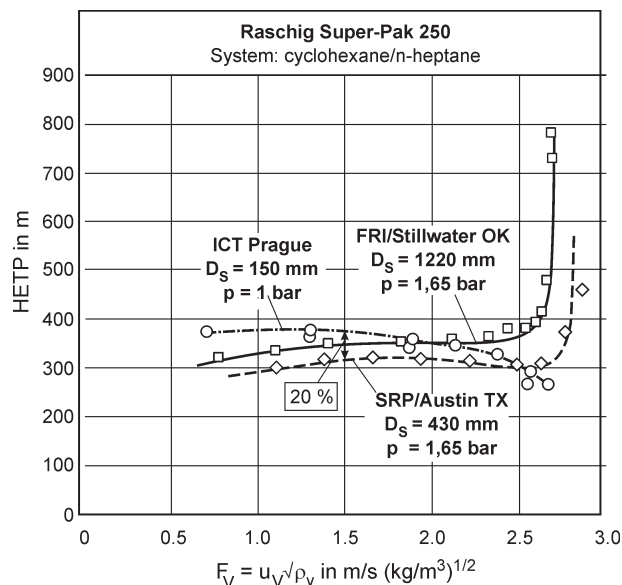


Figure 5. Separation efficiency of Raschig Super-Pak 250 tested at various test institutes with the same test system but different column diameters.

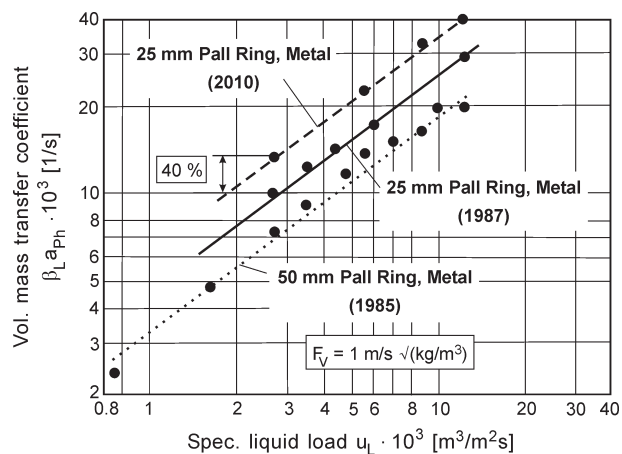


Figure 6. Volumetric liquid side mass transfer coefficient versus liquid load for a constant F -factor. Data for 25- and 50-mm metal Pall-Rings.

25-mm metal Pall-Rings in 1987 [5,6]. These measurements show the typical improvement in efficiency one can expect when switching from 50-mm to 25-mm Pall-Rings. In 2010, Grünewald [7] retested the 25-mm Pall-Ring in the same test facility but with different results. They eliminated the wall effect in their experiments, and the efficiency difference between both measurements was 40%.

Fig. 7 provides information to determine the overall height of a mass transfer unit, HTU_{OV} , for another typical test system which facilitates the absorption of ammonia from air into water under ambient conditions. For these tests, the liquid load was held constant and the F -factor varied from low values up to the flood point of the packing. The results for the 25- and 50-mm metal Pall-Ring published by Billet and co-

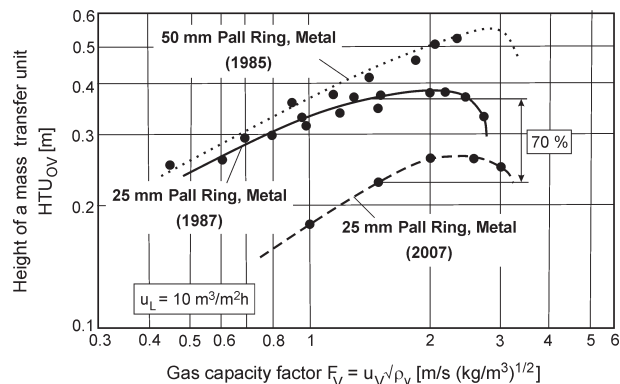


Figure 7. Height of an overall mass transfer unit versus F -factor for a constant liquid load. Data for 25- and 50-mm metal Pall-Rings.

workers [5, 6] are presented in Fig. 7. In 2007, Hoffmann et al. retested the 25-mm Pall-Rings in another test facility but under equivalent test conditions [8]. The HTU_{OV} values obtained by Hoffmann et al. were significantly lower than those obtained by Billet et al., with the data sets differing by approximately 70%.

First conclusion

Currently, the performance of an efficient column internal device depends significantly on the test setup and the institute that has studied the hardware device. Thus, it is necessary to standardize the experimental layout of test facilities to obtain equal results from different test facilities. Simple standard test systems to establish the fluid dynamics and separation efficiency of trays and random or structured packing materials should be defined. To prove a suitable facility setup, universities, institutes, and industries must agree on the same pressure drop, capacity, liquid holdup, and separation efficiency for a standard packing or mass transfer tray. Furthermore, test systems with significantly different physical properties should be included to cover industrially relevant conditions. In all cases, the recommendations should be associated with a suitable thermodynamic model to describe the phase equilibrium and physical properties. A list of geometric parameters should be elaborated to report trays and packing materials in the literature. A report by Olujić represents an initial example of such communication that provides information for standard structured packing tests under distillation conditions [4].

2.2 Reliability of Calculation Models

As an alternative, current industrial column design can rely on calculation models. For example, various equations to estimate the fluid-dynamic performances or mass transfer efficiencies of trays are available from Chan-Fair, Zuiderweg, Hughmark, Harris, AIChE, Locket, Glitsch, and Kister-Haas. Similar equations for packing materials are offered from Billet-Schultes, Stichmair, Onda, Bravo-Fair, Bravo-Rocha-Fair, and Mackowiak, among others. The criteria for selecting a proper industrial column design rely on extensive experience because the

equations can only be applied when actual conditions do not differ significantly from the test systems and packing materials they correlate with. These equations may result in major errors when used for extrapolations. The following examples demonstrate how diverse (extrapolated) calculated results can be.

To predict the sieve tray mass transfer efficiency and pressure drop, the clear liquid height on a tray deck has to be determined. Rahimi et al. [9] presented a study on various sieve tray hydraulic parameters with CFD modeling. Parameters were predicted with state-of-the-art published equations and compared the results with experimental data. For example, Fig. 8 displays the clear liquid height on a sieve tray deck that was experimentally studied by Solari and Bell. Calculated results that apply equations from Colwell et al. and Bennet et al. are also depicted in Fig. 8. In addition, two CFD studies, i.e., Gesit et al. with outlet downcomer and Rahimi et al. with inlet and outlet downcomer, are incorporated in the comparison which shows a deviation of up to 42 % between calculated and experimental data.

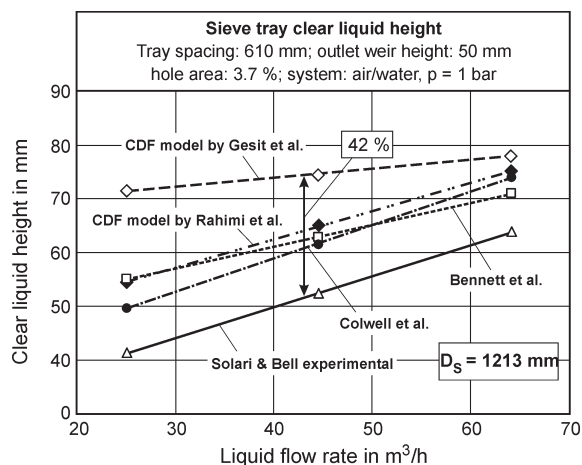


Figure 8. Clear liquid height on a sieve tray as a function of liquid flow rate. Experimental values by Solari & Bell, CDF modeling by Gerit et al. and Rahimi et al., modeling by Bennet et al. and Colwell et al.; hole diameter: 12.7 mm; downcomer clearance: 38 mm; downcomer area: 13 % (total area); hole area: 3.7 % (total area) or 5 % (bubbling area); pitch: 50 mm triangular; weir length: 925 mm [9].

Kister [10] also provided three different procedures for designing a sieve tray column to remove propane. He concluded with the following statement: “The example reflects the prime difficulty often encountered by tray designers; inconsistent predictions from different correlations. The three entrainment flood correlations used gave predictions that widely differed; the differences were up to 50 to 60 percent.” Klemola and Ilme [11] came to the same conclusions by comparing two-pass valve tray efficiency data from an industrial *i*-butane/*n*-butane fractionator with experimental data from FRI and 18 different prediction methods. The differences between predictions and measured tray efficiencies approached 50 %. The experimental efficiency data measured by FRI closely approximated the industrial data.

Currently, the same conclusion is still given for packing performance data. The black solid circles in Fig. 9 represent experimentally determined HETP values to rectify the ethanol/water system published by Linek [12]. Concentration profiles were measured along the test column to obtain the HETP values along the bed. In 2005, Linek applied an equation to predict the gas and liquid side mass transfer coefficients, $\beta_{Va_{ph}}$ and $\beta_{La_{ph}}$, using the liquid load, u_L , and the gas velocity, u_v , as process variables. Constants and exponents served as fitting parameters in the equations ($\beta_{Va_{ph}} = C_1 u_L^{a_1} u_v^{b_1}$; $\beta_{La_{ph}} = C_2 u_L^{a_2}$) to predict HETP values. The curve crossing the experimental points in Fig. 9 is based on the best fit of constants and exponents in these correlations. Linek and his colleagues also applied the original Billet-Schultes model published in 1993 [13] and adapted the liquid- and gas load-related exponents such that the deviation from their experimental results was minimized (black solid curve). In addition to this effort, they calculated the HETP values by means of the original Onda model from 1968 [14] and the original Billet-Schultes model from 1993 [13]. Finally, Fig. 9 shows the HETP values calculated from an equation published by Linek and colleagues in 1995 [15]. This equation was correlated using different test systems. These results demonstrate that the calculated values depend strongly on the calculation method used and can vary significantly from experimentally determined values.

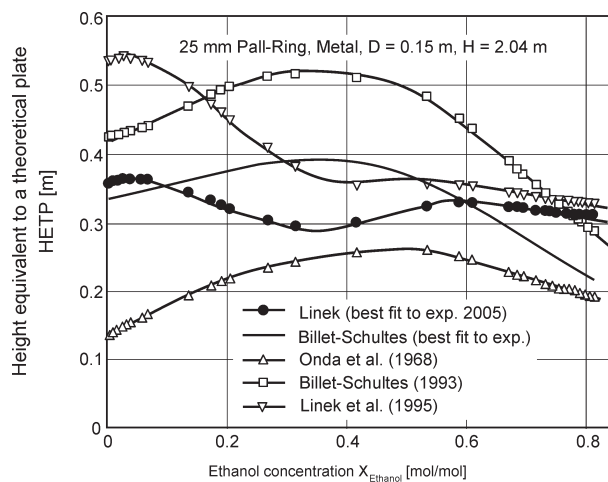


Figure 9. Measured HETP values for the rectification of the ethanol/water system and calculated values according to different models [12–15].

Recently, another attempt was made to pre-emptively calculate the effective specific mass transfer area, a_{ph}/a , using correlations known to industry. These calculations resulted in comparisons between the measurement results and the calculations presented in Fig. 10 for the ideal CO_2 -air/sodium hydroxide test system. These comparisons were published by Lewis et al. [16]. Fig. 10 also incorporates the new Tsai model that was published in 2011 [17]. The Tsai equation was originally developed for structured packing materials but could be applied to random packing materials if the flow parameter defined by Tsai, Q_L/L_p ($m^3 s^{-1} m^{-1}$), is substituted by the specific liquid load,

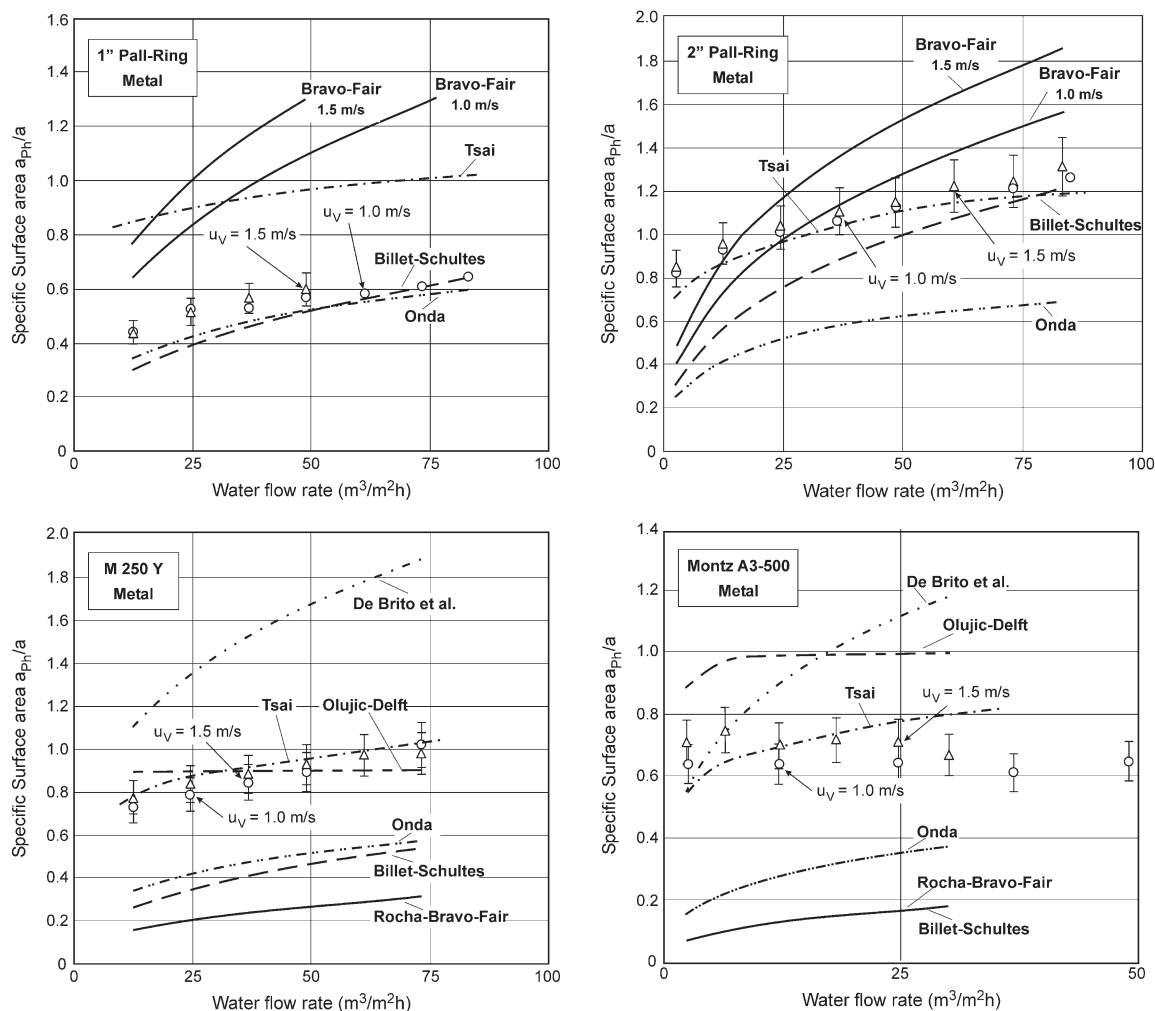


Figure 10. Experimentally determined specific mass transfer area and calculated values from different correlations [17].

u_L ($\text{m}^3 \text{m}^{-2} \text{s}^{-1}$) and hydraulic packing diameter, $d_h = 4\epsilon/a$ (m). This substitution can effectively be summed up by the following expression: $Q_L/L_p = u_L d_h/4$.

The Billet-Schultes and Onda models can predict the effective specific surface area for 25-mm metal Pall-Rings. The new Tsai model overestimates the effective surface area. For the 50-mm Pall-Rings, the Bravo-Fair model can be applied at low liquid rates, while the Billet-Schultes model predicts values close to those observed during experiments at high liquid rates. The Tsai model closely approximates all of the experimental data. For the Mellapak 250 Y packing, the Olujic and Tsai models can predict the effective surface area. The model proposed by De Brito and co-workers generates values for the Montz A3-500 structured packing that correlate with experimental data at very low liquid rates. Only the Tsai model correlates well with experimental data at low and mid-range liquid rates. Any deviations can be explained by the fact that the equations reported were only in good agreement with experimental data if the packing was incorporated in the model validation. Therefore, significant deviations can result if models are applied to packing materials that were not incorporated in model validation.

Second conclusion

As observed in Figs. 8–10, different correlations provide different answers to questions regarding fluid dynamics, separation efficiencies, and specific mass transfer area. Additionally, it certainly bears repeating that different test institutes may also obtain different performance figures.

Based on the author's experience, the deviation between calculated fluid-dynamic data and experimental values is approximately $\pm 5\%$ if one test system and one test facility is considered for the studies. If the experiments are conducted in different test facilities but with the same test system and device, differences of $\pm 15\%$ can occur between the predictions and experiments. If different ideal test systems are used beyond this scope and the results are correlated, 90% of all test data can be predicted with an accuracy of $\pm 20\%$ using state-of-the-art equations. When different model equations are applied to industrial systems, significantly higher differences can occur if predictions extrapolate the experimental data from which these models were derived.

Based on the author's experience, the experimental mass transfer efficiencies of trays and random or structured packing

materials are even more inhomogeneous which explains the larger deviations from precalculations. If a model is correlated to mass transfer efficiency measurements in one column for one device, the calculated and measured data deviate from one another by approximately $\pm 10\%$. If multiple mass transfer data sets, i.e., absorption, desorption, and rectification, are used for a single type of tray or packing device, the state-of-the-art correlations will predict the experimental performance with a deviation of $\pm 20\text{--}30\%$.

If the physical properties or flow rates in an industrial column differ from the secured test systems, any correlations necessarily extrapolate from experimental data which significantly increases uncertainty levels. State-of-the-art equations will accurately predict 80–90% of industrial applications by a $\pm 20\text{--}30\%$ error margin, but in 10–20% of industrial cases, the calculated separation efficiency can deviate by up to $\pm 80\%$ depending on the equation employed. Similarly, the calculated effective mass transfer area can deviate by more than $\pm 100\%$, as indicated in Tab. 1. The predictive uncertainties increase further if foaming or fouling systems are to be evaluated. The same applies to extreme physical properties, such as highly viscous (ionic) media.

3 Design of Non-Separating Column Internals

In addition to trays or random and structured packing materials that serve as mass transfer efficient elements, each mass transfer column also has non-separating tower internals such as liquid distributors, hold-down grids, support grids, gas distributors, and liquid collectors, to name a few (Fig. 1). These column internals must function properly to comply with the performance figures of a mass transfer column. To date, limited fluid-dynamic guidelines have been published for dimensioning these internals.

For example, feeding a trayed or packed column with a two-phase gas/liquid mixture (flashing feed) is a common industrial practice. The phases often have to be separated with special flashing feed devices inside the column, such as circular flashing galleries. The gas and liquid flow are directed around the inner column shell in a gallery-type construction as illustrated in Fig. 11. The phases are separated by the centrifugal forces of the spinning gas and liquid. The design criteria for flashing feed



Figure 11. Flashing gallery for a packed mass transfer tower.

devices have not yet been thoroughly investigated, resulting in numerous empirical design concepts. For example, the kinetic energy of a flashing feed entering the column is often used for column nozzle sizing. As a design criterion, the kinetic energy of the phase mixture is set by a limited maximum allowable value which varies noticeably from company to company. The piping isometrics also influence the flash gallery design but are seldom taken into consideration. Based on the author's experience, this is an important design parameter because it noticeably impacts the flow regime of flashing feeds.

Flow maps according to Baker, e.g., are applied to characterize the flashing feed flow regime [18]. Other flow maps are also in use, but they only apply to fully developed flow regimes. As stated above, the piping configurations outside industrial columns often preclude the full development of a two-phase flow regime before entering the column. The droplet size initiated by the flow regime of flashing feeds is another important design aspect that has rarely been studied to date. This initiation can occur either in the feed line or when leaving the gallery with the gas phase to the column section above. However, this type of information is essential for designing droplet separators (vane or wire mesh type). Entrainment to the bed above can also impact column performance. The distances between flashing galleries and droplet separators, packed beds or overhead gas nozzles are also very empirical. These distances often require significant column heights and, consequently, investment cost.

In contrast to flashing feed designs, the study of single gas feed configurations has been reported in various papers. For example, gas feed nozzles, baffle plates, vane type distributors, and horn devices have been investigated [19–23]. The conclusions given by the authors are not always the same which further underlines interest in this research area. These differing

Table 1. Mean deviations between the experimentally determined performance data and calculated values with state-of-the-art models.

Fluid-dynamic performance numbers		Number of known test systems	Mean deviation for interpolation	Mean deviation for extrapolation	
Capacity/flooding limit	$F_{V,FI}, C_{S,FI}$	13	$\pm 20\%$	$\pm 30\%$	
Specific pressure drop	$\Delta p/H$	25	$\pm 20\%$	$\pm 30\%$	
Liquid holdup	h_L	20	$\pm 10\%$	$\pm 30\%$	
Mass transfer	Absorption and desorption	$\beta_V a_{Ph}, \beta_L a_{Ph}$	31	$\pm 20\%$	$\pm 50\%$
	Rectification	$HTU_{OV}, HETP$	14	$\pm 30\%$	$\pm 80\%$
	Chemisorption	a_{Ph}/a	4	$\pm 80\%$	$> \pm 100\%$

interpretations can also be attributed to somewhat arbitrary gas feed nozzle sizing. The design was partly based on velocity or kinetic energy criteria. Some authors considered the pressure drop of trays or packing materials, others the ratio of column diameter to gas feed nozzle.

The application of computational fluid dynamics (CFD) modeling is a modern approach to these uncertainties. CFD modeling provides flow profiles and can be applied to complex geometries (Fig. 12) [24]. However, CFD studies often cannot be validated experimentally, especially if large-scale and complex configurations are studied. Ali et al. [25] demonstrate how CFD can help to understand the gas distribution at liquid redistributors and liquid collectors if the results are experimentally validated. However, the liquid phase is usually neglected because it increases the complexity of mathematical modeling. Nevertheless, CFD often provides qualitative results as long as the liquid phase is not considered.

The distances between tower internals are another important design consideration in industry. Currently, these distances are often empirical, and different design criteria are applied in industry. For example, the distance between a gas feed nozzle and the tray or packed section above or the liquid level below is often arbitrary. The distance to the trays or packing materials above can be related to various parameters: the column diameter, the pressure drop of trays or packing materials or the gas inlet velocity (or kinetic energy) in conjunction with the type of gas feed device. Others simply recommend a constant distance for all column diameters.

The design of a liquid distribution system is another well-understood concept. Distributors in packed mass transfer towers are one of the most important hardware items. If the liquid distribution is not homogeneous over the column cross-sectional area, the separation efficiency is noticeably reduced, as indicated by Moore and Rukovena [26]. In Kister's failure list for column troubleshooting cases, distributor design failures are ranked at number six [27].

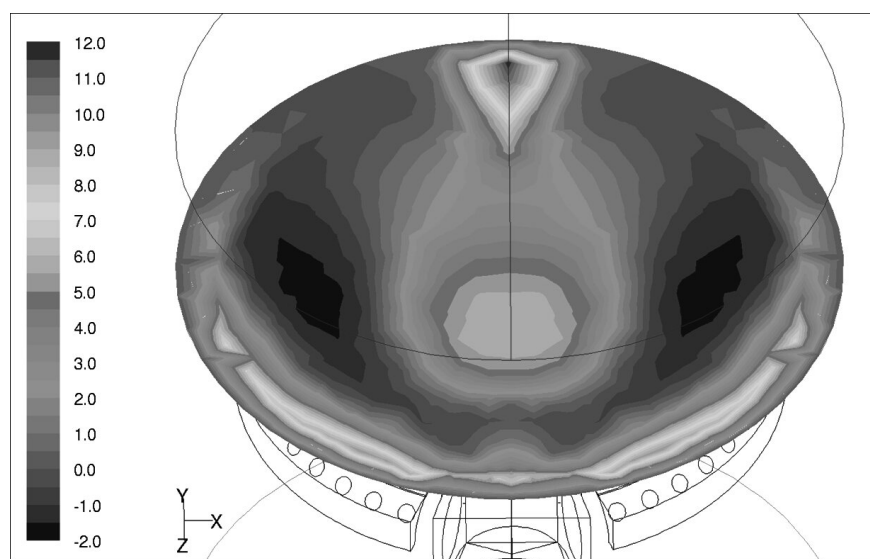


Figure 12. CFD simulation of a circular gas inlet pipe [24].

Based on the author's experience, a proper liquid distribution system design starts with the piping isometric outside the column and the suitable sizing of the feed nozzle diameter. The design process subsequently includes the feed pipe arrangement inside the column to direct the liquid towards the trough- or deck-type distributor. The design principles of a liquid distributor seem to be simple, but the discharge of liquid through holes becomes difficult to predict if industrial conditions are considered. For example, the industry lacks consensus about the effect of various parameters on the flow through distribution orifices. These parameters include the ratio of the hole diameter to deck thickness, hole shape (e.g., laser-cut, punched or drilled), hole orientation (e.g., horizontal, perpendicular, flat or round in pipe), physical properties (e.g., liquid density and viscosity), and fluid-dynamic properties (e.g., overflow velocity). Furthermore, the impact of foam and solids on the liquid discharge performance is also under discussion in industry. Further parameters to be considered are reported by Schultes et al. [28].

Third conclusion

Kister [1] describes multiple instances of design failures for *non-separating* column internals. Industrial vendors of trays and packing materials have confidential design rules to predict the fluid dynamics of liquid distributors, gas distributors, liquid collectors, two-phase flashing feeds, wire mesh droplet separators, and internal piping, to name but a few. These criteria are also of strong interest to those operating industrial processes. More university research should help to provide independent design rules for non-separating tower internals to the open domain.

4 Industrial Column Design

Because of the aforementioned uncertainties, the appropriate design of an industrial column still requires significant experience. It is recommended to consider the following steps when sizing new industrial columns:

- Search for industrial performance data in applications similar to the new design process condition. Apply the same design to the new column. Alternatively, study the performance of industrial trays or packing materials using empirical fluid-dynamic and efficiency correlations. Use possible model parameters to fit the model to the existing industrial performance data. Apply this model to the prediction of fluid dynamics and mass transfer of the new column. Unfortunately, detailed industrial performance information is rarely available.
- Search for published test systems in a similar pressure range and with comparable physical properties to the new design conditions. Transfer these test results to the new process conditions by applying an empirical

correlation. Use possible model parameters to fit the model to the published test data. Use the fitted correlation for the new column design. Experience is required to approve published test data for a new column design as described in the previous sections.

- Perform new laboratory tests if uncertainties are too significant, e.g., very high viscosity, the reaction kinetics are unknown or the equilibrium prediction is uncertain. The thermodynamic properties and system pressure of the test system should be as close as possible to the new process conditions. Apply or transfer the test results with empirical correlations to the new design conditions.
- Make use of experience of similar devices in industrial columns to select and design non-separating column internals. For liquid distributors, demand a liquid flow test at the manufacturer's workshop to prove the offered design and eliminate fabrication errors. Use CFD to visualize the fluid dynamics of tower internals. Use this information to select the most appropriate device. Prove a sub-supplier's detailed design with personal experience or design criteria.
- Make use of safety margins where information and experience is incomplete or missing.

5 Future Research Tasks

As explained in the preceding sections, mass transfer column research certainly cannot be deemed complete (Tab.2). The field of thermodynamics has thus far shown good progress on the advanced calculation of physical properties or phase equilibrium. Further research activities will still be necessary to assure the precise prediction of the thermodynamic properties of new fluids, such as ionic solvents. New model approaches may also be needed. In addition, thermodynamics should also continue to study reactive and catalytically active fluids because industrial processes will continue to be energetically optimized and promoters/additives will have to be added to the fluids to achieve targeted effects. The development of amine solvents for CO₂ capture technology in power plants is a typical example of such fluids. Various additives, such as piperazine and antioxidants, are under discussion to improve the reactivity and hinder degradation of the amine by oxygen.

Based on various research activities in the field of thermodynamics over the past decades, a large number of thermodynamic models are available to predict gas/liquid or gas/liquid/liquid equilibrium data as well as physical, thermal, and kinetic parameters. A universal decision tree would be very helpful to properly select thermodynamic models for industrial applica-

Table 2. Main preferable research areas in the field of mass transfer columns.

Area of research	Focus of activities
Thermodynamics	<ul style="list-style-type: none"> – Description of new systems (e.g., ionic solvents) – New models for more accurate descriptions of physical and equilibrium properties – Development of a universal decision tree for recommending a thermodynamic model for mass transfer systems
Fluid dynamics	<ul style="list-style-type: none"> – Description of standards for test equipment and experimental testing procedure – Influence of foaming and fouling systems – Influence of industry-relevant extreme physical properties and column loads – Further development of the calculation model using modern analysis procedures/measurement techniques – Predictive description of modern trays and packings
Separation efficiency	<ul style="list-style-type: none"> – Description of standards for test equipment and experimental testing procedure – Further development of the calculation models – Influence of foaming and fouling systems – Influence of industry-relevant extreme physical properties and column loads – Separation efficiency at very high product purity levels – Development of generalized performance figures for systems with chemical reactions (HETP_R or HTU_R). – Predictive description of modern trays and packings
Non-separating column internals	<ul style="list-style-type: none"> – Fluid dynamics and hardware design criteria for distribution systems for – liquids, gases, gas/liquid mixtures, liquid-liquid systems, solid-liquid systems, collectors and redistributors – Influence of foaming and fouling systems – CFD modeling (with experimental verification)

tions. Research institutes can play an important role in this approach.

A wide range of unanswered questions remains in the field of fluid-dynamic and mass transfer efficiency of trayed or packed towers. Research should also be oriented towards practical problems, such as the influence of impurities on fluid dynamics or the separation efficiency of trays and packing materials. Grosserichter and Stichlmair as well as Zhou and Zhang published well-recognized papers about fouling principles in trayed and packed columns [29–31]. The same interest applies to the description of foam development and foam impact on mass transfer efficiency. Thiele et al., Ratman et al., and Senger and Wozny [32–34] presented published work in progress to significantly improve understanding of the development and behavior of foam in packed columns.

Modern fluids, such as ionic solvents, have high viscosities and molecular weights, which raises questions about fluid dynamics and mass transfer. It is not certain if the former regularity for the prediction of diffusion coefficients and mass transfer coefficients can still be applied to highly viscous media.

Due to considerable industrial demand for investment- and energy-saving processes, the flow conditions applied in mass transfer columns are sometimes becoming increasingly extreme. As a result, very high liquid loads are partly coupled with extremely low gas rates. Ionization of drinking water and CO₂ absorption from natural gas into physical solvents are typical examples of processes with low gas velocities and high liquid rates. Further experiments that study very low gas rates and very high liquid rates should help to design such towers in industry. Requirements for product purity levels also continue to increase, so that an increasing number of industrial columns operate at very high purities (ppm and ppb concentrations). Thus, the mass transfer efficiency measured in the medium concentration range may not be applicable.

The definition of test facility setup standards is also relevant to experimentally determine and prove fluid dynamics and separation efficiencies. Olujić [35] attempted to measure the standard mass transfer with structured packing materials in distillation test equipment. Such approaches are also needed for absorption and desorption processes as well as random packing materials and trays. Hoffman et al. [8] and most recently Kunze et al. [36] reported research activities on the standardization of mass transfer measurements for absorption and desorption processes. Lastly, tray and packing vendors as well as the applying industry have to agree on these standards. It should be also developed a performance figure (pressure drop, capacity, liquid holdup, and efficiencies) of standard trays and packing materials that are ultimately agreed upon. Such agreement is important to validate any test facility throughout the world. The work in progress by Kunze et al. [36] is a promising approach as the industry partners are involved.

As described in the previous sections, the calculation methods to determine pressure drop, liquid holdup, and capacity limits, mass transfer coefficients and effective mass transfer areas must be developed further. High safety margins are incongruent with the energy-saving concepts. Industrial engineers not only require calculative results but also the opportunity to analyze the simulation results. For example, figures that apply to chemical reactions in mass transfer columns and can also be used for plausibility testing should be developed. For instance, a system-related HETP_R or HTU_R value can be evaluated from experimental data as a function of the chemical reaction speed and component diffusivity (Hatta number) in reversible chemical reactions. For irreversible chemical reactions, the HETP_R value becomes meaningless and the HTU_R value can help to characterize the reactive system.

To date, the pre-emptive calculation of the fluid dynamics or separation efficiency has not been possible in the development of new mass transfer trays of packing materials. A wide range of complex tests must be performed before a new product can be placed on the market. This testing leads to considerable disadvantages in industry because potential improvements in mass transfer are identified and implemented late. The development of criteria that drastically reduce the product development and market introduction timeline would be preferable to current testing. CFD and computational mass transfer (CMT) can be important tools for this type of research; Liu and Yin have published initial interesting approaches to this effect [37, 38].

In general, CFD should play an important tool to visualize the dynamics of fluid flow around efficient and non-separating tower internals. However, experimental validation data should correlate with this type of modeling as much as possible.

Non-separating column internals should also be a growing area for research. The fluid-dynamic performance of non-separating tower internals at very low surface tensions is of major interest in industry. These non-separating tower internals include distributors, droplet separators, and/or flashing feed devices. Similarly, the fluid-dynamic performance of narrow gas and liquid densities at high pressure are also industrially interesting. Examples of industrial applications are the high-pressure removal of methane and ethane. At very low surface tensions and under higher pressures, the separation efficiency of trays and packing materials as well as that of droplet separators are often significantly reduced which repeatedly creates cases for troubleshooting in industry.

The operation of mass transfer towers under fouling and foaming conditions represents another industrial challenge. As described above, research in this area has already begun but should be developed further.

6 Conclusions

Mass transfer column research at the university level, as described in the above sections, is still essential. As indicated in this article, continuous research activity is important to generate thermodynamic equilibrium data for modern gas/liquid or gas/liquid/liquid systems as well as the prediction of their physical, thermal, and kinetic properties. Standardization of experimental test facilities for distillation, absorption, desorption, and liquid-liquid extraction (including the test procedures) should be another topic of research. It is important that these standards are approved by all parties involved (universities, test institutes, vendors, and applying industry). Test results with standard packing materials and trays have to be agreed upon as well. The result of such standardization should be improved prediction of column performance data (pressure drop, capacity limit, liquid holdup, mass transfer efficiency, etc.) by modeling.

The fluid dynamics of non-separating column internals (feed pipe arrangements, distributor designs, flashing feed devices, etc.) should be investigated with the same priority because these parameters are as important for proper column performance as trays and packing materials.

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Symbols used

a	[m ² m ⁻³]	specific dry surface area of packing
a_1, a_2	[-]	constants
b_1	[-]	constant
a_{ph}	[m ² m ⁻³]	specific effective surface area of packing
C_S	[m s ⁻¹]	capacity factor = $u_V(\rho_V/(\rho_L-\rho_V))^{1/2}$

C_1, C_2	[-]	constants
D_S	[m]	column diameter
F_V	$[\text{m s}^{-1}(\text{kg m}^{-3})^{1/2}]$	gas capacity factor = $u_V(\rho_V)^{1/2}$
H	[m]	section height
h_L	$[\text{m}^3\text{m}^{-3}]$	superficial liquid holdup
HETP	[m]	height equivalent to a theoretical plate
HTU _{OV}	[m]	overall gas side height of a transfer unit
n_{th}	[-]	number of theoretical stages
n_p	[-]	number of practical stages
p	[bar]	pressure
$\Delta p/H$	$[\text{mbar m}^{-1}]$	specific pressure drop
u_L	$[\text{m s}^{-1}]$	superficial liquid velocity
u_V	$[\text{m s}^{-1}]$	superficial gas velocity
x_{eth}	$[\text{mol mol}^{-1}]$	mole composition of ethanol

Greek symbols

$\beta_{V, \text{Ph}}$	$[\text{s}^{-1}]$	volumetric mass transfer coefficient in gas phase
$\beta_{L, \text{Ph}}$	$[\text{s}^{-1}]$	volumetric mass transfer coefficient in liquid phase
ε	$[\text{m}^3\text{m}^{-3}]$	void fraction of packing
ρ_L	$[\text{kg m}^{-3}]$	liquid density
ρ_V	$[\text{kg m}^{-3}]$	gas density

Subscripts

exp	experimental value
Fl	flooding condition
L	liquid phase
R	reaction
V	vapor phase

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