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Term Projects (5): Mass Transfer Operations

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Term Project 28.1

An Improved Absorber Design Procedure

One of the authors [1–3] has developed a procedure that enables one to quantitatively outline how to size (diameter, height) a packed tower to achieve a given degree separation without any information on the physical and chemical properties of a gas to be absorbed. To calculate the height, one needs both the height of a gas transfer unit H_{OG} and the number of gas transfer units N_{OG} (see also Chapter 7). Since equilibrium data are not available, assume that m (slope of equilibrium curve) approaches zero. This is not an unreasonable assumption for most solvents that preferentially absorb (or react with) the solute. For this condition:

$$N_{OG} = \ln\left(\frac{y_1}{y_2}\right) \quad (28.1)$$

where y_1 and y_2 represent inlet and outlet concentrations, respectively. Since it also reasonable to assume the scrubbing medium to be water or a solvent that effectively has the physical and chemical approaching that properties of water, H_{OG} can be assigned values usually encountered for water systems. These are given in Table 28.1. For plastic packing, the liquid and gas flow fluxes are both typically in the range of 1500 – 2000 lb/(h·ft² of cross-sectional area). For ceramic packing, the range of flow rates is 500 – 1000 lb/h·ft². For difficult-to-absorb gases, the gas flow rate is usually lower and the liquid flow rate higher. Superficial gas velocities (velocity of the gas if the column is empty) are in the 3 – 6 ft/s range. The height Z may then be calculated from

Table 28.1 Packing Diameter versus H_{OG}

Packing diameter, inches	Plastic packing H_{OG} , feet	Ceramic packing H_{OG} , feet
1.0	1.0	2.0
1.5	1.25	2.5
2.0	1.5	3.0
3.0	2.25	4.5
3.5	2.75	5.5

$$Z = (H_{OG})(N_{OG})(SF) \quad (28.2)$$

where SF is a safety factor, the value of which can range from 1.25 – 1.5. Pressure drops can vary from 0.15 – 0.40 inch H₂O/ft packing. Packing size increases with increasing tower diameter. Packing diameters of 1 inch are recommended for tower diameter in the 3 ft range. One should use large packing for larger-diameter units; for smaller towers, smaller packing is usually employed [4,5].

You have been requested to improve the procedure originally proposed by Theodore [2]. In effect, you are being asked to convert the semi-qualitative procedure; i.e., the calculation for N_{OG} , H_{OG} , ΔP , and D need to be improved.

Term Project 28.2

An Improved Adsorber Design Procedure

There are several factors to be considered in the design of an adsorber [6].

1. The adsorbent particle size
2. The physical adsorbent bed depth
3. The gas velocity
4. The temperature of the inlet gas stream and the adsorbent
5. The solute concentration to be adsorbed
6. Any solute concentration(s) not to be adsorbed, including moisture
7. The required or desired removal efficiency
8. Possible polymerization on the adsorbent
9. The frequency of operation
10. Regeneration conditions
11. The system pressure.

Refer to the Overview in Part II, Chapter 7. Employing the above considerations/factors, Theodore [7,8] developed a rather simplified overall design procedure for a carbon system adsorbing an organic that consists of two horizontal units (one on/one off) that are regenerated with steam [9,10].

Dissatisfied with the semi-qualitative nature of Theodore's [7] design procedure, your boss has requested that the procedure not only be improved both technically and quantitatively but also be applicable for vertical columns, more than two units, and all methods of regeneration.

Term Project 28.3

Multicomponent Distillation Calculations

The Fenske-Underwood-Gilliland (FUG) procedure for sizing multicomponent distillation columns is well documented in the literature. Theodore and Ricci [11] provide the following.

The Fenske equation is used to calculate the minimum number of theoretical stages when the column is being operated under total reflux. While many forms of the Fenske equation have been presented in the literature, Equation (28.3) is preferred because it is highly useful from a design perspective.

$$N_{\min} = \frac{\ln \left[\left(\frac{r_{LK}}{1-r_{LK}} \right) \left(\frac{r_{HK}}{1-r_{HK}} \right) \right]}{\ln \alpha_{LK}} \quad (28.3)$$

where N_{\min} = minimum number of theoretical stages (including partial reboiler)

$$\alpha_{LK} = \sqrt{\alpha_{D,LK} \alpha_{B,LK}}, \text{ a geometric mean of light key relative volatilities}$$

$$r_{LK}, r_{HK} = \text{fractional recoveries}$$

While the Fenske equation calculates the minimum number of equilibrium stages required for separation at total reflux, Underwood developed equations that estimate the minimum reflux ratio. The Underwood equations are a set of two mathematical expressions that are generally solved sequentially (unless there are one or more components in between the light and heavy keys, in which case they should be solved simultaneously in order to determine the correct root). These equations are listed below:

$$N_{\min} = \sum_{i=1}^N \frac{\overline{\alpha}_i x_{F,i}}{\overline{\alpha}_i - \Theta} = 1 - q; \quad (1 < \Theta < \overline{\alpha}_{LK}) \quad (28.4)$$

$$R_{\min} = \sum_{i=1}^N \frac{\overline{\alpha}_i x_{D,i}}{\overline{\alpha}_i - \Theta} - 1 \quad (28.5)$$

where q = the q -factor, dependent upon the thermal condition of the feed
 Θ = root of the first Underwood equation (28.4)

$$\bar{\alpha}_i = \sqrt{\alpha_{D,i} \alpha_{D,i}}, \text{ mean relative volatility of the } i\text{th component}$$

Now that both Fenske and Underwood equations have been utilized to determine N_{min} and R_{min} , respectively, the last step in the FUG procedure is to employ the Gilliland correlation in order to determine the number of theoretical trays. The Gilliland correlation is shown in Figure 28.1 [12].

As is evidenced by Figure 28.1, one need only know the minimum reflux ratio and the operating reflux ratio in order to compute the abscissa of the correlation. The corresponding ordinate is then read from the plot. Since the minimum number of theoretical stages is known, the actual number of theoretical stages, N_p may be calculated. However, it should be noted that the Gilliland correlation was derived for systems with nearly constant relative volatilities throughout the column. Therefore, the Gilliland correlation may not be a suitable short-cut method for non-ideal systems, which relative volatilities may vary drastically.

While the Gilliland correlation itself is immensely useful for use as a shortcut method in column quick-sizing, it is inconvenient in that the graph must be read manually. In order to make the Gilliland correlation more applicable to computer programming, several analytical expressions

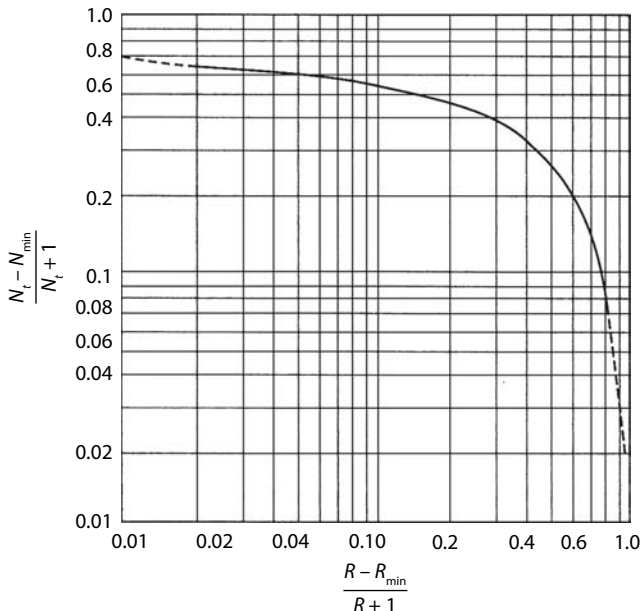


Figure 28.1 Gilliland correlation.

have been developed. One such outstanding correlation, as suggested by Chang, [13] is provided in Equation 28.6 below:

$$Y = 1 - \exp \left[1.490 + 0.315X - \frac{1.805}{X^{0.1}} \right] \quad (28.6)$$

where X is the abscissa (x -axis) of this correlation, given by

$$X = \frac{R - R_{\min}}{R + 1} \quad (28.7)$$

and Y is the ordinate (y -axis) of the correlation,

$$Y = \frac{N_t - N_{\min}}{N_t + 1} \quad (28.8)$$

which may be rearranged to:

$$N_t = \frac{N_{\min} + Y}{1 - Y} \quad (28.9)$$

where N_t = the number of theoretical stages.

Thus, Gilliland's graphical correlation may be replaced by using Chang's convenient mathematical expression.

The calculation of the number of theoretical trays is the last step in the FUG procedure. However, the sizing of distillation columns does not end there. To the contrary, there are several other considerations which must be taken into account in order to completely quick-size a distillation column, including: theoretical location of the feed tray (Kirkbride equation), calculation of the actual number of trays, calculation of column diameter, calculation of column height, etc.

In the addition to the FUG, the Kirkbride equation may be employed to determine the location of the *theoretical feed tray* and the O'Connell correlation [14] may be used to obtain a column's overall efficiency.

You have been assigned the task of developing an improved multicomponent distillation calculational procedure. Your revised equations need not necessarily be based on either the Fenske, or the Underwood, or the Gilliland correlations.

Term Project 28.4

A New Liquid-Liquid Extraction Process

Extraction is a term that is used for an operation in which a constituent of a liquid or a solid is transferred to another liquid (the *solvent*). The term *liquid – liquid extraction* describes the processes in which both phases in the mass transfer process are liquids. The term *liquid – solid extraction* is restricted to those situations in which a solid phase is present and includes those operations frequently referred to as *leaching*, *lixiviation*, and *washing* [15–18].

Extraction involved the following two steps: contact of the *solvent* with the liquid or solid to be treated so as to transfer the *solute* (soluble component) to the solvent, and separation or washing of the resulting solution. The complete process may also include a separate recovery procedure involving the solute and solvent; this is normally accomplished by another operation such as evaporation, distillation, or stripping. Thus, the streams leaving the extraction system usually undergo a series of further operations before the finished “product” is obtained; either one or both solutions may contain the desired material. In addition to the recovery of the desired product or products, recovery of the solvent for recycling is also an important consideration [15–18].

Liquid – liquid extraction is used for the removal and recovery of primarily organic solutes from aqueous and nonaqueous streams. Concentrations of solute in these streams range from either a few hundred parts per million to several mole/mass percent. Most organic solutes may be removed by this process. Extraction has been specifically used in removal and recovery of phenols, oils, and acetic acid from aqueous streams, and in removing and recovering freons and chlorinated hydrocarbons from organic streams [15–18].

Treybal [15] has described the liquid – liquid extraction process in the following manner. If an aqueous solution of acetic acid is agitated with a liquid such as ethyl acetate, some of the acid but relatively little water will enter the ester phase. Since the densities of the aqueous and ester layers are different at equilibrium, they will settle on cessation of agitation and may be decanted from each other. Since the ratio of acid to water in the ester layer is now different from that in the original solution and also different from that in the residual water solution, a certain degree of separation has occurred. This is an example of stage-wise contact and it may be carried out either in a batch or continuous fashion. The residual water may be repeatedly extracted with more ester to additionally reduce the acid content. One may arrange a countercurrent cascade of stages to accomplish

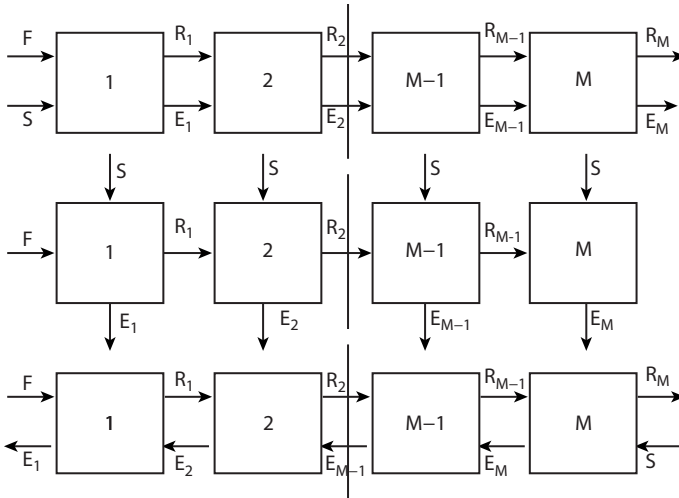


Figure 28.2 Multistage extractors.

the separation. Another possibility is to use some sort of countercurrent or crosscurrent continuous-contact device where discrete stages are not involved. These are pictured in Figure 28.2.

The solution whose components are to be separated is the *feed* (F) to the process. The feed is composed of a *diluent* and *solute*. The liquid contacting the feed for purposes of extraction is referred to as the *solvents* [19]. The solvent-lean, residual feed solution, with one or more constituents removed by extraction, is referred to as the *raffinate* (R). The solvent-rich solution containing the extracted solute(s) is the *extract* (E).

There are two major categories of equipment for liquid extraction. The first is *single-stage* units, which provide one stage of contact in a single device or combination of devices. In such equipment, the liquids are mixed, extraction occurs, and the insoluble liquids are allowed to separate as a result of their density differences. Several separate stages may be used in an application. Second, there are *multistage* devices, where many stages may be incorporated into a single unit. This type is normally employed in practice. (See also Figure 28.2).

In addition to *concurrent* flow (rarely employed) provided in Figure 28.2 for an n -stage system, *crosscurrent* extraction is a series of stages in which the raffinate from one extraction stage is contacted with additional fresh solvent in a subsequent stage. *Crosscurrent* extraction is usually not economically appealing for large commercial processes because of the high solvent usage and low solute concentration in the extract. Figure 28.2 also illustrates countercurrent extraction in which the extraction solvent enters

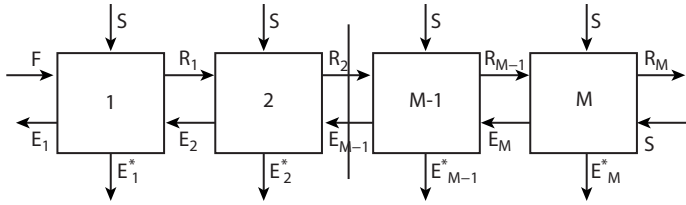


Figure 28.3 A hybrid extraction process

a stage at the opposite end from where the feed enters and the two phases pass each other countercurrently.

You have been assigned to a project that is concerned with developing a new liquid-liquid extraction process that integrates/combines both features of the countercurrent and crosscurrent processes. A diagram of one proposed process is provided in Figure 28.3 [19].

Term Project 28.5

Designing and Predicting the Performance of Cooling Towers

Cooling towers find application in industry. The same operation that is used to humidify air may also be used to cool water. There are many cases in practice in which warm water is discharged from condensers or other equipment and where the value of this water is such that it is more economical to cool it and reuse it than to discard it. Water shortages and thermal pollution have made the cooling tower a vital part of many plants in the chemical process industry. Cooling towers are normally employed for this purpose and they may be destined to have an increasingly important role in almost all phases in industry. Modern (newer) power-generating stations remain under construction or in the planning stage, and both water shortages and thermal pollution are serious problems that must be addressed [20,21].

The cooling of water in a cooling tower is accomplished by bringing the water into direct contact with unsaturated air under such conditions that the air is humidified and the water is brought approximately to the wet-bulb temperature. This method is applicable only in those cases where the wet-bulb temperature of the air is below the desired temperature of the exit water [22].

Qualitatively speaking, water is cooled in cooling towers by the exchange of sensible heat, latent heat, and water vapor with a stream of relatively cool dry air. The basic relationships developed for dehumidifiers also apply to

cooling towers although the transfer is in the opposite direction since the unit acts as a humidifier rather than as a dehumidifier of air [23].

Brown and Associates [24] have provided empirical correlations from the literature [25,26] for estimating (roughly) sizes and capacities of conventional towers.

One of the authors [27] has repeatedly claimed that no simple, easy-to-use procedure is available for either predicting the performance or the design of cooling towers. Employing the citations above and sound mass transfer theory, develop a cooling tower model/equation(s) that the practicing chemical engineer can apply.

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