

# We are IntechOpen, the world's leading publisher of Open Access books Built by scientists, for scientists

6,900

Open access books available

186,000

International authors and editors

200M

Downloads

Our authors are among the

154

Countries delivered to

TOP 1%

most cited scientists

12.2%

Contributors from top 500 universities



WEB OF SCIENCE™

Selection of our books indexed in the Book Citation Index  
in Web of Science™ Core Collection (BKCI)

Interested in publishing with us?  
Contact [book.department@intechopen.com](mailto:book.department@intechopen.com)

Numbers displayed above are based on latest data collected.  
For more information visit [www.intechopen.com](http://www.intechopen.com)



---

# Design of Fractionation Columns

---

Hassan Al-Haj Ibrahim

Additional information is available at the end of the chapter

<http://dx.doi.org/10.5772/57249>

---

## 1. Introduction

Distillation is a physical process used to separate a fluid mixture of two (binary) or more (multi-component) substances into its component parts. In most cases, the components to be separated are miscible liquids with different volatilities and boiling points. This separation process is a thermal unit operation that utilizes the differences of vapour pressure to produce the separation. In this process, the vapour or liquid mixture is heated whereby the more volatile components are evaporated, condensed and allowed to drip or drip apart, i.e. distil or destillare, as it was originally called in Latin. It is from this fact of 'Dripping' that the name of the distillation process was derived.

Distillation may be carried out in an intermittent or batch process or it may be carried out in a continuous steady-state process with a continuous feed stream. Distillation systems may also operate at different pressures, with higher pressures used for the separation of the more volatile materials and reduced pressures, in vacuum distillation systems, for heavier higher boiling-point materials, lowering thereby their boiling points.

Simple distillation of alcoholic beverages and petroleum has been practiced for a very long time. In the Middle Ages, Damascus played a part as a distilling centre, where distillation came into its own as the best and quickest way of obtaining pure chemical substances [1]. In the writings of the thirteenth century Syrian Scholar, Al-Dimashki, we find the earliest reference to the industrial distillation of crude oil or petroleum. Al-Dimashki says: "Many types of petroleum are water white by nature and so volatile that they can not be stored in open vessels. Others are obtained from a kind of pitch or bitumen in a turbid and dark condition, but by further treatment they can be made clear and white by distilling them like rose-water" [1].

An earlier account of crude oil distillation was given by Rhases (865-925) in his book *Sirr alasarar* (Secretum secretorum or Secret of secrets), and in 1512 Hieronymus Brunschwig wrote a book

on the true art of distillation (*Das Buch der wahren Kunst zu destillieren*). The theory of distillation, however, was first studied by Sorel in 1893 in his book on the distillation (Rectification) of alcohol, in which, interestingly, he says that the aim of distillation is the production (Fabrication) of alcohol and alcoholic liquors [2]. Mathematical analysis of the distillation (of binary mixtures) was first carried out in 1902 by Lord Rayleigh [3].

## 2. Fractional distillation

In common industrial jargon, distillation is used sometimes to mean fractional, and not merely simple, distillation. Fractional distillation, or fractionation, however, is in fact a special type of distillation, and as a separation technique, is much more effective than simple distillation and more efficient. In effect, fractionation is equivalent to a series of distillations, where the separation is achieved by successive distillations or repeated vaporization-condensation cycles<sup>1</sup>. Each vaporization-condensation cycle makes for an equilibrium stage, commonly known as a theoretical stage. A number of such theoretical stages may be required for the efficient fractionation and separation of the vapour or liquid mixture. The McCabe-Thiele method is a graphical method that may often be used to calculate the required number of theoretical stages. These theoretical stages will then have to be converted to actual plates or an equivalent packed height depending upon separation efficiency for a particular service.

In fractional distillation the components are separated through continuous heat and mass transfer between counter-current streams of a rising vapour and a descending liquid. As in all thermal separation processes, the motive force for the separation is the drive towards thermodynamic equilibrium between the different phases (VLE or vapour liquid equilibrium). This equilibrium is continuously disturbed by the mixing of the colder descending liquid and the hotter rising vapour, where the more volatile components of the descending liquid are vaporized and the less volatile components of the rising vapour are condensed and the driving force for the separation process is thereby maintained. The concentration of the lighter components will be greater in the vapour phase and conversely the concentration of the heavier components will be greater in the liquid phase, and only in the case of pure components or azeotropic mixtures will the equilibrium composition be the same in both phases.

Fractional distillation is much more recent than simple classical distillation. Its invention is often attributed to an illiterate French workman from Rouen, Edouard Adam by name, who succeeded in obtaining concentrated alcohol in a single distillation (*Seule distillation*). Although he patented his rectification still in 1801, he was not successful in commercializing his invention and several distillers built similar stills and he died poor in 1807 [4].

In recent years, fractional distillation has become one of the most important unit operations in chemical engineering industries. It is by far the most common specialized separation technology. It has been variously estimated that the capital investment in separation

---

<sup>1</sup> In fact, the term used in Arabic for fractional distillation (*Takreer*) is derived from this repeated vaporization-condensation cycles.

equipment is 40-50% of the total for a conventional fluid processing unit. In the USA, fractionators are used in 90 to 95% of the separations [5]. In 1992, Darton estimated the world-wide throughput of distillation columns at 5 billion tonnes of crude oil and 130 million tonnes of petrochemicals per year.

In petroleum refineries in particular, atmospheric fractionation of crude oil, variously known as topping, is the first and the most important process of the series of oil refinery operations [6]. The crude oil feedstock, which is a very complex multi-component mixture, is separated into a number of products or fractions, with each product or fraction being composed of groups of compounds within a relatively small range of boiling points. These straight-run products or fractions are the origin of the term fractional distillation or fractionation. Fractionation, however, is not used in the petroleum refining industry only, but is also used in other chemical, petrochemical, beverage and pharmaceutical industries and in natural gas processing plants.

Fractionation, though widely used, is one of the most energy intensive operations. In fact, distillation may be the largest consumer of energy in petroleum and petrochemical processing. Of the total energy consumption of an average unit, the separation steps accounts for about 70% and distillation alone can consume more than 50% of a plant's operating energy cost [7]. Fractional distillation accounts for about 95% of the total distillation consumption [8]. Back in 1988, speaking of distillation, Dr. Frits Zuiderweg wrote:

"Distillation is the most important and most visible separation technology. The skyline of many refineries and chemical plants is dominated by tall distillation towers, and it cannot be denied that their spatial arrangements often amount to architectural beauty. Less spectacular, but also visible are the smoke stacks of these industrial complexes; they represent the energy used in processing, the major part of which is consumed by distillation processes" [9].

Replacing fractionation by other separation techniques in such industries as petroleum and petrochemical industries is, however, highly unlikely despite its energy intensiveness. The pre-eminence of distillation for the separation of fluid mixtures is fundamental for both kinetic and thermodynamic reasons [9]. It may also be the most economical process for the separation and production of high-purity products. In general, distillation is favoured over other separation techniques when large rates are desired of products that are thermally stable with relative volatility greater than 1.2 and where no extreme corrosion, precipitation, sedimentation or explosion issues are present [10].

The lunar environment would offer some unique advantages for distillation processes, as has been pointed out in an interesting paper on fractional distillation in a lunar environment [11]. Due to the cryogenic temperatures available on the moon, vacuum distillation would be possible with lower heat loads and cleaner separations [11].

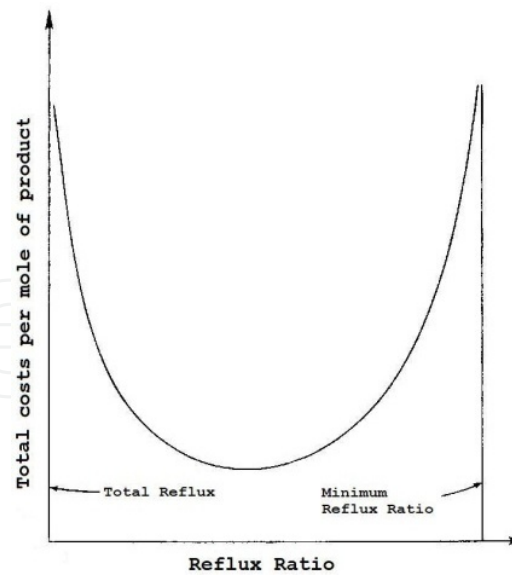
### 3. Fractionation columns

In normal practice, fractionation is carried out in isolated, vertical, cylindrical fractionation columns or towers using different types of contacting devices, with condensers at the top for

cooling and partially condensing the top products and reboilers at the bottom for heating and partial evaporation of the bottom products. Fractionation towers may also have several outlets at intervals up the column allowing for the withdrawal of different products having different boiling points or boiling ranges. When the liquid mixtures to be separated are similar in boiling points, spinning band distillation may sometimes be used in which a spinning helical band made of an inert material is used to push the rising vapours and descending liquids to the sides of the column, bringing them into close contact with each other and providing for a greater number of vaporization-condensation cycles and thereby speeding equilibration. Spinning band distillation was first used by Harold Clayton Urey for the separation of different compositions of isotopic water, but was later used in normal, non-isotopic, distillations, where it proved to be much superior to other distillation techniques.

In industrial fractionation towers, reflux is normally used to improve the efficiency of separation of the fractionation products. Reflux refers to the portion of the condensed overhead product that is returned to the tower. The reflux flowing downwards provides the cooling required for condensing the vapours flowing upwards. The reflux ratio, which is the ratio of the (internal) reflux to the overhead product, is conversely related to the theoretical number of stages required for efficient separation of the distillation products, with the two extreme theoretical and impractical possibilities being the use of a minimum reflux ratio with an infinite number of stages or else the use of total reflux with a minimum number of stages. From the standpoint of design, total reflux can be thought of as a column of infinite diameter operating at infinitely large vapour and liquid rates [12]. In practice, the optimum choice of the reflux ratio and the number of theoretical stages is based on economical considerations. Increasing the reflux ratio reduces the number of theoretical stages and thus reduces the construction cost of the column. On the other hand, increasing the reflux ratio leads at the same time to an increase in both the construction and operating costs of the fractionation columns, as increasing the reflux ratio leads to an increase of the column diameter and a corresponding increase of the size of coolers and condensers required for cooling and condensing the reflux (Fig. 1). Most columns are designed to operate between 1.2 to 1.5 times the minimum reflux ratio because this is approximately the region of minimum operating costs [13]. For an existing distillation tower with a fixed number of stages, the reflux ratio can be used to influence the steady-state operation of the unit and maximize the operating profit [8].

Fractionation columns are generally composed of two sections: an upper rectifying section and a lower stripping section. Rectification in general is used to mean purification or purification by distillation in particular, and in other languages, such as French (Rectification) or German (Rektifikation), it is used to refer to fractionation. In the strict technical sense, however, rectification in English refers only to that part of the fractionation process in which the concentration of the more volatile components is increased in the upper section of the fractionation column. Stripping, on the other hand, refers to the part of the fractionation process in which the more volatile components are stripped from the liquid in the lower section of the fractionation column or, sometimes, in a separate stripping column or stripper. In most stripping sections or columns, steam is normally used to strip the volatile components from the descending liquid.



**Figure 1.** Total costs for a specified separation [12]

The two major types of contacting devices used in fractionation columns or towers are packing and trays. Packed columns, also called continuous-contact columns, use packing that may be grid, random dumped or structured (sheet metal) packing, while tray columns, also called staged-contact columns, use trays or plates. Trays used in tray columns can be of many different types and designs including not only conventional trays such as bubble-cup, perforated and valve trays but also baffle, tunnel, high capacity, dual-flow and multiple downcomer trays. The first vertical columnar continuous distillation still was developed by Collier-Blumenthal in France in 1813, but tray columns, along with feed preheating and the use of internal reflux, first appeared on the scene in the 1820s. The first generation of random packings was developed as early as 1907, but structured packing was only developed in the 1940s.

Packing is usually preferred for loads that are temperature sensitive or corrosive, as the range in packing materials is wider than that commonly available for trays. Packing is further preferred for foaming systems where the low gas and liquid velocities in packing suppress foam formation. Packing pressure drop is also much lower than trays. Kister et al. states that packing pressure drop is typically three to five times lower than that of trays [14]. For this reason, packing columns are favoured in vacuum services and wherever pressure drop is important or where it is economical to minimize it. In addition, packing columns are normally used for lower loads because of the mass transfer characteristics of packing. The mass transfer in packing applications takes place on a thin film of liquid that is spread over the surface area of the packing. If the liquid rate is high this boundary layer will increase, reducing the mass transfer. Packing columns are also used for smaller towers, with diameters less than 3 ft, as it is difficult to access such towers from the inside in order to install and maintain the trays. Although an oversized diameter may be used to overcome this difficulty, packing is normally a cheaper and more desirable alternative [9, 10, 15].



Tray columns, on the other hand, are used for larger columns with high liquid loads. With lower liquid loads trays can have high residence times leading to undesired effects such as fouling and sedimentation. In addition trays will have difficulty maintaining a good weir loading and distribution across the tray, resulting in lower than expected tray efficiencies. For such reasons, it is often more economical to handle high liquid loads in tray columns. Whereas packings are prone to severe maldistribution problems in large-diameter columns, these problems are far less severe in tray columns. Tray columns are also to be preferred where fouling or solid accumulation and deposition is anticipated because of the long history of success trays have had in fouling service applications. Trays can handle solids a lot easier than packed columns, as solids tend to accumulate in packing voids, but the higher gas and liquid velocities on trays provide a sweeping action that keeps tray openings clear. In addition, cleaning trays is easier than cleaning random packing, while cleaning structured packing is practically impossible. In many situations, tray columns are favoured because of performance prediction which is generally much better with tray columns. A major difficulty often encountered in packed-tower design is the inability to predict performance. This difficulty was, for many years, circumvented by grossly oversizing packed towers. Finally, tray columns weigh less than packed columns, which mean lower costs for foundations, support and column shell, the cost of the column being usually estimated on the basis of weight [9, 10, 15]. Tower casing cost, for example, can be estimated from this equation, in USA dollars [16]:

$$\text{Tower casing cost} = 8750 \left( \frac{wt, lb}{6000} \right)^{0.75}$$

In most oil refinery operations, tray columns are mainly used for the separation of petroleum fractions at different stages of oil refining [6]. Apart from the general advantages, already referred to, of using trays for larger columns and higher liquid loads, other factors as well argue for the use of tray columns in the oil industry. These include better performance prediction, feed composition variation and the need to incorporate side drawoffs, reboilers, condensers and other complexities. Feed composition variation is generally allowed for by installing alternate feed points, but in packed towers, every alternate feed point requires expensive distribution equipment. Complexities are also more easily incorporated in tray than in packed columns where every complexity requires additional distribution and/or liquid collection equipment [9, 15].

#### 4. Design of fractionation columns

Fractionation columns are designed to achieve the required separation of fluid mixtures or miscible liquids efficiently. The energy intensiveness of distillation added to the continued increase in energy costs has made it of paramount importance to develop rapid calculation procedures for the design of distillation columns.

In general, the design methods for fractionation columns are similar to the design methods used for other separation processes such as absorption and extraction. With reference to the different types of fractionation columns, design procedures for tray columns are better

established than for packed columns. This is particularly so with respect to separation efficiency since tray efficiency can be estimated more accurately than packed height equivalent to a theoretical plate (HETP) [17].

The design of fractionation columns is normally made in two steps; a process design, followed by a mechanical design. The purpose of the process design is to calculate the number of required theoretical stages and stream flows including the reflux ratio, heat reflux and other heat duties. The purpose of the mechanical design, on the other hand, is to select the tower internals and calculate column diameter and height. For the efficient selection of tower internals and the accurate calculation of column height and diameter, many factors must be taken into account. Some of the factors involved in design calculations include feed load size and properties and the type of distillation column utilized. This phase of column design has a major impact on column costs, for the choice of internals influences all costs of the distillation system including the column, attendant structures, connecting piping and auxiliaries such as reboiler, condenser, feed heater, and control instruments [15].

Over the years, a considerable amount of work has been done to develop efficient and reliable computer-aided procedures for the mechanical design of fractional distillation columns. Reliable design methods for fractionation columns and other equipment have been developed by commercial organisations such as Fractionation Research Inc. (FRI), but their work is of a proprietary nature and is not freely available in technical literature. Besides being costly to license, using the available rigorous commercial simulation programmes of fractional distillation columns often requires some time to learn. Short-cut design calculation methods have also been presented but mainly for specific and limited applications such as for binary or pseudo-binary systems [18, 19]. In most cases, however, the correct design of multi-component fractionation columns with multiple feeds and side-streams is not straightforward and is not always a simple task, particularly in the oil industry where the fractions to be separated are not simple mixtures composed of two or three components but are complex mixtures of varying composition and properties and which are composed of tens or hundreds of different compounds.

In the oil refining industry, the design and operation of fractionation towers is still largely accomplished on an empirical basis. The calculations involved in the design of petroleum fractionation columns require in the usual practice the use of numerable charts, tables and complex empirical equations. Such charts, tables, equations and other design aids have long been available in the literature, but their use is often lengthy and tedious, and the results are not always dependable [6]. Using Matlab computer programming in the design of fractionation columns can, therefore, be of great help and assistance in facilitating the design calculations and improving their accuracy and dependability. The use of charts, tables and empirical equations can then be dispensed with and the different available calculation methods can be directly compared and evaluated in order to achieve better design results. Using Matlab computer programming is, furthermore, time saving which is often an important consideration from a designer's perspective. In the remaining part of this chapter, the clear advantages of using Matlab programming in the design calculations of fractionation columns will be illustrated by the calculation of column diameters.



## 5. Calculation of column diameter

In industrial applications, diameters of fractionation columns vary greatly and may range from about 65 cm in smaller towers to about 6 m and more in larger columns, even up to 15 m or 50 feet in some applications [12]. In general, column diameter is relatively insensitive to changes in operating temperature or pressure. Nonetheless, the proper sizing of the column diameter is essential for the efficient performance and operation of fractionation columns. If the column diameter is not sized properly, operational problems may occur and the desired separation will not be achieved. Proper sizing of the column diameter is also crucial for other economic considerations as the costs of fractionation equipment are markedly influenced by the column diameter. The investment cost of a distillation column is a function of its diameter, and so is the cost factor for capacity for the estimation of the cost of trays [20]. For sieve trays, for example, the cost, in USA dollars, has been estimated as equal to  $151 \left(\frac{D}{3}\right)^{1.63}$  where D is the tower diameter in feet [16].

The accurate calculation of column diameter will depend on many factors of column design and, conversely, many factors of column design and construction will depend on the column diameter. Furthermore, in the choice of the contacting devices to be used in fractionation columns, the column diameter is the main factor to consider. As was indicated above, packing is preferred for smaller towers while trays are mainly used in larger columns, with diameters greater than 3 ft. The use of tray columns with diameters in the 1-ft, 6-in. to 2-ft range is not usually economical and a packed tower in such cases will prove the best economically. On the other hand, packed towers are not limited to small units and the use of larger-diameter packing columns may still provide the less expensive choice for some specific applications [21].

In packed columns, some of the ultimate performance depends on the column diameter. When random packings are used, there are many reports of an increase in HETP with column diameter [15]. Tower diameter is related also to the packing size, and Robbins recommends that the tower diameter should be at least 8 times the packing size, and if not, the diameter calculations should be repeated with different packing. The ratio of maximum random packing size to tower diameter depends, however, on the type of packing used. Although the 1:8 ratio is in more common use for most packings, some types of packing require a larger ratio of about 1:15, and recent data indicates that Raschig rings require a ratio approaching 1:20 [7, 21].

In tray column design, the calculation of the column diameter is of special significance. Factors of column design that depend on the column diameter include such factors as the choice of plate type and liquid flow arrangement. The choice of plate type is often related to column diameter. Using dual-flow trays, for example, has numerous advantages over other types of trays, particularly in heavy fouling services, but their use in larger diameter towers may lead at the same time to maldistribution. The choice of both liquid flow, reverse or direct flow, and number of passes, single or multiple pass, depend on the column diameter. The common design practice is to minimize the number of passes, and trays smaller than 1.5-m diameter seldom use more than a single pass; those with 1.5- to 3-m diameters seldom use more than two passes. Four-pass trays are only used in towers larger than 5-m diameter [15]. In general,

increasing column diameter leads to liquid flow rate increase as the square of the diameter. As the area available for liquid flow also increases in proportion to the diameter, the liquid load per unit length of outlet weir will increase in proportion to the column diameter. In larger column diameters, liquid level difference across the tray increases leading to inefficient mixing of liquid and vapour and trays may have to be fitted with multiple downcomers to reduce the liquid load across each active area section. This reduces the weir load and liquid head on the tray deck resulting in higher vapour capacity, lower pressure drop and improved operating turndown range. Tray efficiency, which typically varies in well-designed trays between 40 and 120%, may be affected by the column diameter. Because of a cross-flow effect, higher efficiency values may be achieved in large-diameter towers [17]. Finally, trays for columns larger than about 900 mm in diameter are normally manufactured in sections sized to fit through the column manways and are assembled inside the column.

Calculation of the column diameter, on the other hand, will in turn depend on many factors of column design such as plate spacing, the type of tray used in the column, surface tension and other physical properties of the vapour and liquid at the tray conditions and other factors. The interrelationship of these different factors is not clearly understood, and diameters are, therefore, determined by relations correlated by empirical factors [21].

Greater plate spacing raises capacity, leading to a smaller tower diameter, but at the same time it raises tower height. In this case, an economic tradeoff is to be considered between tower height and diameter. With accessibility in mind, plate spacing should be from 450 to 600 mm. In towers with larger than 1.5-m diameter, tray spacing is typically 600 mm to allow easy access for maintenance, but in very large towers (>6-m diameter), tray spacings of 750 mm are often used. For a tower diameter below 1.5 m, on the other hand, a tray spacing of 450 mm is adequate as the column wall can be reached from the manway [7, 15].

The factors influencing the different tray types are somewhat different. Column diameter is a function of both tray efficiency and its capacity characteristics as represented by velocity effects including entrainment, and the pressure of the operation. Low-pressure operations require generally larger diameter Columns [21].

The reflux ratio is another factor that affects the calculation of tower diameter. Increasing the reflux ratio means more vapour being boiled up and the required column diameter will be greater.

The principal factor that determines the column diameter, however, is the vapour flow rate. The minimum vapour flow rate required is determined by weeping, but the flooding condition fixes the upper limit of vapour velocity. A high vapour velocity will be needed for high plate efficiencies, but the vapour velocity must at the same time be below that which would cause excessive liquid entrainment or a high pressure drop. In general, vapour velocity will normally be between 70 to 90 per cent of that which would cause flooding. For design purposes, a value of 80 to 85 per cent of the flooding velocity should be used [22].

Calculation of the column diameter is often based on a trial and error approach. The preliminary diameter calculated may have to be refined by checking against the performance correlations of the column. In any case the calculated diameter should also be rounded

appropriately for fabrication standardisation. A number of general rules of thumb may be used as a general guide before carrying out the actual calculations, but these rules often have exceptions and may not apply in all cases. Such rules as related to column diameter include the following four rules [10]:

1. The length to diameter ratio should be less than 30, preferably below 20, and tower height is to be limited to 60 meters because of wind load and foundation concerns. If the tower is higher than 60 m, then a design with smaller tray spacing should be considered [7, 23].
2. The ratio of tower diameter to random packing size is greater than 10.
3. The tower diameter should be maintained at 1.2 meters at the top for vapour disengagement.
4. The tower diameter should be maintained at 2 meters at the bottom for liquid level and reboiler return.

In normal practice, however, only one diameter is calculated for the whole column. Different column diameters would only be used where there is a considerable change in flow-rate. Changes in liquid rate can be allowed for by adjusting the liquid downcomer areas [22]. If two or more diameters are calculated, say for the top and bottom sections of the column, then roughly speaking, when the difference in the calculated diameters exceeds 20%, different diameters for the top and bottom sections are likely to be economical and sections having different diameters should be at least 600 cm (20 ft) in length. Otherwise the diameter should be uniform. The preliminary column diameter would then be the larger of the two calculated diameters [9].

## 6. Calculation procedures for column diameter

The equation most often used for the calculation of column diameter is based on the well-known empirically-correlated Souders and Brown equation. Equation 1, or the chart of Fig. 2, may then be used to estimate the maximum allowable superficial vapour velocity, and hence the column area and diameter.

$$U_V = \frac{K}{3600} \left( \frac{\rho_L - \rho_V}{\rho_V} \right)^{0.5} \quad (1)$$

Where:

$U_V$  = Maximum allowable superficial vapour velocity (ft/sec or m/sec)

$\rho_L$  = Liquid density (kg/m<sup>3</sup> or lb/ft<sup>3</sup>)

$\rho_V$  = Vapour density (kg/m<sup>3</sup> or lb/ft<sup>3</sup>)

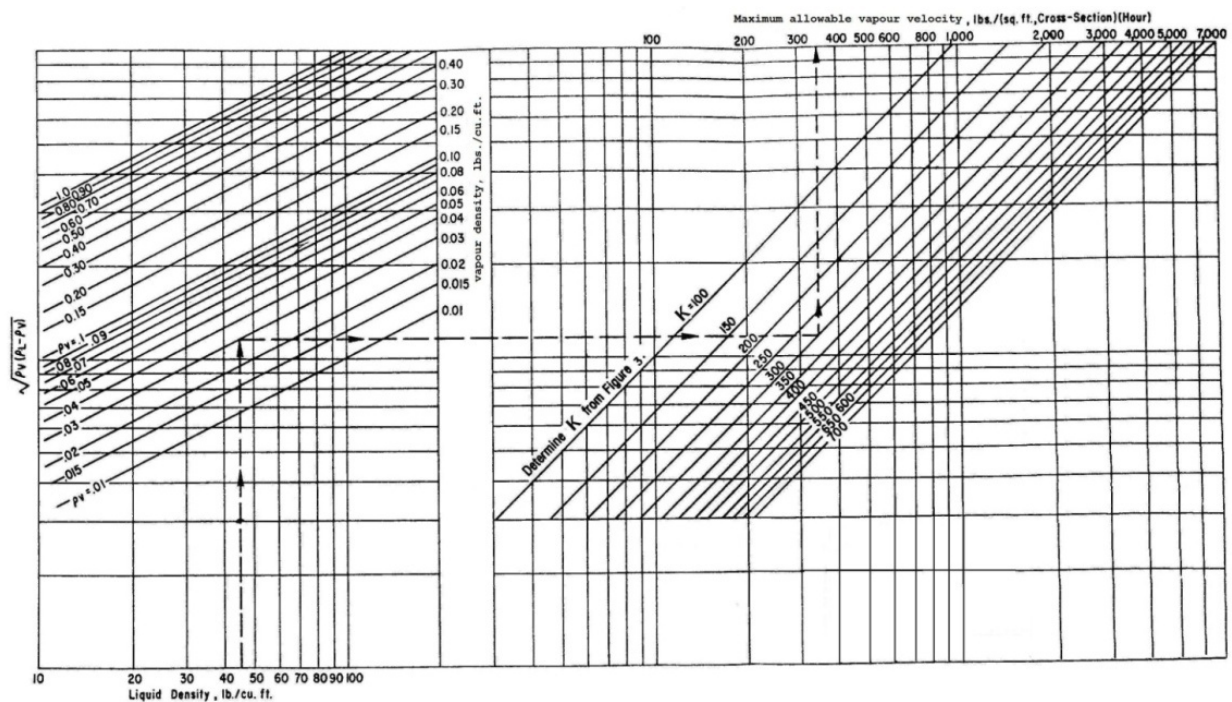
K in the above equation is an entrainment-related factor that depends on several variables which differ from one correlation to another. From a consideration of the balance of forces acting on the liquid droplets, the acceleration due to gravity must be included in the K factor, although it does not appear as a separate parameter in the above equation. This is to be explained by the fact that the Souders and Brown equation is based on empirical correlation of experimental data and has no theoretical derivation. The other main variables that affect the K factor include plate spacing, type of tray, physical properties, liquid load, hole diameter and fractional hole area.

Of such variables, plate spacing is the most important. In general, the K factor rises with plate spacing, where it is proportional to plate spacing to the power of 0.5 to 0.6. At low plate spacing (<38 cm) the power maybe somewhat higher due to the proximity of the froth envelop and/or excessive splashing from dispersion at the plate [9]. For the calculation of the K factor in terms of plate spacing, the following equation may be used:

$$K = 3600 \left( -0.171 T^2 + 0.27 T - 0.047 \right) \quad (2)$$

Where:

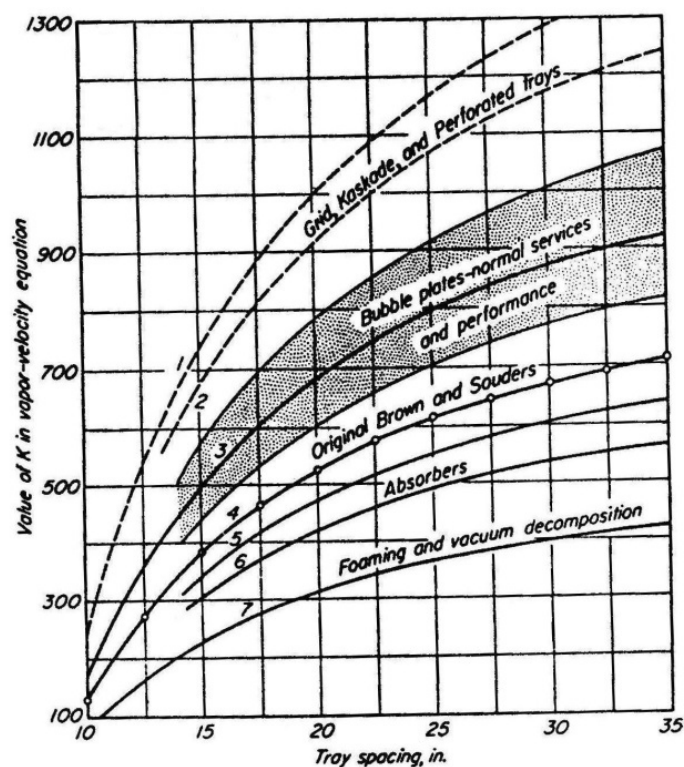
T = plate spacing (m).



**Figure 2.** Allowable vapour velocity for fractionation, absorption, and stripping columns.

This K factor may also be estimated from charts available in the literature such as shown in Figure 3 (curve 3, for normal performance of bubble plates through normal range of liquid





**Figure 3.** Effect of tray spacing and type of service allowable vapour load in fractionators [24]

loads at atmospheric and higher pressures, or curve 2 for other tray types). Using the value of the K factor from this chart often gives a lower estimate of the column diameter as compared to the other charts and procedures. This will be clearly seen when the K factor from this chart is used in the calculation of the two fractionators given later on in this chapter.

Although it has been suggested by certain workers that the K factor is at most a weak function of physical properties, there are several correlations that tend to contradict this suggestion [9]. Surface tension, in particular, may have a significant effect on the value of the K factor, as evidenced by such charts as are shown in figures 4 and 5. Equation 2 may also be modified to include the effect of surface tension for bubble cap trays as follows:

$$K = (36.71 + 5.456 T - 0.08486 T^2) \ln \sigma - 312.9 + 37.62 T - 0.5269 T^2$$

Correlation ranges are:

$K = 0$  to  $700$

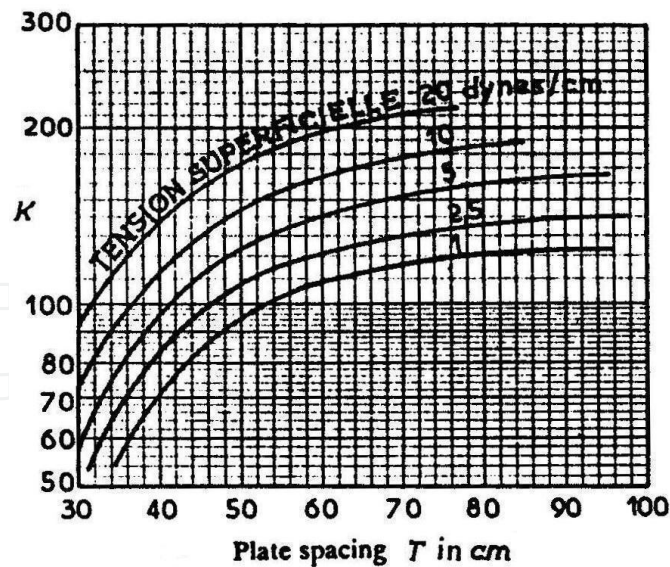
$\sigma$  = Surface tension (dynes/cm)  $0.1$  to  $100$ .

$T$  = plate spacing (in) =  $18$  to  $36$ .

The K factor increases with fractional hole area increase and hole diameter decrease. Roughly speaking, it increases with the reciprocal of hole diameter to a power of  $0.1$  to  $0.2$  [9].

Finally, liquid load can have a significant effect on the value of the K factor. As liquid load increases, the K factor first rises then declines [9].





**Figure 4.** Coefficient of entrainment as a function of plate spacing and surface tension [25]

Taking all the above effects into consideration, the K factor may be calculated (for valve and sieve trays) using the Kister and Haas Correlation [9]:

$$K = 0.144 \left( \frac{d_H^2 \sigma}{\rho_L} \right)^{0.125} \left( \frac{\rho_V}{\rho_L} \right)^{0.1} \left( \frac{T}{h_c} \right)^{0.5}$$

Where:

$d_H$  = Hole diameter (in)

$\rho_L$  = Liquid density (kg/m<sup>3</sup> or lb/ft<sup>3</sup>)

$\rho_V$  = Vapour density (kg/m<sup>3</sup> or lb/ft<sup>3</sup>)

$h_c$  = Clear liquid height (in).

In Fair's correlation, the value of the K factor is a function of the flow parameter  $F_{lv}$ , plate spacing, surface tension and fractional hole area (Fig. 6), where:

$$F_{lv} = \frac{L}{V} \sqrt{\frac{\rho_v}{\rho_L}} \quad (3)$$

$L$  = Liquid flow rate (lb/h)

$V$  = Vapour flow rate (lb/h)

Fair's correlation was recommended by most designers, but, according to some workers, it may be on the conservative side.

The K factor is normally determined at the top and bottom (or intermediate) positions of the column in order to evaluate the point of maximum required diameter.

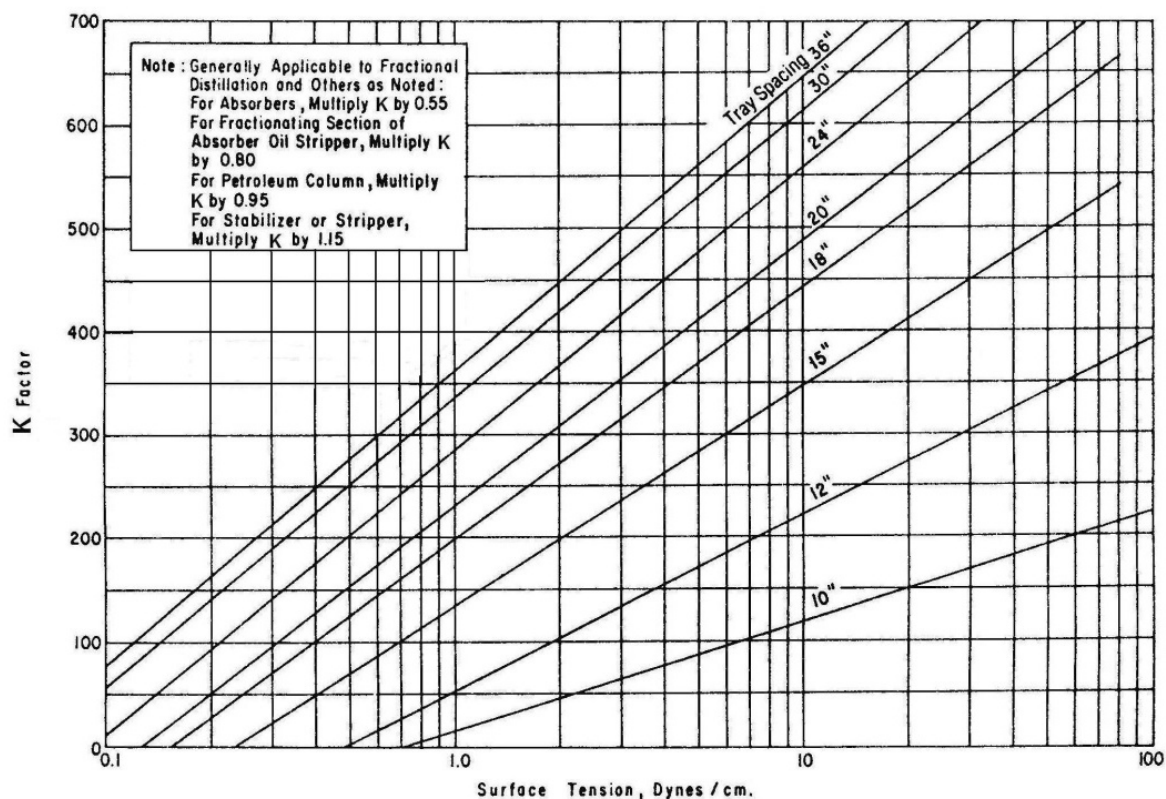


Figure 5. K factors for column diameter using bubble cap trays

It has been suggested that, for perforated trays in particular, the Souders-Brown method is rather conservative and that a better first approximation and often a more economical tower diameter is determined using the following equation for the calculation of vapour velocity [21]:

$$e_w = 0.22 \left( \frac{73}{\sigma} \right) \left( \frac{U_v}{T'} \right) \quad (4)$$

where:

$e_w$  = Weight of liquid entrained/unit weight of vapour flowing in perforated tray column. Entrainment values of 0.05 lbs liquid/lb vapour are usually acceptable, with 0.001 and 0.5 lb/lb being the extremes.

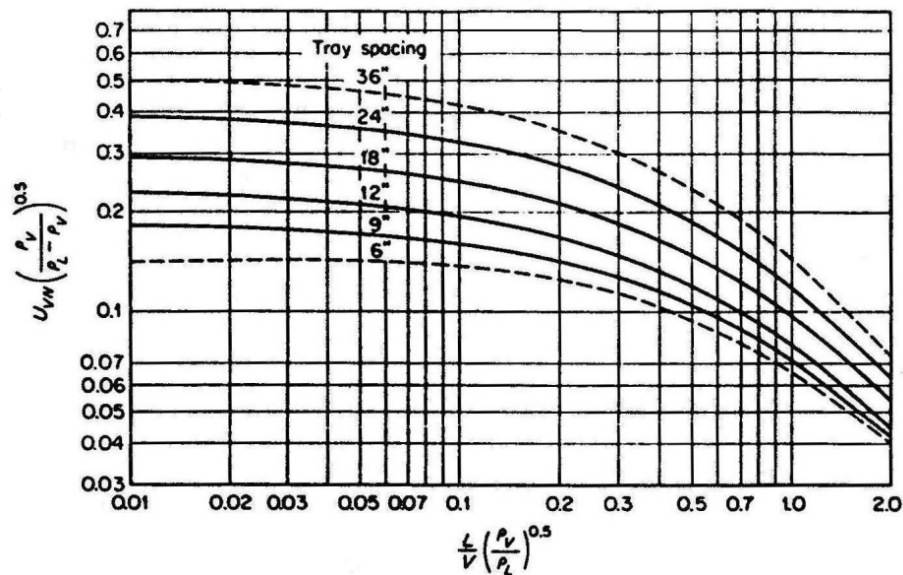
$T'$  = Effective tray spacing, distance between top of foam and next plate above

$$= T - 2.5 h_c$$

$h_c$  = Height of clear liquid in bubbling zone. This is based on an aerated mixture density of 0.4 that of the clear liquid on the tray, and has been found to be a reasonable average for several mixtures.

Equation 4 was derived primarily for perforated trays with downcomers, but, because of the very close similarity between a perforated plate without downcomers and one with down-

comers, it can also be used for trays without downcomers, although the liquid level and foam-froth height will generally be higher on trays without downcomers. The value of  $h_{\text{c}}$ , clear liquid on the tray, may therefore range from 1 in. to 6 in. depending on the service [21].



**Figure 6.** Flooding limits for bubble-cap and perforated plates [26]

From the calculated vapour velocity  $U_v$ , the column diameter can then be calculated using the following equation:

$$D = \left[ \frac{4V}{3600\pi\rho_v U_v} \right]^{0.5} \quad (5)$$

Where:

$D$  = Column diameter (m or ft)

$V$  = Maximum allowable vapour flow rate (kg/h or lb/h)

The diameter obtained in this solution, corresponding to the maximum allowable flow rate, is the minimum acceptable diameter for operation with essentially no entrainment carryover from plate to plate. If irregularities in capacity, system pressure or other significant variables are anticipated, the maximum allowable flow rate can then be divided by a safety, or ignorance, factor of 1.10 to 1.25. Alternatively, the maximum flow rate may be multiplied by 1.05 to 1.15 for satisfactory operation at conditions tolerating some entrainment with no noticeable loss in fractionation efficiency or capacity.

In another procedure based on Fair's correlation that may also be used for the estimation of column diameter, the chart of Figure 6 may be used to determine the maximum vapour velocity relative to the net area for vapour flow ( $U_{\text{VN}}$ ) for a given value of flow parameter and plate

spacing, the net area being defined as the tower cross-sectional area minus the downflow area. This may be used for both bubble-cap plates and perforated plates with the following restrictions:

1. The distilling system is non-foaming or low-foaming type. Foaming makes liquid disengagement from vapour flow difficult, and when foaming becomes severe it can lead to a reduction in capacity and loss of efficiency. To counter this effect distillation equipments that operate in foaming systems are run at lower vapour, and liquid, rates to reduce the amount of froth generated.
2. The weir height is less than 15% of the tray spacing.
3. In perforated trays, the perforation diameters are equal to or less than 0.25 in. Entrainment may be greater with larger hole sizes.
4. The plate spacings of 6 and 9 in. do not apply to bubble-cap plates.
5. The ratio of slot area, or hole area, to active plate area must be equal to or greater than 0.1, and the system surface tension 20 dynes/cm.

For hole-to-active area ratios of less than 0.1, it is recommended by Fair that the vapour velocity from the chart be modified by the following factors:

$A_s/A_A$	Multiply $U_{VN}$ by
0.10	1.00
0.08	0.90
0.06	0.80

The base surface tension assumed in the chart of Figure 6 is 20 dynes/cm. The decrease in surface tension as equilibrium temperature rises promotes foaming in distillation systems, and, as mentioned earlier, distillation equipment may have to be run in foaming systems at lower vapour rates. For surface tensions different from 20 dynes/cm, the velocity from the chart is therefore modified by:

$$\text{Corrected } U_{VN} = U_{VN} \left( \frac{\sigma}{20} \right)^{0.2}$$

Alternatively, the maximum vapour velocity ( $U_{VN}$ ) from the chart of Fig. 6 may be used for estimating the flooding vapour velocity ( $U_f$ ). Taking the above considerations into account and assuming operation at 80% flooding, the following equation may then be used for calculating the flooding velocity.

$$U_f = 0.80 U_{VN} \cdot C_s \cdot C_F \cdot C_A$$

Where:

$C_s$  = Surface tension correction factor =  $\left(\frac{\sigma}{20}\right)^{0.2}$

$C_F$  = Foaming correction factor (1, for non-foaming systems; 0.50-0.75 for foaming systems)

$C_A$  = Hole-to-active area ratio correction factor (1, for  $A_s/A_A \geq 0.1$ ;  $5(A_s/A_A)+0.5$ , for  $0.06 \leq (A_s/A_A) < 0.1$ ). For valve trays  $C_A = 1$ .

The flooding vapour velocity is usually about 50-60% of the maximum vapour velocity value, and can be as high as 85% or even 95% if sufficient information is available about the system foaming characteristics.

The net area for vapour flow ( $A_N$ ) can then be calculated, and hence, using the flooding vapour velocity, the column diameter can be calculated.

Assuming that the downflow area ( $A_d$ ) is 10% of the total cross-sectional area of the column ( $A$ ),

$$A_N = 0.90 A = 0.90 \times 0.785 D^2 = \frac{Q_V}{3600 U_F}$$

$$D^2 = \frac{1.4}{3600} \frac{Q_V}{U_F}$$

Where:

$Q_V$  = Vapour volume flow rate ( $m^3/h$  or  $ft^3/h$ ).

A correlation similar to Fair's correlation, applicable to columns having bubble caps, sieve or valve trays, was later proposed by Smith et al [27]. In this correlation, settling height, defined as tray spacing minus clear liquid height in bubbling zone, was used as parameter instead of tray spacing, where:

$$\text{Clear liquid height} = h_c = h_w + h_{ow}$$

$h_w$  = Height of outlet weir.

$h_{ow}$  = Height of liquid over weir.

In a similar correlation by Gerster, this clear liquid height ( $h_c$ ) was used as a parameter [28]. The deviation in the flooding velocity data from manufacturers of valve trays, as reported by Smith et al., was six percent on average [27].

For checking designs one can roughly relate tower diameter (ft) to reboiler duty as follows [29]:

Situation	Reboiler Duty, MM/BTU/hr
Pressure distillation	$0.5 D^2$
Atmospheric pressure distillation	$0.3 D^2$
Vacuum distillation	$0.15 D^2$



In a lunar environment with its reduced gravity, fractionation column area and diameter would be larger, as both are inversely related to the acceleration due to gravity. In his paper on distillation in a lunar environment, Pettit concluded that the ratio of the column diameters on earth and moon will be proportional to the square root of the accelerations due to gravity [11].

$$\frac{A_m}{A} \propto \sqrt{\frac{g}{g_m}}$$

Where:

$A$  = Cross-sectional area of the column on earth

$A_m$  = Cross-sectional area of the column on the moon

$g$  = Gravity constant on earth

$g_m$  = Gravity constant on the moon.

Considering that the lunar gravity is about one sixth of the earth gravity, an earthly fractionation column of one-meter diameter would correspond to a lunar column of 1.6 m diameter.

## 7. MATLAB programming for column diameter calculation

For the calculation of the fractionation column diameter using the various charts and calculation methods available in the literature, a computer programme is written in Matlab (See appendix). Using this programme simplifies the procedure of the calculation of column diameter and dispenses with the tedious use of the different charts and eliminates the errors that may result from reading such charts. There is a minimal data requirement for the calculation using this programme, with the possibility of calculating the surface tension and other required data being built in the programme. Furthermore, using the programme makes comparison of the different calculation methods an easy and straightforward matter. The order of the calculated diameters using the programme is as follows:

$D_1$  based on Fig. 3;

$D_2$  based on Fig. 4;

$D_3$  based on Fig. 5;

$D_4$  based on Fig. 6;

$D_5$  based on Eq. 2.

The use of this programme will be illustrated by two examples in which the diameters of two fractionation columns used at the Homs oil refinery, in Syria, are calculated.

### Example 1

Fractionation column data

$T$  = Plate spacing = 22 in. = 56 cm

$Q_V$  = Vapour volume flow rate =  $29 \times 10^2 \text{ m}^3/\text{h} = 10 \times 10^4 \text{ ft}^3/\text{h}$

$L$  = Liquid flow rate =  $65.3 \times 10^2 \text{ kg/h} = 14.4 \times 10^3 \text{ lb/h}$

$V$  = Maximum allowable vapour flow rate =  $83.3 \times 10^2 \text{ kg/h} = 18.3 \times 10^3 \text{ lb/h}$

$\rho_L$  = Liquid density =  $684 \text{ kg/m}^3 = 42.7 \text{ lb/ft}^3$

$\rho_V$  = Vapour density =  $2.9 \text{ kg/m}^3 = 0.18 \text{ lb/ft}^3$

$t$  = Temperature =  $141^\circ\text{C}$

$M$  = Average molecular weight = 110

The calculated tower diameter varied between 1.0 and 1.2 m depending on the method or chart used for calculation. The results are as follows:

$D_1 = 1.0 \text{ m}$

$D_2 = 1.2 \text{ m}$

$D_3 = 1.2 \text{ m}$

$D_4 = 1.2 \text{ m}$

$D_5 = 1.1 \text{ m}$

The diameter of the actual tower is 1.2 m which happens to be also the diameter calculated using the charts of Figures 4, 5 and 6. As was earlier indicated, the surface tension of the system is not taken into account in the estimation of the  $K$  factor using Fig. 3 or Eq. 2, and using the values of the  $K$  factor from Eq. 2 or from the chart of Fig. 3 often gives lower estimates of the column diameter, with Eq. 2 giving a better estimate of the tower diameter (1.1 m) than Fig. 3 (1.0 m). Neglecting the diameter calculated using the  $K$  value of Fig. 3 gives an average calculated diameter of  $1.2 \pm 0.09 \text{ m}$  with a maximum difference between the calculated values of less than 9% of the average, which is quite acceptable and falls well within the accuracy limits of the methods of calculation.

### Example 2

Fractionation column data

$T$  = plate spacing = 18 in. = 46 cm

$Q_V$  = Vapour volume flow rate =  $96 \times 10^2 \text{ m}^3/\text{h} = 33.1 \times 10^4 \text{ ft}^3/\text{h}$

$L$  = Liquid flow rate =  $25.6 \times 10^3 \text{ kg/h} = 56.5 \times 10^3 \text{ lb/h}$

$V$  = Maximum allowable vapour flow rate =  $11.5 \times 10^3 \text{ kg/h} = 25.4 \times 10^3 \text{ lb/h}$

$\rho_L$  = Liquid density =  $7.2 \times 10^2 \text{ kg/m}^3 = 45 \text{ lb/ft}^3$

$\rho_V$  = Vapour density =  $1.2 \text{ kg/m}^3 = 0.075 \text{ lb/ft}^3$

$t$  = Temperature =  $149^\circ\text{C}$

$M$  = Average molecular weight = 98

The calculated diameter varied between 1.6 and 2.0 m, and the results are as follows:

$D_1 = 1.6 \text{ m}$

$D_2 = 2.0 \text{ m}$

$D_3 = 2.0 \text{ m}$

$D_4 = 2.0 \text{ m}$

$D_5 = 1.8 \text{ m}$

The actual column diameter in this case is 2.0 m, which is also the diameter calculated using the charts of Figures 4, 5 and 6. As mentioned before, using Eq. 2 or the chart of Fig. 3 gives lower estimates of the tower diameter (1.8 and 1.6 m respectively). Neglecting the diameter calculated using the  $K$  value of Fig. 3 gives an average diameter of  $2.0 \pm 0.18 \text{ m}$  with a maximum difference of 10%.

## 8. Conclusion

The use of the Matlab programme for the calculation of fractionation column diameters greatly simplified the calculation procedure and made it possible to compare calculated diameters based on different calculation procedures and/or assumptions.

The results obtained on applying the programme on the two fractionation columns at the Homs Oil Refinery indicate clearly the significance of surface tension in the calculation of tower diameters. In all three procedures in which the surface tension is taken into account, accurate and equal estimates of the tower diameter were obtained. When using, however, Eq. 2 or the chart of Fig. 3, where the surface tension is not considered, lower estimates were generally obtained, although in both instances Eq. 2 gave a much better estimate than the chart of Fig. 3. The calculation results obtained were found to be comparable in general and mostly fall well within the accuracy limits of the methods of calculation.

As indicated above, the calculated column diameter will have to be further refined, by checking against the performance correlations of the column. Following this refinement, the eventual choice of the column diameter will be made.

## Nomenclature

$A$  = Cross-sectional area of the column ( $\text{m}^2$  or  $\text{ft}^2$ )

$A_A$  = Active area ( $\text{m}^2$  or  $\text{ft}^2$ )

$A_m$  = Cross-sectional area of the column on the moon

$A_N$  = Net area for vapour flow ( $\text{m}^2$  or  $\text{ft}^2$ )

$A_s$  = Slot area (or hole area) ( $\text{m}^2$  or  $\text{ft}^2$ )

$C_A$  = Hole-to-active area ratio correction factor (1, for  $A_s/A_A \geq 0.1$ ;  $5(A_s/A_A)+0.5$ , for  $0.06 \leq (A_s/A_A) < 0.1$ )

$C_F$  = Foaming correction factor (0.85-0.95, for non-foaming systems; 0.50-0.60 for foaming systems)

$C_s$  = Surface tension correction factor =  $\left(\frac{\sigma}{20}\right)^{0.2}$

$D$  = Column diameter (m or ft)

$d_H$  = Hole diameter (in).

$e_w$  = Weight of liquid entrained/unit weight of vapour.

$g$  = Gravity constant on earth

$g_m$  = Gravity constant on the moon.

$h_c$  = Height of clear liquid in bubbling zone (in).

$h_{ow}$  = Height of liquid over weir.

$h_w$  = Height of outlet weir.

$K$  = A factor that depends on plate spacing.

$L$  = Liquid flow rate (kg/h or lb/h)

$M$  = Average molecular weight

$Q_V$  = Vapour volume flow rate ( $\text{m}^3/\text{h}$  or  $\text{ft}^3/\text{h}$ ).

$T$  = plate spacing (in., cm or m)

$T'$  = Effective tray spacing (in.)

$t$  = Temperature ( $^{\circ}\text{C}$ )

$U_F$  = Flooding vapour velocity (m/sec or ft/sec).

$U_V$  = Maximum allowable superficial vapour velocity (m/sec or ft/sec).

$U_{VN}$  = Vapour velocity relative to the net area for vapour flow (m/sec or ft/sec).

$V$  = Maximum allowable vapour flow rate (kg/h or lb/h)

$\rho_L$  = Liquid density ( $\text{kg}/\text{m}^3$  or  $\text{lb}/\text{ft}^3$ )

$\rho_V$  = Vapour density ( $\text{kg}/\text{m}^3$  or  $\text{lb}/\text{ft}^3$ )

$\sigma$  = Surface tension (dynes/cm)

## Appendix

```
disp('Matlab programme for the calculation of fractionation column diameter')
```

```
N = input('N = ')
```

```
switch N
```

```
Case 1
```

```
disp('Fig -3')
```

```
t = input('T = ')
```

```
x = t;
```

```
y = ((0.0254 * x ^ 3 - 2.793 * x ^ 2 + 110.6 * x - 616.6) + 0.5);
```

```
k = y
```

```
disp(['K=' int2str(k)])
```

```
EL = input('EL');
```

```
EV = input('EV');
```

```
V = input('V');
```

```
Uv = k / 3600 * ((EL - EV) / EV) ^ 0.5;
```

```
W = 3600 * EV * Uv;
```

```
D = (4 * V / (3.14 * W)) ^ 0.5;
```

```
disp(['D=' num2str(D * 0.3048)])
```

```
Case 2
```

```
disp(' Fig - 4')
```

```
t = input('t= ');
```

```
M = input('M = ');
```

```
if t == -20 Then
```

```
x = M;
```

```
y = 0.000006446 * x ^ 3 - 0.0033908 * x ^ 2 + 0.61513 * x - 9.3259;
```

```
s = y;
```

```
Else if t == 40
```

```
x = M;
```

```
y = -0.000000018 * x ^ 4 + 0.000014 * x ^ 3 - 0.0044 * x ^ 2 + 0.675 * x - 17;
```

```
s = y;
```

```
Else if t > -20 & t < 40
```

```
x = M;
```

```
y = 0.000006446 * x ^ 3 - 0.0033908 * x ^ 2 + 0.61513 * x - 9.3259;
```

```
S1 = y;
```

```
x = M;
```

```
y = -0.000000018 * x ^ 4 + 0.000014 * x ^ 3 - 0.0044 * x ^ 2 + 0.675 * x - 17;
```

```
S2 = y;
```

```
s = S1 - (S1 - S2) / (-20 - 40) * (-20 - t);
```

```
Else if t == 90
```

```
x = M;
```

```
y = -0.00000001 * x ^ 4 + 0.00001 * x ^ 3 - 0.0036 * x ^ 2 + 0.601 * x - 20;
```



```

s = y;
Else if (t > 40) & (t < 90)
x = M;
y = -0.000000018 * x ^ 4 + 0.000014 * x ^ 3 - 0.0044 * x ^ 2 + 0.675 * x - 17;
S1 = y;
x = M;
y = -0.00000001 * x ^ 4 + 0.00001 * x ^ 3 - 0.0036 * x ^ 2 + 0.601 * x - 20;
S2 = y;
s = S1 - (S1 - S2) / (90 - 40) * (90 - t);
Else if t == 150
x = M;
y = 0.00000235 * x ^ 3 - 0.00154 * x ^ 2 + 0.387 * x - 17.71;
s = y;
Else if (t > 90) & (t < 150)
x = M;
y = -0.00000001 * x ^ 4 + 0.00001 * x ^ 3 - 0.0036 * x ^ 2 + 0.601 * x - 20;
S1 = y;
x = M;
y = 0.00000235 * x ^ 3 - 0.00154 * x ^ 2 + 0.387 * x - 17.71;
S2 = y;
s = S1 - (S1 - S2) / (150 - 90) * (150 - t);
end
disp(['s=' num2str(s)])
t = input('T = ');
if s == 20
x = t;
y = -0.049 * x ^ 2 + 7.941 * x - 97.82;
k = y;
Else if s == 10
x = t;
y = -0.045 * x ^ 2 + 7.359 * x - 107.2;
k = y;
Else if s > 10 & s < 20
x = t;
y = -0.049 * x ^ 2 + 7.941 * x - 97.82;
k1 = y;
x = t;
y = -0.045 * x ^ 2 + 7.359 * x - 107.2;
K2 = y;
k = k1 - (k1 - K2) / (20 - 10) * (20 - s);
Else if s == 5
x = t;
y = -0.034 * x ^ 2 + 5.834 * x - 81.98;

```

```

k = y;
Else if (s > 5) & (s < 10)
x = t;
y = -0.045 * x ^ 2 + 7.359 * x - 107.2;
k1 = y;
x = t;
y = -0.034 * x ^ 2 + 5.834 * x - 81.98;
K2 = y;
k = k1 - (k1 - K2) / (10 - 5) * (10 - s);
Else if s == 2.5
x = t;
y = -0.032 * x ^ 2 + 5.4 * x - 83.02;
k = y;
Else if (s > 2.5) & (s < 5)
x = t;
y = -0.034 * x ^ 2 + 5.834 * x - 81.98;
k1 = y;
x = t;
y = -0.032 * x ^ 2 + 5.4 * x - 83.02;
K2 = y;
k = k1 - (k1 - K2) / (5 - 2.5) * (5 - s);
Else if t == 1
x = t;
y = -0.029 * x ^ 2 + 4.918 * x - 77.79;
k = y;
Else if s > 1 & s < 2.5
x = t;
y = -0.032 * x ^ 2 + 5.4 * x - 83.02;
k1 = y;
x = t;
y = -0.029 * x ^ 2 + 4.918 * x - 77.79;
K2 = y;
k = k1 - (k1 - K2) / (2.5 - 1) * (2.5 - s);
end
disp(['K=' int2str(k)])
EL = input('EL');
EV = input('EV');
V = input('V');
Uv = k / 3600 * ((EL - EV) / EV) ^ 0.5;
W = 3600 * EV * Uv;
D = (4 * V / (3.14 * W)) ^ 0.5;
disp(['D=' num2str(D)])
Case 3

```

```

disp(' Fig - 5')
t = input('t= ');
M = input('M = ');
if t == -20
    x = M;
    y = 0.000006446 * x ^ 3 - 0.0033908 * x ^ 2 + 0.61513 * x - 9.3259;
    s = y;
Else if t == 40
    x = M;
    y = -0.000000018 * x ^ 4 + 0.000014 * x ^ 3 - 0.0044 * x ^ 2 + 0.675 * x - 17;
    s = y;
Else if t > -20 & t < 40
    x = M;
    y = 0.000006446 * x ^ 3 - 0.0033908 * x ^ 2 + 0.61513 * x - 9.3259;
    S1 = y;
    x = M;
    y = -0.000000018 * x ^ 4 + 0.000014 * x ^ 3 - 0.0044 * x ^ 2 + 0.675 * x - 17;
    S2 = y;
    s = S1 - (S1 - S2) / (-20 - 40) * (-20 - t);
Else if t == 90
    x = M;
    y = -0.00000001 * x ^ 4 + 0.00001 * x ^ 3 - 0.0036 * x ^ 2 + 0.601 * x - 20;
    s = y;
Else if (t > 40) & (t < 90)
    x = M;
    y = -0.000000018 * x ^ 4 + 0.000014 * x ^ 3 - 0.0044 * x ^ 2 + 0.675 * x - 17;
    S1 = y;
    x = M;
    y = -0.00000001 * x ^ 4 + 0.00001 * x ^ 3 - 0.0036 * x ^ 2 + 0.601 * x - 20;
    S2 = y;
    s = S1 - (S1 - S2) / (90 - 40) * (90 - t);
Else if t == 150
    x = M;
    y = 0.00000235 * x ^ 3 - 0.00154 * x ^ 2 + 0.387 * x - 17.71;
    s = y;
Else if (t > 90) & (t < 150)
    x = M;
    y = -0.00000001 * x ^ 4 + 0.00001 * x ^ 3 - 0.0036 * x ^ 2 + 0.601 * x - 20;
    S1 = y;
    x = M;
    y = 0.00000235 * x ^ 3 - 0.00154 * x ^ 2 + 0.387 * x - 17.71;
    S2 = y;
    s = S1 - (S1 - S2) / (150 - 90) * (150 - t);

```

```

end
disp(['s=', num2str(s)])
t = input('T = ')
if t == 36
    x = s;
    y = 123.2 * Log(x) + 363.9;
    k = y;
Else if t == 30
    x = s;
    y = 120.7 * Log(x) + 337.3;
    k = y;
Else if t < 36 And t > 30
    x = s;
    y = 123.2 * Log(x) + 363.9;
    k1 = y;
    x = s;
    y = 120.7 * Log(x) + 337.3;
    K2 = y;
    k = k1 - (k1 - K2) / (36 - 30) * (36 - t);
Else if t == 24
    x = s;
    y = 118.6 * Log(x) + 287.4;
    k = y;
Else if t < 30 And t > 24
    x = s;
    y = 120.7 * Log(x) + 337.3;
    k1 = y;
    x = s;
    y = 118.6 * Log(x) + 287.4;
    K2 = y;
    k = k1 - (k1 - K2) / (30 - 24) * (30 - t);
Else if t == 20
    x = s;
    y = 113.2 * Log(x) + 231.1;
    k = y;
Else if t < 24 And t > 20
    x = s;
    y = 118.6 * Log(x) + 287.4;
    k1 = y;
    x = s;
    y = 113.2 * Log(x) + 231.1;
    K2 = y;
    k = k1 - (k1 - K2) / (24 - 20) * (24 - t);

```

```

Else if t == 18
x = s;
y = 107 * Log(x) + 197.9;
k = y;
Else if t < 20 And t > 18
x = s;
y = 113.2 * Log(x) + 231.1;
k1 = y;
x = s;
y = 107 * Log(x) + 197.9;
K2 = y;
k = k1 - (k1 - K2) / (20 - 18) * (20 - t);
Else if t == 15
x = s;
y = 92.91 * Log(x) + 136.55;
k = y;
Else if t < 18 And t > 15
x = s;
y = 107 * Log(x) + 197.9;
k1 = y;
x = s;
y = 92.91 * Log(x) + 136.55;
K2 = y;
k = k1 - (k1 - K2) / (18 - 15) * (18 - t);
Else if t == 12
x = s;
y = 74.62 * Log(x) + 53.18;
k = y;
Else if t < 15 And t > 12
x = s;
y = 92.91 * Log(x) + 136.55;
k1 = y;
x = s;
y = 74.62 * Log(x) + 53.18;
K2 = y;
k = k1 - (k1 - K2) / (15 - 12) * (15 - t)
Else if t == 10
x = s;
y = 46.68 * Log(x) + 15.74;
k = y;
Else if t < 12 And t > 10
x = s;
y = 74.62 * Log(x) + 53.18;

```



```

k1 = y;
x = s;
y = 46.68 * Log(x) + 15.74;
K2 = y;
k = k1 - (k1 - K2) / (12 - 10) * (12 - t);
end
disp(['k=' int2str(k)])
EL = input('EL');
EV = input('EV');
V = input('V');
Uv = k / 3600 * ((EL - EV) / EV) ^ 0.5;
W = 3600 * EV * Uv;
D = (4 * V / (3.14 * W)) ^ 0.5;
disp(['D=' num2str(D * 0.3048)])

Case 4
disp('Fig - 6')
t = input('t= ');
M = input('M = ');
if t == -20 Then
x = M;
y = 0.000006446 * x ^ 3 - 0.0033908 * x ^ 2 + 0.61513 * x - 9.3259;
s = y;
Else if t == 40
x = M;
y = -0.000000018 * x ^ 4 + 0.000014 * x ^ 3 - 0.0044 * x ^ 2 + 0.675 * x - 17;
s = y;
Else if t > -20 & t < 40
x = M;
y = 0.000006446 * x ^ 3 - 0.0033908 * x ^ 2 + 0.61513 * x - 9.3259;
S1 = y;
x = M;
y = -0.000000018 * x ^ 4 + 0.000014 * x ^ 3 - 0.0044 * x ^ 2 + 0.675 * x - 17;
S2 = y;
s = S1 - (S1 - S2) / (-20 - 40) * (-20 - t);
Else if t == 90
x = M;
y = -0.00000001 * x ^ 4 + 0.00001 * x ^ 3 - 0.0036 * x ^ 2 + 0.601 * x - 20;
s = y;
Else if (t > 40) & (t < 90)
x = M;
y = -0.000000018 * x ^ 4 + 0.000014 * x ^ 3 - 0.0044 * x ^ 2 + 0.675 * x - 17;
S1 = y;
x = M;

```

```

y = -0.00000001 * x ^ 4 + 0.00001 * x ^ 3 - 0.0036 * x ^ 2 + 0.601 * x - 20;
S2 = y;
s = S1 - (S1 - S2) / (90 - 40) * (90 - t);
Else if t == 150
x = M;
y = 0.00000235 * x ^ 3 - 0.00154 * x ^ 2 + 0.387 * x - 17.71;
s = y;
Else if (t > 90) & (t < 150)
x = M;
y = -0.00000001 * x ^ 4 + 0.00001 * x ^ 3 - 0.0036 * x ^ 2 + 0.601 * x - 20;
S1 = y;
x = M;
y = 0.00000235 * x ^ 3 - 0.00154 * x ^ 2 + 0.387 * x - 17.71;
S2 = y;
s = S1 - (S1 - S2) / (150 - 90) * (150 - t);
end
disp(['s=' num2str( s)])
L = input('L = ');
V = input('V = ');
EV = input('EV = ');
DL = input('DL = ');
t = input('T = ');
Qv = input('Qv = ');
pF = (L / V) * (EV / dL) ^ 0.5;
if t == 36
x = pF;
y = 0.104 * x ^ 4 - 0.52 * x ^ 3 + 0.974 * x ^ 2 - 0.916 * x + 0.508;
pc = y;
Else if t == 24
x = pF;
y = 0.111 * x ^ 4 - 0.524 * x ^ 3 + 0.903 * x ^ 2 - 0.765 * x + 0.399;
pc = y;
Else if t > 24 And t < 36
x = pF;
y = 0.104 * x ^ 4 - 0.52 * x ^ 3 + 0.974 * x ^ 2 - 0.916 * x + 0.508;
PC1 = y;
x = pF;
y = 0.111 * x ^ 4 - 0.524 * x ^ 3 + 0.903 * x ^ 2 - 0.765 * x + 0.399;
PC2 = y;
pc = PC1 - (PC1 - PC2) / (36 - 24) * (36 - t);
Else if t == 18
x = pF;
y = 0.058 * x ^ 4 - 0.295 * x ^ 3 + 0.555 * x ^ 2 - 0.518 * x + 0.298;

```

```

pc = y;
Else if (t > 18) And (t < 24)
x = pF;
y = 0.111 * x ^ 4 - 0.524 * x ^ 3 + 0.903 * x ^ 2 - 0.765 * x + 0.399;
PC1 = y;
x = pF;
y = 0.058 * x ^ 4 - 0.295 * x ^ 3 + 0.555 * x ^ 2 - 0.518 * x + 0.298;
PC2 = y;
pc = PC1 - (PC1 - PC2) / (24 - 18) * (24 - t);
Else if t == 12
x = pF;
y = -0.053 * x ^ 5 + 0.292 * x ^ 4 - 0.618 * x ^ 3 + 0.663 * x ^ 2 - 0.438 * x + 0.234;
pc = y;
Else if (t > 12) And (t < 18)
x = pF;
y = 0.058 * x ^ 4 - 0.295 * x ^ 3 + 0.555 * x ^ 2 - 0.518 * x + 0.298;
PC1 = y;
x = pF;
y = -0.053 * x ^ 5 + 0.292 * x ^ 4 - 0.618 * x ^ 3 + 0.663 * x ^ 2 - 0.438 * x + 0.234;
PC2 = y;
pc = PC1 - (PC1 - PC2) / (18 - 12) * (18 - t);
Else if t == 9
x = pF;
y = 0.034 * x ^ 4 - 0.161 * x ^ 3 + 0.284 * x ^ 2 - 0.269 * x + 0.186;
pc = y;
Else if t > 9 And t < 12
x = pF;
y = -0.053 * x ^ 5 + 0.292 * x ^ 4 - 0.618 * x ^ 3 + 0.663 * x ^ 2 - 0.438 * x + 0.234;
PC1 = y;
x = pF;
y = 0.034 * x ^ 4 - 0.161 * x ^ 3 + 0.284 * x ^ 2 - 0.269 * x + 0.186;
PC2 = y;
pc = PC1 - (PC1 - PC2) / (12 - 9) * (12 - t);
Else if t == 6
x = pF;
y = -0.013 * x ^ 4 + 0.042 * x ^ 3 - 0.002 * x ^ 2 - 0.11 * x + 0.147;
pc = y;
Else if t > 6 And t < 9
x = pF;
y = 0.034 * x ^ 4 - 0.161 * x ^ 3 + 0.284 * x ^ 2 - 0.269 * x + 0.186;
PC1 = y;
x = pF;
y = -0.013 * x ^ 4 + 0.042 * x ^ 3 - 0.002 * x ^ 2 - 0.11 * x + 0.147;

```

```

PC2 = y;
pc = PC1 - (PC1 - PC2) / (9 - 6) * (9 - t);
end
disp( pc)
disp('pF=')
disp (pF)
disp( 'pc=')
disp (pc) )
corpc = pc * (s / 20) ^ 0.2;
Uvn = corpc / (EV / (DL - EV)) ^ 0.5;
D = Sqrt((4 * Qv / (0.54 * 3.14 * Uvn))/60;
disp(['D=' num2str(D * 0.3048)])

Case 5
disp( 'Eq')
T = input ('T');
EL = input ('EL');
EV = input ('EV');
V = input ('V');
k = 3600 * (-0.171 * T ^ 2 + 0.27 * T - 0.047);
Uv = k / 3600 * ((EL - EV) / EV) ^ 0.5;
W = 3600 * EV * Uv;
D = ((4 * V) / (3.14 * W))^0.5;
disp( ['k=' num2str(k)])
disp( ['D=' num2str(D)])

End

```

## Author details

Hassan Al-Haj Ibrahim\*

Address all correspondence to: [sanjim84@yahoo.com](mailto:sanjim84@yahoo.com)

Dept of Chemical Engineering, Al-Baath University, Homs, Syria

## References

- [1] Forbes RJ. A short history of the art of distillation. Leiden: E. J. Brill; 1970.
- [2] Sorel E. La rectification de l'alcool. Paris: Gauthier-Villars et fils; 1893.

- [3] (Lord) Rayleigh. On the distillation of binary mixtures. *Philosophical Magazine* 1902; 4:521-537.
- [4] Urbain G and Boll M. *La science, ses progrès, ses applications*. Vol. 2, *Les applications et les théories actuelles*, Paris: Librairie Larousse; p. 304, 1934.
- [5] Humphrey JL. and Keller, G. E., *Separation Process Technology*, New York: McGraw-Hill; 1997.
- [6] Al-Haj Ibrahim H and Daghistani N. *Petroleum refinery engineering (In Arabic)*. vol. 3, Homs, Syria: Al-Baath University; 2008.
- [7] *Distillation Column Selection and Sizing*. KLM Technology group, 2011, available at: [www.klmtechgroup.com](http://www.klmtechgroup.com).
- [8] Chen H and Lin Y. Case Studies on Optimum Reflux Ratio of Distillation Towers in Petroleum Refining Processes. *Tamkang Journal of Science and Engineering* 2001; 4(2):105-110.
- [9] Kister HZ. *Distillation design*. New York: McGraw-Hill Book Co.; 1992.
- [10] Kolmetz K et al. Optimize Distillation Column Design for Improved Reliability in Operation and Maintenance. 2nd Best Practices in Process Plant Management, Kuala Lumpur, Malaysia, March 14-15, 2005.
- [11] Pettit DR. Fractional distillation in a lunar environment. Los Alamos National Laboratory, MS P952, Los Alamos, NM87545.
- [12] Holland CD. *Fundamentals of multi-component distillation*. New York: McGraw-Hill; 1997.
- [13] Tham MT. *Distillation, an introduction*. available at: <http://lorien.ncl.ac.uk/ming/distil/distil0.htm>.
- [14] Kister HZ et al. How Do Trays and Packing Stack Up? *Chemical Engineering Progress*, February 1994.
- [15] Kister HZ et al. Equipment for Distillation. Gas Absorption, Phase Dispersion and Phase Separation, Section 14, *Perry's Chemical Engineers Handbook*, 8<sup>th</sup> Edition, New York: McGraw-Hill; 2008.
- [16] Razia SS. Equipment Design and Costs for Separating Homogeneous Mixtures. Bangladesh University of Engineering and Technology, Dhaka.
- [17] Seader JD, Siirola JJ, Barnicki SD. Distillation. Ch. 13 of *Perry's Chemical Engineers Handbook*, 7<sup>th</sup> Edition, New York: McGraw-Hill; 1997.
- [18] Youssef S et al. Calcul d'une colonne de rectification à multiples alimentations et soutirages latéraux. *The Chemical Engineering Journal* 1989, 42(3):153–165.



- [19] Jevric J, and Fayed ME. Shortcut Distillation Calculations via Spreadsheets. Dec. 2002, available at: [www.cepmagazine.org](http://www.cepmagazine.org).
- [20] Guthrie KM. Data and techniques for preliminary capital cost estimating. Chem. Eng. 142, Mar. 24, 1969.
- [21] Ludwig EE. Applied process design for chemical and petrochemical plants. vol. 2, Gulf Professional Publishing; 1997.
- [22] Sinnott RK. Coulson & Richardson's Chemical Engineering. Elsevier, vol. 6, 2006.
- [23] Douglas J. Conceptual Design of Chemical Processes. McGraw-Hill, Section A.3, 1988.
- [24] Nelson WL. Petroleum refinery engineering. McGraw-Hill Book Co., New York, 1958.
- [25] Wuithier P. editor. Le pétrole: Raffinage et génie chimique. Paris: L'institut français du pétrole; Vol. 2, 1972.
- [26] Van Winkle M. Distillation. New York: McGraw-Hill Book Co.; 1967.
- [27] Smith RB et al. Tower capacity rating ignores trays. Hydrocarbon process. and pet. Refiner 1963; 40(5):183.
- [28] Gerster JA. Recent advances in distillation. Davis-Swindin memorial lecture, University of Loughborough, 1964.
- [29] Branan CR. The Process Engineer's Pocket Handbook. Vol. 1, Gulf Publishing Co., 1976.

