Energy Considerations in Distillation

Megan Jobson
School of Chemical Engineering and Analytical Science, The University of Manchester, Manchester, UK

CHAPTER OUTLINE

6.1 Introduction to energy efficiency ............................................................... 226
   6.1.1 Energy efficiency: technical issues ..................................................... 227
      6.1.1.1 Heating .................................................................................... 227
      6.1.1.2 Cooling—above ambient temperatures ........................................ 229
      6.1.1.3 Cooling—below ambient temperatures ....................................... 230
      6.1.1.4 Mechanical or electrical power .................................................. 232
      6.1.1.5 Summary—technical aspects of energy efficiency .................... 233
   6.1.2 Energy efficiency: process economics ................................................. 233
      6.1.2.1 Heating .................................................................................... 233
      6.1.2.2 Cooling .................................................................................... 235
      6.1.2.3 Summary—process economics and energy efficiency ............ 237
   6.1.3 Energy efficiency: sustainable industrial development ...................... 237

6.2 Energy-efficient distillation ....................................................................... 237
   6.2.1 Energy-efficient distillation: conceptual design of simple columns .... 238
      6.2.1.1 Degrees of freedom in design .................................................. 238
      6.2.1.2 Column operating pressure ...................................................... 240
      6.2.1.3 Pressure drop .......................................................................... 242
      6.2.1.4 Number of theoretical stages .................................................... 242
      6.2.1.5 Feed condition .......................................................................... 243
      6.2.1.6 Feed stage location ................................................................. 244
      6.2.1.7 Condenser type ........................................................................ 245
      6.2.1.8 Summary—column design for energy efficiency .................... 245

6.3 Energy-efficient distillation: operation and control .................................. 246
   6.3.1 Energy-efficient column operation ..................................................... 246
   6.3.2 Process control .................................................................................. 246
   6.3.3 Summary—operation and control ....................................................... 247

6.4 Heat integration of distillation .................................................................. 247
   6.4.1 Heat exchange and heat recovery ....................................................... 248
      6.4.1.1 Summary—heat exchange and heat integration ...................... 249
6.1 Introduction to energy efficiency

Distillation processes involve mass transfer between a liquid phase (or two liquid phases) and a vapor phase flowing in countercurrent fashion. The vapor and liquid phases are generated by vaporization of a liquid stream and condensing a vapor stream, which in turn requires heating and cooling. Distillation is thus a major user of energy in the process industries and globally. The sudden rise in crude oil prices in the 1970s, the general price increases since the start of the twenty-first century [1] and growing concerns about the environmental impact of using energy, in particular increasing emissions of carbon dioxide [2], drive efforts to increase the energy efficiency of this inherently energy-intensive process.

The underpinning physical and thermodynamic phenomena of distillation result in its energy intensity. Distillation design and operation need to be based on a clear understanding of these phenomena, “energy efficiency” and the process economics of distillation processes and the associated energy-supply processes.

This chapter first introduces the concept of energy efficiency in the context of distillation in the process industries. The influence of design and operation of
individual, conventional, simple, continuous distillation columns on energy efficiency is explored and the role of heat recovery for enhancing energy efficiency is discussed. Advanced and complex distillation configurations that exploit opportunities to reduce the energy requirements or increase heat recovery are presented. A short concluding section provides a summary and notes areas of ongoing research.

### 6.1.1 Energy efficiency: technical issues

It is significant that energy may take various forms, including heat (or cooling) and work, in the form of mechanical or electrical power. This section provides an overview of energy supply technologies applied in the process industries; more detail may be found in various sources (e.g. [3,4]). The process economics of energy efficiency is discussed in Section 6.1.2.

#### 6.1.1.1 Heating

In the process industries, process heating requirements may be met efficiently and relatively safely using steam, although electrical heating is also applied, especially at smaller scales where steam generation is impractical. Heat released by the steam on condensing (dominated by the latent heat of condensation of the steam, rather than its sensible heat), is transferred to the process stream being heated. On condensing, the steam needs to be hotter than the stream being heated, by the second law of thermodynamics. There is an upper limit to the temperature at which the steam can condense, since the critical temperature of steam is 374 °C. In practice, steam is used in the process industries at pressures of up to about 4000 kPa, corresponding to a condensing temperature of 250 °C [5].

Heat is typically generated on a process site at various temperatures in a site utility system, as illustrated in Figure 6.1, where fuels are combusted to generate steam at a very high pressure (e.g. 11,000 kPa) with a relatively low heat of vaporization. This very high pressure (VHP) steam then has its pressure reduced to two or more lower pressure levels. While pressure-reducing valves may be used to “let down” the steam pressure, steam turbines facilitate generation of mechanical and/or electrical power as a by-product. On smaller sites, steam may be generated at a high pressure (around 4000 kPa) rather than a very high pressure [3].

The steam is then distributed across the process site at various pressures (e.g. as high-pressure (HP) steam, medium-pressure (MP) steam at around 2000 kPa and low-pressure (LP) steam at around 400 kPa), with corresponding condensing temperatures and additional power generation [3]. While the condensing temperature decreases as the pressure decreases, its latent heat of vaporization increases, as does power generation. It is therefore beneficial to use steam for heating at the lowest pressure available that allows effective heat transfer—a minimum temperature difference between the process stream and condensed steam of 20—40 °C is typical.

At higher temperatures, it is common practice to use fired heaters, as shown in Figure 6.2, where a fuel (gas, liquid, or solid) is combusted and the flue gases of the
FIGURE 6.1 Site Utility System

Reprinted from Klemes et al., Targeting and design methodology for reduction of fuel, power and CO$_2$ on total sites, Applied Thermal Engineering 17 (8–10) 993–1003, Copyright 1997, with permission from Elsevier.

FIGURE 6.2 Fired Heater (or Furnace)

combustion process provide heat to the process by radiant and convective heat transfer. Typically, modern furnaces transfer 80–90% of the heat of combustion of the fuel to the process stream \[^5,6\], with temperatures within the radiant section of the furnace of 700–900 °C \[^3\]. The flue gas, which contains mainly nitrogen, carbon dioxide and water, undergoes sensible cooling to its dew point, typically 150–160 °C, where corrosive liquids form that can damage the chimney stack. In the case of natural gas, temperatures below 100 °C can be achieved, allowing latent heat of the water to be recovered \[^3\]. The temperature to which the flue gas is cooled represents a limitation to the amount of heat that can be recovered from the stack gas.

Heat transfer fluids are also used to provide heat at higher temperatures, e.g. 350 °C; these fluids in turn require heating by a high-temperature heat source such as a fired heater, as shown in Figure 6.3. Using heat transfer fluids, pumped to a location some distance away from the fired heater, can bring safety or layout benefits. However, the need for intermediate heat exchange (i.e. flue gas—hot oil, hot oil—process stream) increases equipment costs and decreases the temperature to which the process stream may be heated.

At temperatures below the condensation temperature of low-pressure steam (130–150 °C), heat may be provided to a process by other utilities, such as hot water.

### 6.1.1.2 Cooling—above ambient temperatures

When cooling is required in a process plant above the ambient temperature, it is usually convenient to use ambient air or cooling water as the cold utility. The process stream may be cooled using an air-cooled heat exchanger, such as that shown in Figure 6.4, to a temperature about 20 °C hotter than the ambient air temperature; to achieve a smaller temperature difference than this would require significantly more heat transfer area. Fans or blowers increase the effectiveness of the coolers but consume power.

When available in an industrial site, cooling water is the most common cooling medium. The cooling water system, illustrated in Figure 6.5, uses cooling towers, in which evaporation of the water takes place, causing the temperature of the

---

**FIGURE 6.3 Hot Oil Circuit [3]**

*Reprinted from Chemical Process Design and Integration, Copyright 2005, with permission from John Wiley and Sons.*
remaining water to fall. The temperature of the cooling water is in practice higher than the wet-bulb temperature of the ambient air. The process stream can be cooled to around 10°C warmer than the cold cooling water. The overall heat transfer coefficient of water cooling in a shell and tube heat exchanger is typically 500–900 W m⁻² K⁻¹, which is significantly higher than that of air cooling (typically around 100 W m⁻² K⁻¹) [5]; therefore, a lower temperature difference is achievable using cooling water, without requiring an excessively large heat exchanger.

6.1.1.3 Cooling—below ambient temperatures

Refrigeration is needed to cool a process stream to temperatures below ambient. In refrigeration systems, a refrigerant fluid acts as a heat sink (i.e. accepts heat) at a low temperature and rejects the heat at a higher temperature. Moving the heat from a lower to a higher temperature requires work to be done on the refrigerant.

The most common way of doing work on the refrigerant is to compress it when it is in the vapor or gas state, as illustrated in Figure 6.6. The refrigerant, a vapor–liquid mixture at a low pressure (point 1 on Figure 6.6), acts as a low-temperature heat sink. The temperature of the refrigerant at this point is lower than that of the
process stream requiring cooling; the temperature difference is usually in the range 1–5 °C. On absorbing heat from the process requiring cooling, the fluid is evaporated (point 2). This vapor is at its saturation temperature at a low pressure; when compressed, its temperature rises (point 3). The high-pressure refrigerant can be condensed using a relatively warm coolant, e.g. ambient air or cooling water, forming a liquid (point 4). The pressure of the refrigerant is then reduced through an expansion valve or an expander, generating a cold, low-pressure, two-phase mixture (point 1) that can take up heat at a low temperature.

The main drawback of compression refrigeration is that raising the pressure of the fluid requires a compressor, which is a relatively expensive equipment item and mechanical or electrical power to drive the compressor. Low-molecular weight hydrocarbons, hydrofluorocarbons (HFCs), ammonia and nitrogen are commonly used as the refrigerant; since 1987, chlorofluorocarbons (CFCs) and hydrochlorofluorocarbons (HCFCs) are being phased out [7,8].

Single cycles, such as that shown in Figure 6.6, can cool process streams to around −40 °C. In so-called “cascaded refrigeration cycles”, temperatures as low as −165 °C can be reached in a refrigeration cycle that rejects heat to one or more other refrigeration cycles.

An alternative to using compression to raise the pressure of the refrigerant fluid is to use an absorption—desorption process, as illustrated in Figure 6.7. As before, a two-phase refrigerant at a low pressure is evaporated by absorbing heat from the process stream requiring cooling. The vapor-phase refrigerant is then absorbed into a liquid absorbent, the pressure of which may be raised simply and inexpensively using a pump. The high-pressure mixture of the refrigerant and absorbent is separated in a desorption unit operating at a suitably high pressure. Desorption requires heat for vapor generation; cooling may also be required. The absorbent is then condensed, at the high pressure; once its pressure has been reduced, using an expansion valve, it can again act as a low-temperature heat sink.

While an absorption refrigeration cycle avoids using vapor compression, it requires heat as well as the absorption-desorption equipment. Absorption refrigeration is particularly attractive when low-grade heat (above 95 °C) is available and
inexpensive [3]. The solutions applied in these schemes are usually water based (common refrigerant—absorbent pairs are ammonia—water and water—lithium bromide) [9]; the lowest temperature reached by the refrigerant in these cycles is limited to around −40°C [3,8].

6.1.1.4 Mechanical or electrical power

It is important to consider energy in the form of work because of the need for refrigerated cooling in distillation processes as well as compression of distillation feeds and products. The steam system may provide power in the form of mechanical work generated by steam turbines, as illustrated in Figure 6.8, or generated using gas turbines. Alternatively, electricity may be imported from off-site generators.
6.1.1.5 Summary—technical aspects of energy efficiency

Energy may be employed in various forms. When considering energy efficiency in distillation it is important to account for the form of the energy being consumed and the quality of the heating and cooling required. Trade-offs exist between the energy sources, in terms of equipment requirements, cost and environmental impact.

6.1.2 Energy efficiency: process economics

Energy required for distillation processes—heat, cooling and power—require combustion of fuels, in fired heaters, for heat and power generation, as well as water or air for cooling. Energy-efficient distillation can reduce the demand for these utilities, bringing a range of benefits, in terms of the associated greenhouse gas emissions and use of water for process cooling, the size and cost of various equipment items and process operating costs. The main benefit is the reduction of operating costs, i.e. costs of fuel, power, electricity and water. A site may export electricity, thereby generating revenue.

This section considers the operating costs of various forms of energy for distillation. Investment and other costs will not be discussed, except briefly in Section 6.6. It is widely recognized that operating costs—in particular, costs of utilities for heating and cooling—dominate distillation process economics. There are also important trade-offs between operating and capital costs.

6.1.2.1 Heating

A site steam system can provide heat at several temperatures. Given that the “path” or process by which the steam is generated depends on the pressure at which it is provided to the process, its associated cost does too. A useful way of evaluating the relative cost of different steam levels is to use “marginal costing” [3], where the value of any power generated offsets the cost of generating the steam using a given fuel. The unit cost of steam is case specific as it depends on the steam conditions and flows at each level, the efficiency of the steam turbines, the type and price of fuel, the cost of importing electricity rather than generating it, etc. Figure 6.9 provides an illustration, Table 6.1 summarizes the energy flows and Table 6.2 shows that the higher the temperature at which heat is required, the higher the cost, where conventional steam tables have been used to determine steam properties such as enthalpies and entropy. (A drawback of this method for costing steam is that in some cases the value of power generated can be so high that the cost of the steam at the lowest pressure becomes negative, which could perversely suggest that using more steam is more energy efficient [3].)

Estimating the operating cost of fired heating is relatively straightforward. The type of fuel and its cost and energy content (heat of combustion), as well as the efficiency of the furnace, determine the cost per unit of heat provided to the process. For a fired heater burning natural gas costing £6.3 GJ\(^{-1}\) with 80% efficiency, the operating costs would be £7.9 GJ\(^{-1}\) of high-temperature heat.
The operating cost of heat delivered by a heat transfer fluid is based on the cost of the heat from the furnace, taking into account heat losses of the heat distribution system. The operating cost of electrical heating, similarly, takes into account the cost or value of the electricity consumed and the efficiency of the heating system.

**FIGURE 6.9 Marginal Steam Costing: Illustrative Example**

*Assumptions:* Overall efficiency of boiler and steam distribution system: 80%; isentropic efficiency of steam turbines: 80%; cost of fuel (natural gas): £6.9 GJ⁻¹; value of power generated: £22.2 GJ⁻¹; steam condenses completely at constant pressure.

**Table 6.1 Steam Properties for Illustrative Example of Figure 6.9**

<table>
<thead>
<tr>
<th>Stream</th>
<th>Temperature (°C)</th>
<th>Condensate Temp. (°C)</th>
<th>Pressure (kPa)</th>
<th>Enthalpy (kJ kg⁻¹)</th>
<th>Entropy (kJ kg⁻¹ K⁻¹)</th>
<th>Enthalpy of Condensate (kJ kg⁻¹)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Boiler feed water</td>
<td>100</td>
<td>101</td>
<td>419</td>
<td>419</td>
<td>1.303</td>
<td>–</td>
</tr>
<tr>
<td>HP steam</td>
<td>400</td>
<td>253</td>
<td>4200</td>
<td>3210</td>
<td>6.742</td>
<td>1101</td>
</tr>
<tr>
<td>MP steam</td>
<td>250</td>
<td>184</td>
<td>1100</td>
<td>2939</td>
<td>6.876</td>
<td>781</td>
</tr>
<tr>
<td>LP steam</td>
<td>159</td>
<td>144</td>
<td>400</td>
<td>2772</td>
<td>6.975</td>
<td>605</td>
</tr>
</tbody>
</table>

The operating cost of heat delivered by a heat transfer fluid is based on the cost of the heat from the furnace, taking into account heat losses of the heat distribution system. The operating cost of electrical heating, similarly, takes into account the cost or value of the electricity consumed and the efficiency of the heating system.
6.1.2.2 Cooling

When air is used as the cooling medium, the dominant operating cost is that of the power consumed by the fans causing a draft through the air-cooled heat exchanger. Typically, the fan power demand is 0.5–1.5% of the cooling duty [10]. The power requirements for a given cooling duty may be estimated [5]. The lower rates of heat transfer to air than to water mean that higher minimum temperature approaches are needed, around 20 °C [10].

The operating cost of cooling using cooling water depends strongly on the operating context. Rules of thumb may be used: the operating cost per unit of cooling is typically 1% of the cost of power per unit of energy [3], or 3–5% of the unit cost of heating using low-pressure steam. In the illustrative example presented in Figure 6.9 and Table 6.2, the unit cost of cooling using water would be around £0.3 GJ⁻¹ cooling.

When subambient cooling is required, compression refrigeration is most commonly applied. Compressors are relatively expensive units and also use a power, a relatively expensive form of energy. Therefore, it is often preferable to avoid using refrigeration if another technical solution is feasible. Compressor costs can be estimated using standard approaches for estimating capital costs (e.g. Ref. [5,11]).

The unit cost of refrigeration depends strongly on both the temperature at which refrigerant is evaporated (point 2 in Figure 6.6) and the temperature of the liquid leaving the refrigerant condenser (point 4). A simple approach is to estimate the power demand of an ideal refrigeration cycle, i.e. a reverse Carnot engine and to apply a factor to consider the inefficiencies that apply to a real refrigeration cycle.

The “Carnot model” (Eqn (6.1)) relates the ideal compression power demand, \( W_{\text{ideal}} \), to raise the refrigerant from its evaporation temperature, \( T_{\text{evap}} \), to its condensing temperature, \( T_{\text{cond}} \), to satisfy the process cooling duty, \( Q_{\text{cool}} \):

\[
W_{\text{ideal}} = \frac{\dot{Q}_{\text{cool}}}{T_{\text{cond}} - T_{\text{evap}}} \tag{6.1}
\]

A factor representing the ratio of ideal to actual power demand of 0.6 is typically used for moderate evaporation temperatures, e.g. down to −40 °C, to estimate the actual compression work [3,7]; this factor accounts for mechanical and thermodynamic inefficiencies in the refrigeration cycle. The condensing and evaporation

<table>
<thead>
<tr>
<th>Steam Generated</th>
<th>Fuel Consumed (kJ kg⁻¹)</th>
<th>Net Power Generated (kJ kg⁻¹)</th>
<th>Heat Provided (kJ kg⁻¹)</th>
<th>Net Cost of Heat (£ GJ⁻¹)</th>
</tr>
</thead>
<tbody>
<tr>
<td>HP steam</td>
<td>3488</td>
<td>−</td>
<td>2108</td>
<td>11.5</td>
</tr>
<tr>
<td>MP steam</td>
<td>271</td>
<td>2158</td>
<td>8.4</td>
<td></td>
</tr>
<tr>
<td>LP steam</td>
<td>167</td>
<td>2167</td>
<td>6.7</td>
<td></td>
</tr>
</tbody>
</table>
temperature of the refrigerant are dictated by the temperature of the heat sink, which accepts heat from the refrigerant condenser and the temperature of the process stream requiring cooling, respectively. Subambient heat exchangers are designed to allow a minimum temperature approach of $1 - 5 \degree C$. A low temperature approach increases the heat transfer area required, but crucially increases the evaporation temperature and reduces the temperature difference $T_{\text{cond}} - T_{\text{evap}}$, which reduces the power demand of the refrigeration cycle. If a low-temperature heat sink is available to accept heat from the refrigerant condenser, this can reduce the compression work significantly. Table 6.3 illustrates unit operating costs for various operating conditions in a refrigeration cycle and rejecting heat to cooling water. Figure 6.10 summarizes illustrative utility costs at various temperature levels.

The compressor power demand and the condenser cooling duty may be estimated more accurately using a relatively rigorous model of the refrigeration cycle, taking into account the thermodynamic properties of the refrigerant fluid, the compressor type and efficiency and the pressure drop within the heat exchangers and pipework of the refrigeration cycle, etc. [3].

The cost of absorption refrigeration systems requires the absorption—desorption process to be designed and modeled to predict its energy requirements and to allow

<table>
<thead>
<tr>
<th>Process Stream Temperature (°C)</th>
<th>Ideal Work (kJ kJ⁻¹ Cooling)</th>
<th>Actual Work (kJ kJ⁻¹ Cooling)</th>
<th>Cost of Cooling (£ kJ⁻¹ Cooling)</th>
</tr>
</thead>
<tbody>
<tr>
<td>-40</td>
<td>0.35</td>
<td>0.59</td>
<td>13.0</td>
</tr>
<tr>
<td>-20</td>
<td>0.24</td>
<td>0.40</td>
<td>9.0</td>
</tr>
<tr>
<td>0</td>
<td>0.15</td>
<td>0.25</td>
<td>5.5</td>
</tr>
</tbody>
</table>

*Assumptions: Refrigerant evaporation temperature is $5 \degree C$ less than process cooling temperature; refrigerant is condensed at $35 \degree C$; ideal work is 0.6 of actual work; cost of electricity is £22.2 GJ⁻¹ (£0.08 kW⁻¹h⁻¹).

![Figure 6.10: Unit Costs of Heating and Cooling Utilities Depend on Their Temperatures](image_url)

**Figure 6.10** Unit Costs of Heating and Cooling Utilities Depend on Their Temperatures

Illustrative costs from Tables 6.2 and 6.3.
sizing of the associated equipment. The costs of this technology are not discussed further in this chapter; more details are provided by Ludwig [8].

6.1.2.3 Summary—process economics and energy efficiency
It is clear from Figure 6.10 that the form, quantity and quality (i.e. temperature) of energy required by distillation processes significantly affect distillation operating costs. Therefore, for energy-efficient distillation, the designer or operator should seek technical solutions taking into account the costs of utilities.

6.1.3 Energy efficiency: sustainable industrial development
Sustainable industrial development means that products that contribute to economic activity, satisfy human needs and support or enhance quality of life need to be produced with reduced consumption of materials and with reduced environmental damage [12]. Achieving this aim needs increased efficiency in the use of materials (including feedstocks, water and materials of construction) and of energy. The products being produced as a part of economic activity also need to add to quality of life. Their production should minimize the impact on the environment, in terms of consumption of energy, materials and water and of emissions of carbon dioxide and other damaging substances to the air, including substances that potentially deplete the ozone layer. A distinction must be made between energy derived from fossil fuels and other finite resources and that obtained from renewable sources.

Energy-efficient distillation processes play an important role in sustainable industrial development: decreasing demand for energy, whether in the form of heat, cooling, or work, can effectively reduce consumption of materials, energy and cooling water and decrease carbon dioxide and other atmospheric emissions. When energy efficiency reduces production costs, it also increases the economic sustainability of industrial activity.

A holistic consideration of the environmental impact of distillation processes should take into account the process life cycle [12], considering the extraction and consumption of fuels, extraction of materials of construction of the processing equipment and the final use of the products of the process. In a “cradle-to-grave” analysis, the use of resources and the emissions and wastes for the life cycle of a product consider the materials—their extraction, processing, transport, use (and reuse or recycling) and disposal—for the product and the equipment used to produce and process the product [12]. Note that the results of life cycle analyses depend significantly on where the boundaries are drawn for the analysis.

6.2 Energy-efficient distillation
Distillation columns are applied in the process industries in order to carry out separations for various reasons, including removal of contaminants that have a negative impact on downstream processes or are harmful or toxic and recovery of material
that would otherwise be wasted. Distillation is a versatile process that allows a wide range of flow rates, of mixtures, of mixture compositions, etc. to be separated into products of any degree of purity or with any degree of recovery (in the absence of azeotropes). The degree of separation is generally dictated by downstream requirements, e.g. the effect of impurities on a catalyst, or environmental regulation related to transport fuels, but the sharpness of the separation is sometimes a degree of freedom in design or operation, e.g. the recovery of a reactant for recycle.

When considering the energy efficiency of distillation processes, it is important that like-for-like comparisons are made between options. In the following discussion, it is assumed that the separation, specified in terms of product purity or recovery (with respect to a single component or a group of components), is fixed.

6.2.1 Energy-efficient distillation: conceptual design of simple columns

A “simple” distillation column is defined as one in which a single feed is separated into two products, where the column has a single reboiler and a single condenser. Conceptual design of the distillation column means selecting the operating conditions and design parameters that will ensure that the column can carry out the specified separation.

A simple distillation column requires heat in the reboiler, to create a vapor stream that flows upwards through the column and cooling in the condenser, to provide liquid reflux in the column. The column operates over a pressure range, where the pressure at the bottom of the column is greater than that at the top of column; the pressure drop is a result of friction in the column and the hydraulic head of liquid within column. The pressure drop is typically around 0.3—1.0 kPa per distillation stage, depending on the type of internals [13–15]. In the following discussion, uniform pressure in the column will be assumed in order to simplify analysis.

Heat must be provided to the reboiler at the bottom of the column, where the mixture being separated has a relatively high boiling point, being enriched in less volatile components. The condenser, at the top of the column, rejects heat at a lower temperature, as the mixture being condensed is concentrated in more volatile components. Unfortunately, because of this temperature difference, heat rejected by the condenser cannot be reused by the reboiler.

6.2.1.1 Degrees of freedom in design

This section considers the design of a simple distillation column to separate a given feed into products that achieve given separation requirements (e.g. product purity or component recovery to a product). The feed flow rate and composition are taken to be fully defined. There are several degrees of freedom in the design of the column that can be manipulated to promote energy efficiency. The feed temperature and pressure can be manipulated. The column operating pressure must be set (and the associated pressure drop must be estimated). The number of theoretical stages, feed stage location and the type of condenser (to produce an overhead product
that is in the liquid phase, vapor phase, or both) must be chosen. All of these design decisions impact on the heating and cooling requirements of the column and the associated temperatures.

Heating and cooling in distillation processes often takes place over a range of temperatures. The following discussion makes the conservative assumption that a hot utility or heat source must be hot enough, or a cold utility or heat sink must be cold enough, to provide all of the heating or cooling, as illustrated in Figure 6.11. Opportunities to use more than one heat source or sink, or heating or cooling utilities that operate over a range of temperatures (e.g. hot oil, mixed refrigerants), are not considered in the following discussions.

Table 6.4 presents data relevant to an industrial distillation process separating \textit{iso}-butane from \textit{n}-butane [16] that will be used to illustrate how column design affects heating and cooling requirements.

![Figure 6.11 Utility Temperatures are Set by Heating and Cooling Temperatures](image)

**Table 6.4 Summary of Illustrative Distillation Process\textsuperscript{a}**

<table>
<thead>
<tr>
<th>Feed</th>
<th>Column Design</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate</td>
<td>26,122 kg h(^{-1})</td>
</tr>
<tr>
<td>Pressure</td>
<td>650 kPa</td>
</tr>
<tr>
<td>Temperature</td>
<td>55.4 °C</td>
</tr>
</tbody>
</table>

**Feed Composition (by Mass)**

<table>
<thead>
<tr>
<th>Feed Composition</th>
<th>Specifications</th>
</tr>
</thead>
<tbody>
<tr>
<td>Propane</td>
<td>1.54%</td>
</tr>
<tr>
<td>\textit{iso}-Butane</td>
<td>29.5%</td>
</tr>
<tr>
<td>\textit{n}-Butane</td>
<td>67.7%</td>
</tr>
<tr>
<td>\textit{iso}-Butene</td>
<td>0.13%</td>
</tr>
<tr>
<td>1-Butene</td>
<td>0.20%</td>
</tr>
<tr>
<td>\textit{neo}-pentane</td>
<td>0.11%</td>
</tr>
<tr>
<td>\textit{iso}-Pentane</td>
<td>0.77%</td>
</tr>
<tr>
<td>\textit{n}-Pentane</td>
<td>0.08%</td>
</tr>
</tbody>
</table>

**Products**

- Distillate: 8114 kg h\(^{-1}\)
- Bottom product: 18,008 kg h\(^{-1}\)

\textsuperscript{a} Adapted from an industrial example [16]; Peng-Robinson equation of state is used to model the column in Aspen HYSYS v. 7.3.
6.2.1.2 Column operating pressure

In the discussion that follows, the influence of the operating pressure on the distillation performance will be considered; the pressure drop is assumed to be negligible. The column operating pressure is the most important distillation design parameter; it affects the temperatures at which heating and cooling are required, the type of heating and cooling utility required, as well as heating and cooling duties. Furthermore, it impacts on the number of theoretical stages needed for the separation and the diameter of the column.

Firstly, the distillation operating pressure affects the temperatures of heating and cooling. The bubble point and dew point temperatures of a mixture of a given composition depend strongly on pressure. For mixtures in which the vapor phase behaves as an ideal gas and the liquid phase behaves as an ideal liquid solution [14], Raoult’s law can be used, together with a suitable vapor-pressure model, to predict the pressure—temperature relationships. For less ideally behaved mixtures, other fluid property models, including cubic equations of state and activity coefficient models, are needed.

Higher operating pressures increase heating and cooling temperatures, as illustrated in Figure 6.12(a) for the case presented in Table 6.4. In general, these higher temperatures may mean that a hotter heat source is needed in the reboiler (e.g. MP steam.
rather than LP steam) and heat is rejected at a higher temperature in the condenser (e.g. more moderate refrigeration, or cooling water rather than refrigerated cooling).

Secondly, as illustrated in Figure 6.12(b), as temperatures increase with increased pressure, the components in the mixture typically become more similar in volatility [17] and their separation becomes more difficult. With this decrease in the driving force for the separation (the difference in volatilities of the components being separated), the distillation column will need more reflux and more theoretical stages to compensate; Figure 6.12(c) indicates that the minimum reflux ratio and minimum number of stages both increase with pressure.

With a higher operating pressure, the capital cost of the column will increase. A taller column will be needed, to accommodate a greater number of theoretical stages and a thicker shell may be needed to withstand the pressure. A wider column may or may not be needed, as the diameter of the column is dictated mainly by volumetric flow of the vapor; although the molar flow of vapor increases with pressure, so does its density.

The increase in reflux resulting from the pressure increase is likely to increase the reboiler and condenser duties (although this may be offset by the decrease in the latent heat of vaporization as the pressure increases); Figure 6.12(d) provides an illustration. The impact on operating costs depends on both the duty and the utility—if, at the higher temperature, steam at a higher temperature were needed, the cost per unit of heat would increase. However, if the higher temperature raised the temperature at which refrigeration were required, or allowed refrigeration to be avoided completely, the cost per unit of cooling could decrease quite significantly.

In general, therefore, the combination of these effects suggests that it is best to operate at as low a pressure as is practical and which avoids the use of refrigeration. While operating under vacuum makes the separation easier, the low vapor density requires a greater cross-sectional area in the column to accommodate the high volumetric flow of material. More importantly, operating under vacuum increases the complexity of the flowsheet, requiring additional, costly equipment to draw the vacuum and recover material drawn into the vacuum system. Vacuum also introduces safety hazards, related to risk of air ingress and resulting fire or explosion of flammable process fluids.

The logical conclusion is that it is best to operate at atmospheric pressure unless there are good reasons not to [3]. In particular, if:

1. increasing the operating pressure allows refrigeration to be avoided or facilitates the use of refrigeration at more moderate conditions;
2. operating under a vacuum avoids degradation of thermally sensitive materials because of the lower temperatures in the column;
3. changing the pressure (up or down) creates an opportunity for heat recovery within the wider process (see Section 6.4);
4. the cost of compressing the feed products (to meet downstream specifications) outweighs the benefits of increased operating pressure.

In the design of distillation columns, it is important that the operating pressure is considered first, as the other degrees of freedom for design are all strongly
influenced by the operating pressure and as the operating pressure influences most strongly which utilities will be required.

6.2.1.3 Pressure drop
Distillation benefits from low operating pressures in terms of heating and cooling duties; the corollary is that high pressure drop in a column is detrimental to energy efficiency, as part of the column will operate at pressures well above the minimum pressure in the column, where volatilities are lower and more reflux is required to compensate. Pressure drops increase column heating and cooling duties, especially in columns operating at atmospheric or subatmospheric pressures [18]. Higher pressure drop is particularly detrimental in vacuum distillation, as the energy required to draw the vacuum is wasted in a context where the separation is particularly sensitive to volatilities. In these cases, additional separation stages may increase, rather than decrease energy requirements, because of the increased pressure drop [18]. Especially in columns operating under vacuum, it is beneficial to use distillation internals and auxiliary equipment (e.g. trays or packing, gas and liquid distributors, mist eliminators, reboiler, condenser) with low pressure drop.

In the example presented in Table 6.4, a pressure drop of 0.5 kPa per stage increases the reboiler duty by up to 2%. An additional pressure drop of 50 kPa each in the condenser and reboiler increases this penalty by a further 3%.

6.2.1.4 Number of theoretical stages
The number of theoretical stages required to carry out a specified separation in a simple distillation column (at a given operating pressure) is not unique. Some minimum number of stages is required to achieve the separation [19]. For different numbers of stages, different reflux ratios will be required, corresponding to different reboiler and condenser duties, as illustrated in Figure 6.13(a). A key design decision is to select the number of stages in the column. There is a trade-off between operating costs and capital investment—increasing the number of stages would increase
the height of the column and therefore its cost, but would decrease reflux requirements and hence reduce duties and operating costs, as well as costs of heat transfer equipment.

A heuristic rule, based on a wide range of mixtures, separation specifications, operating pressures, etc., is that the reflux ratio should be in excess of the minimum reflux ratio by approximately 10%, as illustrated in Figure 6.13(b). In difficult separations, where high reflux ratios are required, or in cases that refrigeration is required, a lower excess reflux ratio may be applied. However, operating too close to the minimum reflux ratio can make the process highly sensitive to operational fluctuations or errors in the design calculations, e.g. estimated tray efficiencies or predicted phase equilibrium behavior [5]. If heating and cooling are inexpensive, e.g. where recovered heat is used in the reboiler, higher reflux ratios may be suitable.

6.2.1.5 Feed condition

The temperature (or “thermal condition”) of the feed is another important degree of freedom. It is generally the case that a feed (at the column pressure) should enter the column as a saturated liquid [20] or at a temperature between its bubble point and dew point temperatures. A superheated feed or a subcooled feed will introduce thermodynamic inefficiencies, as it will need to be cooled or heated, respectively, to saturation conditions to participate in the separation processes within the column.

The thermal condition of the feed entering a distillation column impacts directly on the flow rate of vapor and of liquid within the column, as illustrated in Figure 6.14(a). In turn, the feed condition affects the amount of vaporization required in the reboiler and the amount of condensing required and thus the heating and cooling duties. The feed condition therefore presents an opportunity to manipulate the heating and cooling duties of the column, by using a feed preheater or pre-cooler, as shown in Figure 6.14(b). The total heating and cooling duties tend to be least when the feed enters as a saturated liquid [20], so from the point of view of duty only, a saturated liquid feed is most energy efficient. It is often appropriate to assume that the total heating and cooling duties of the column remain unchanged when the feed condition changes. However, as shown in Figure 6.14(c), the total heating duty for a saturated vapor feed is around 16% more than that for a saturated liquid feed for the example of Table 6.4.

The hot utility suitable for heating the feed may be cooler and cheaper than that used in the reboiler, as the bubble point and dew point temperatures of the feed are usually lower than that in the reboiler. Using feed heating may thus decrease the total cost of providing heat (in both the reboiler and preheater). An analogous argument applies to precooling the feed. Whether preheating (or precooling) the feed reduces operating costs depends mainly on whether a cheaper, more energy-efficient heating medium can be employed. For the illustrative example, Figure 6.14(d) shows that the feed and reboiler temperatures are relatively similar in this near-binary separation—it is likely that heating the feed would actually increase total heating costs.
6.2.1.6 Feed stage location

When a feed stream enters a distillation column, it mixes with the liquid and vapor streams at the feed stage and participates in the mass transfer processes on the feed stage. The stage onto which the feed is introduced is a degree of freedom in the design. If the feed composition or temperature are very different to those on the feed stage, the mixing of the feed with the material within the column disrupts the composition profile in the column. These “mixing effects” are thermodynamically inefficient and cause the heating and cooling duties of the column to increase [17]. On the other hand, if the composition and temperature of the feed and the feed stage are similar, then the feed scarcely influences the mass transfer taking place on the feed stage, which is more energy efficient.

The above argument is applicable to both single-phase and two-phase feeds; the compositions of both the liquid phase and the vapor phase of the feed should ideally match those on the feed stage [3,5]. For the illustrative example of Table 6.4,
selecting a feed stage four stages above or below the best feed stage (35) increases the heating and cooling duty by around 1.5%.

6.2.1.7 Condenser type
The phase of the overhead product of a distillation column may be a design degree of freedom, if, for example, the phase of the distillation product is not specified. Figure 6.15 illustrates that the overhead product may be a liquid and/or a vapor. A “total condenser” produces a liquid product; all the overhead vapor is condensed to its bubble point, so the condenser duty is maximized and the product temperature is minimized. Where the cooling medium is expensive, as in refrigerated condensers, this may lead to high operating costs. Instead, a “partial condenser” (in which additional separation takes place [3]) can provide a vapor-phase product and liquid reflux; as less material is condensed, the condenser duty is lower and cooling is required to a more moderate temperature.

Especially where components are present that are difficult to condense at the column operating pressure, a “mixed condenser” may offer a compromise [3], i.e. providing a liquid product and liquid reflux but also producing a vapor product. If the overhead product is itself the feed to a distillation column, the energy demand of the downstream column will be affected by the type of condenser of the upstream column.

6.2.1.8 Summary—column design for energy efficiency
Key degrees of freedom for column design affect both the quantity and quality of heating and cooling required. The column operating pressure is significant, as it affects heating and cooling duties, as well as the temperatures involved. The number of stages in the column also determines duties; the trade-off between energy demand and capital investment is noteworthy. The feed condition, feed stage location and phase of the distillate product should also be selected considering energy efficiency. This section assumes that all heating and cooling is provided by utilities, such as steam, cooling water and refrigeration. Section 6.4 discusses how exploiting heat recovery opportunities can also enhance energy efficiency. Furthermore, Section 6.5 presents a range of complex column configurations that can improve energy efficiency.
It is sometimes assumed that there are strong trade-offs between energy efficiency and capital investment; however, energy-efficient designs may also have reduced capital costs, for example, if a more compact column, with a smaller diameter, is needed because of reduced vapor and liquid flows in the column.

6.3 Energy-efficient distillation: operation and control

6.3.1 Energy-efficient column operation

In practice, distillation columns designed for energy efficiency may not be operated efficiently. Plants need to be monitored, maintained, and managed for energy efficiency and causes of inefficiency and solutions to overcome these, need to be identified [21]. A column operated inefficiently, compared to design conditions, may use utilities in excessive amounts or at excessively high (or low) temperatures, may not achieve expected throughputs, or may not meet product specifications, impacting on process economics.

Economic trade-offs exist between reflux, production rate, and product purity in an operational distillation column. If a higher-purity product can be sold for a higher price that more than compensates for the additional energy required, it may make sense to use higher reflux in the column. However, higher reflux will increase energy demand and may also lead to flooding, restricting the production capacity of the column. Conservative operating policies, where reflux and product purities are higher than required, thus reduce energy efficiency and potentially production capacity, i.e. increase costs and limit product revenue [21].

A number of operational problems can reduce energy efficiency. Fouling or polymerization in the reboiler can reduce heat transfer rates, increasing steam demand or requiring steam at higher temperatures. Steam flows will also be excessive if some of the steam does not condense in process heaters; appropriate maintenance and operation of the steam system and steam traps can avoid this problem. A high pressure drop may indicate plugging or fouling of column internals or flooding in the column, with associated increases in heating and cooling duties [21]. If column internals are broken, damaged, or poorly installed, there may be effectively fewer theoretical stages in the column, so additional reflux will be needed to compensate. Poorly insulated equipment may have significant heat losses, increased utility demand. If the feed or product specifications change, compared to the design case, column operating conditions, e.g. feed temperature, reflux ratio, reflux temperature, may need to be adjusted accordingly to operate efficiently.

6.3.2 Process control

The control system for a distillation process manipulates operating parameters to meet process specifications. A column that is controlled inappropriately may operate inefficiently with respect to energy consumption and production rates. Further detail on the control of distillation processes is presented in a later chapter [22].
If control of the column is not tight but the set points are appropriate, then the product quality may, on average, meet specifications and the energy demand, on average, should be acceptable. In other words, deviations around the set point are of less importance than the set point itself [23]. Process simulation and optimization can identify the most appropriate set points for control variables to minimize the use of the most costly utilities, given the physical constraints of existing equipment. Control to avoid subcooling the reflux may reduce demand for both hot and cold utilities.

In winter months, it may be possible to reduce the column operating pressure to take advantage of colder ambient temperatures. The higher volatilities at the lower pressure should allow the reflux to decrease and thus reduce the condenser and reboiler duties as well as the feed preheat temperature—the control set points should all be adjusted accordingly.

### 6.3.3 Summary—operation and control

The energy efficiency of a well-designed distillation process can be promoted by effective management, maintenance, control and operation of the plant. How the column is operated affects the separation performed, production capacity and energy demand and in turn process economics and energy efficiency. Judicious choice of control set points, along with effective control strategies, can facilitate energy-efficient column operation.

### 6.4 Heat integration of distillation

A heat-integrated process is one in which heat is recovered from process streams requiring cooling and reused by process streams requiring heating. Heat is recovered using heat exchangers, which facilitate heat transfer from one process stream to another. Figure 6.16 illustrates a heat exchanger in which heat is transferred from a “hot” stream, requiring cooling, to a “cold” stream, requiring heating. The hot stream is also known as a heat source and the cold stream as a heat sink.

The recovery and reuse of heat reduces demand for heating and cooling utilities and thus increases the energy efficiency of a process; the improved energy efficiency

![FIGURE 6.16 A Heat Exchanger Facilitates Heat Recovery from a “Hot” Stream to a “Cold” Stream](image)
has a positive impact on operating costs, environmental impact and sustainability, as discussed in Section 6.1.

### 6.4.1 Heat exchange and heat recovery

For heat to flow from one stream to another in a heat exchanger, the heat source must be at a temperature that is hotter than the heat sink. The greater the temperature difference between the streams, the greater the driving force for heat transfer, so the lower the heat transfer area required. Equation (6.2) presents a simple model for heat exchangers that relates the heat transfer duty \( \dot{Q}_{ht} \) to the logarithmic mean temperature difference \( \Delta T_{lm} \), the overall heat transfer coefficient \( U \) and the heat transfer area \( A_{ht} \) [5]:

\[
\dot{Q}_{ht} = U \cdot A_{ht} \cdot \Delta T_{lm}
\]

(6.2)

The cost of the heat exchanger, which is highly correlated with heat transfer area, is lowest for large temperature differences. However, fewer opportunities for heat recovery will exist if a high minimum temperature approach in the exchanger is specified, which will in turn increase operating costs. In above-ambient processes, this trade-off typically leads to a minimum temperature approach of 10°C or greater; in subambient processes, the high cost of refrigeration can justify the use of much lower temperature differences, of around 5°C or even 1°C [3].

A process may employ several heat exchangers to recover heat from one or more hot streams. Together, these exchangers form a heat exchanger network. The design of the network can be challenging, especially when several process streams requiring heating or cooling are present. Systematic approaches have been developed for design of heat exchanger networks that maximize heat recovery. One such approach, known as pinch analysis, which is based on thermodynamic analysis, is useful for identifying the maximum heat recovery that can be achieved, for a given minimum temperature approach [3—5]. Furthermore, the “pinch” in pinch analysis indicates a constraint for heat recovery: below the pinch temperature, it does not make sense for the process to act as a sink for heat available for recovery and above the pinch temperature it is not sensible for the process to reject heat.

In the context of energy-efficient distillation, it is pertinent to ask under which conditions it is beneficial for heat rejected by a column to be recovered for reuse and under which conditions it makes sense for recovered heat to be reused by a column. The pinch concept helps resolve this question.

For a given process (that is, including various streams requiring heating and cooling but excluding the distillation column), the pinch temperature can be determined for a set value of the minimum temperature approach [3,4]. Compared to the case where utilities supply all the heating and cooling of a distillation column, the total utility demand (of the column and overall process together) will not be reduced if the column rejects heat at a temperature below the pinch and acts as a heat sink at a temperature above the pinch. This condition may be quickly and conveniently applied to identify situations in which there is no benefit of recovering heat between the process and the column. In the converse situation, where the temperatures at which
heat is rejected by the column and reused by the column are both above or both below the pinch, the total heating or cooling demand (of the column and process together) may be reduced. A decrease in total utility demand may not lead to a decrease in operating cost, as utility costs depend strongly on temperature. Note that heat may be rejected by the condenser or feed cooler, as discussed in Section 6.4.2 and the reboiler and feed heater are heat sinks. Section 6.5 considers heat integration of complex column configurations.

6.4.1.1 Summary—heat exchange and heat integration
Heat integration can greatly enhance the energy efficiency of distillation processes. For heat recovery, an available heat source needs to be hotter than a process heat sink; furthermore, pinch analysis recommends that heat recovery between a distillation column and an associated process is only implemented if the temperatures of all streams exchanging heat are either above or below the pinch temperature.

6.4.2 Distillation design for heat integration
Heat recovery can increase the energy efficiency of a distillation column of a given design (given operating conditions, number of theoretical stages, feed condition, etc.). Moreover, the design of the column could potentially be manipulated to enhance or create opportunities for heat integration. The key design degrees of freedom discussed in Section 6.2.1 impact on heating and cooling requirements (both quality and quantity) and therefore on opportunities for heat integration.

6.4.2.1 Operating pressure
As discussed in Section 6.2.1, increasing the column operating pressure raises the temperature in the condenser and reboiler but is also likely to increase their duties. If a feed heater or cooler is present and operates at an increased pressure, the temperatures of the feed heater or cooler will also increase. Changing the operating pressure may therefore present an opportunity to manipulate the temperatures of the column heat sinks and sources so they lie entirely above, or entirely below, the pinch temperature. In this case, it may be possible to satisfy column heating and cooling requirements using process heat sinks and sources and therefore reduce overall utility requirements.

In columns operating at subambient temperatures, in which refrigerated condensing is required, the operating pressure impacts significantly on the distillation condenser temperature and therefore on the refrigerant evaporation temperature. Furthermore, the column reboiler may provide a useful heat sink for the refrigeration cycle rejecting heat from its condenser. As Eqn (7.1) shows, the energy efficiency of a refrigeration cycle can be increased by reducing the temperature of the condenser (i.e. allowing heat to be rejected to a colder heat sink). A column reboiler may act as such a heat sink; the column operating pressure can sometimes be manipulated to increase the energy efficiency of a refrigeration cycle by reducing the compression power demand of the cycle. Tables 6.5, 6.6 and 6.7 summarize an illustrative example.

In this example, in which light hydrocarbons are distilled to separate ethene and methane, the column and refrigeration system performance are modeled using simple
### Table 6.5 Summary of Light Hydrocarbons Distillation Process

<table>
<thead>
<tr>
<th>Feed</th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Flow rate</td>
<td>6000 kmol h⁻¹</td>
<td></td>
</tr>
<tr>
<td>Pressure</td>
<td>1000 kPa</td>
<td></td>
</tr>
<tr>
<td>Temperature</td>
<td>−70 °C</td>
<td></td>
</tr>
<tr>
<td>Vapor fraction</td>
<td>0.248</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Feed composition (mole fraction)</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Methane</td>
<td>0.24</td>
</tr>
<tr>
<td>Ethene</td>
<td>0.46</td>
</tr>
<tr>
<td>Ethane</td>
<td>0.20</td>
</tr>
<tr>
<td>Propene</td>
<td>0.08</td>
</tr>
<tr>
<td>Propane</td>
<td>0.02</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Specifications</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Methane recovery to distillate</td>
<td>99%</td>
</tr>
<tr>
<td>Ethene recovery to bottom product</td>
<td>96%</td>
</tr>
</tbody>
</table>

### Table 6.6 Summary of Column Design for Distillation of Light Hydrocarbons

<table>
<thead>
<tr>
<th>Column Design Parameter</th>
<th>1000 kPa</th>
<th>2000 kPa</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ratio R/Rmin</td>
<td>1.10</td>
<td>1.10</td>
</tr>
<tr>
<td>Minimum reflux ratio</td>
<td>0.607</td>
<td>0.750</td>
</tr>
<tr>
<td>Actual reflux ratio</td>
<td>0.668</td>
<td>0.825</td>
</tr>
<tr>
<td>Number of stages</td>
<td>10.9</td>
<td>13.4</td>
</tr>
<tr>
<td>Feed stage (from top)</td>
<td>3.1</td>
<td>3.8</td>
</tr>
<tr>
<td>Distillate flow rate</td>
<td>1543 kmol h⁻¹</td>
<td>1542 kmol h⁻¹</td>
</tr>
<tr>
<td>Bottom product flow rate</td>
<td>4457 kmol h⁻¹</td>
<td>4458 kmol h⁻¹</td>
</tr>
<tr>
<td>Feed pressure</td>
<td>1000 kPa</td>
<td>2000 kPa</td>
</tr>
<tr>
<td>Feed temperature</td>
<td>−70 °C</td>
<td>−52.4 °C</td>
</tr>
<tr>
<td>Condenser duty</td>
<td>12.4 GJ h⁻¹</td>
<td>12.2 GJ h⁻¹</td>
</tr>
<tr>
<td>Condenser temperature</td>
<td>−104.8 °C</td>
<td>−91.6 °C</td>
</tr>
<tr>
<td>Reboiler duty</td>
<td>16.8 GJ h⁻¹</td>
<td>24.1 GJ h⁻¹</td>
</tr>
<tr>
<td>Reboiler temperature</td>
<td>−43.4 °C</td>
<td>−18.3 °C</td>
</tr>
</tbody>
</table>

*a Peng-Robinson equation of state is used to model the column in Aspen HYSYS v. 7.3. The shortcut distillation model used to design and simulate the distillation column. The pressure drop within the column is assumed to be negligible.*
6.4 Heat integration of distillation

models—the short-cut distillation design models and the Carnot model. Refrigeration is required to condense the methane-rich overhead product. Table 6.6 shows how the condenser temperature is affected by operating pressure and Table 6.7 illustrates the effect on the compression power demand. In the case that there is no heat integration, the higher-pressure column has a lower power demand, considering both feed compression and the refrigeration system. Table 6.7 also illustrates the potential benefits of heat integration—rejecting heat from the refrigeration condenser to the column reboiler reduces the power demand of the refrigeration system. (Note, however, that for the column operating at 1000 kPa, the reboiler duty is smaller than the refrigeration condenser duty, so the proposed heat recovery scheme is only partially feasible.)

An additional benefit of using a higher pressure in the distillation column may be obtained if the difference in temperature between the evaporator and condenser temperature of the cycle is small enough to avoid using cascaded refrigeration cycles

<p>| Table 6.7 Summary of Refrigeration Requirements for Light Hydrocarbons Separation&lt;sup&gt;a,b&lt;/sup&gt; |
|-----------------------------------------------|------------------|------------------|------------------|------------------|</p>
<table>
<thead>
<tr>
<th>Column Operating Pressure</th>
<th>Without Heat Integration</th>
<th>With Heat Integration</th>
<th>Without Heat Integration</th>
<th>With Heat Integration</th>
</tr>
</thead>
<tbody>
<tr>
<td>1000 kPa</td>
<td>2000 kPa</td>
<td>1000 kPa</td>
<td>2000 kPa</td>
<td></td>
</tr>
<tr>
<td>Condenser duty (GJ h&lt;sup&gt;-1&lt;/sup&gt;)</td>
<td>12.4 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>12.2 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>12.4 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>12.2 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
</tr>
<tr>
<td>Condenser temperature (°C)</td>
<td>-104.8 °C</td>
<td>-91.6 °C</td>
<td>-104.8 °C</td>
<td>-91.6 °C</td>
</tr>
<tr>
<td>Reboiler duty (GJ h&lt;sup&gt;-1&lt;/sup&gt;)</td>
<td>16.8 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>24.1 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>16.8 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>24.1 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
</tr>
<tr>
<td>Reboiler temperature (°C)</td>
<td>-43.4 °C</td>
<td>-18.3 °C</td>
<td>-43.4 °C</td>
<td>-18.3 °C</td>
</tr>
<tr>
<td>Refrigerant evaporation temperature (°C)</td>
<td>163.4 K</td>
<td>176.6 K</td>
<td>163.4 K</td>
<td>176.6 K</td>
</tr>
<tr>
<td>Refrigerant condensing temperature (°C)</td>
<td>Ambient</td>
<td>Ambient</td>
<td>Reboiler</td>
<td>Reboiler</td>
</tr>
<tr>
<td>Ideal compression power demand (GJ h&lt;sup&gt;-1&lt;/sup&gt;)</td>
<td>11.0 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>9.1 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>5.5 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>5.8 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
</tr>
<tr>
<td>Actual compression power demand (GJ h&lt;sup&gt;-1&lt;/sup&gt;)</td>
<td>18.4 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>15.2 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>9.1 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>9.6 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
</tr>
<tr>
<td>Refrigerant condensing duty (GJ h&lt;sup&gt;-1&lt;/sup&gt;)</td>
<td>30.8 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>27.4 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>21.5 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>21.8 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
</tr>
<tr>
<td>Feed compression power demand&lt;sup&gt;b&lt;/sup&gt; (GJ h&lt;sup&gt;-1&lt;/sup&gt;)</td>
<td>-</td>
<td>2.0 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
<td>-</td>
<td>2.0 GJ h&lt;sup&gt;-1&lt;/sup&gt;</td>
</tr>
</tbody>
</table>

---

<sup>a</sup> Refrigerant power demand is estimated using the Carnot model, assuming that the ideal power demand is 60% of the actual power demand and the minimum temperature approach is 5 °C in the evaporator and condenser. In the "With heat integration" cases, heat from the refrigerant condenser is rejected to the column reboiler.

<sup>b</sup> A centrifugal compressor with an isentropic efficiency of 85% is used to compress the vapor fraction of the feed.

---
that are required to overcome the limitations in the operating range of refrigerants [3]. In the example presented in Table 6.7, it would be feasible to use a single refrigerant, ethane, when heat integration is implemented for the column operating at 2000 kPa, while a cascade of two or three refrigeration cycles would be required if the refrigerant were condensed using an ambient cooling medium.

6.4.2.2 Number of theoretical stages

The reflux ratio required to carry out a given separation at a given operating pressure depends on the number of theoretical stages present. When heating and cooling are provided by process utilities, the trade-off between operating and capital costs suggests a reflux ratio of 10% greater than the minimum reflux ratio. In a heat-integrated column, where heating and cooling are provided by recovered heat, the operating costs per unit of heating or cooling are likely to be much lower than for utilities or even negligible. Therefore, a very different trade-off—between the column capital cost, the cost of heat recovery equipment and column operating costs—is likely to exist and the most appropriate number of theoretical stages (and thus reflux ratio) could be very different to that in the case that the heating and cooling needs of the column are met solely by utilities [3].

6.4.2.3 Feed condition

The effect of heating or cooling the feed stream is discussed in Section 6.2.1. Compared to a design in which the column feed is a saturated liquid, preheating the feed can allow the use of heat at a more moderate temperature than that in the reboiler. This more moderate temperature may allow the feed heater to use process heat sources that are not sufficiently hot to serve the reboiler; although the total heating duty may increase slightly as a result of feed heating, energy efficiency may increase because of heat recovery. An analogous effect may be obtained by using a feed cooler, where heat is rejected at a more moderate (warmer) temperature that allows its reuse by the associated process.

6.4.2.4 Summary—distillation design for heat integration

Significant benefits, with respect to energy demand, operating costs and environmental impact, can be achieved by heat integrating distillation columns with an associated process. Direct heat recovery, between heat sinks and sources of a distillation column and an associated process, as well as indirect heat recovery, between a distillation column and a refrigeration cycle, can significantly reduce utility demand. The design degrees of freedom of a distillation column can potentially be exploited to create or enhance opportunities for heat recovery.

6.5 Energy-efficient distillation: advanced and complex column configurations

So far, the discussion has only considered simple distillation columns. Complex column configurations can increase the energy efficiency of distillation. These
configurations include those related to a single separation, i.e. the use of intermediate heating and cooling, “double-effect distillation” (also known as two-pressure distillation), open-cycle heat pumping (also known as vapor recompression) and internally heat-integrated column arrangements, as well as those related to two or more separations, including thermally coupled columns and prefractionation arrangements.

6.5.1 Intermediate heating and cooling (side-reboilers and side-condensers)

In a conventional distillation column (and ignoring feed heating), heat is provided to the column in the reboiler, with the highest temperature in the column and heat is rejected from the condenser, with the lowest temperature. Therefore, all the heat is supplied at a temperature above that in the reboiler. Instead, heat can be supplied at an intermediate temperature, at an intermediate point in the column, using a so-called side-reboiler or inter-reboiler, as illustrated in Figure 6.17(a).

Compared to the simple column, introducing some of the heat at a more moderate temperature could improve energy efficiency. On the other hand, the internal reflux is lower between a reboiler and a side-reboiler. As a result, less separation may take place on those stages—the total heating duty may increase, or additional stages may be needed, to compensate for the lower reflux. Thus, there is a trade-off between the quality and the quantity of heat that is needed. Furthermore, there is a limit to the amount of heat that can be provided by the side-reboiler. If the side-reboiler duty is too high, then the internal reflux ratio may be reduced to below the minimum level for the required separation to be feasible (i.e. a new pinch may form in the column) [24]. The use of side-condensers is analogous to that of side-reboilers. Where the temperature of the side-reboiler allows the use of recovered heat, the demand for a hot utility may be reduced, in spite of the increased total

FIGURE 6.17 Intermediate Heating and Cooling

(a) Column with a side-reboiler; (b) Column with a pump-around.
heat duty and so energy efficiency may increase, compared to the case without inter-
reboiling. Side-reboilers incur capital expenditure and increase the complexity of the
distillation column, however.

Inter-cooling, or side-condensing, similarly, can reduce the amount of heat rejected
by the condenser (at the lowest temperature in the column) and provide a potentially
useful heat source at a more moderate temperature. In practice, it is difficult to
withdraw a vapor stream from a column, condense it and return it to the column. It
is often more practical to withdraw a liquid stream, subcool it and return it to the
column; on entering the column, the subcooled liquid causes some of the vapor to
condense and to flow down the column, increasing the reflux below the return stage.

Such an arrangement is commonly applied in petroleum refining, for example, in
the atmospheric crude oil distillation unit, as a pump-around or pump-back loop
[25], as shown in Figure 17(b). The pump-around (or pump-back) streams are sub-
cooled before being returned to the column; the heat rejected in the pump-around
plays an important role in preheating the column feed, reducing the requirement
for fired heating.

6.5.2 Double-effect distillation

In a double-effect (or “two-pressure”) distillation configuration, a single separation
is carried out in two separate distillation columns operating at two different oper-
ating pressures, as illustrated in Figure 6.18(a). The feed to the process is split
into two streams and fed to columns 1 and 2; the pressure of the feed to column 2
is raised. This high pressure is selected to allow the heat rejected from the condenser

![Figure 6.18 Double-Effect Distillation Configurations](image)

(a) Simple column; (b) Prefractionator configuration [20].

Adapted and reprinted from Journal of Natural Gas Science and Engineering, Copyright 2012, with permission from Elsevier.
of column 2 to be transferred to the reboiler of the lower-pressure column 1 in a heat
exchanger. By exchanging heat between the condenser of column 2 and the reboiler
of column 1, the requirement for hot and cold utilities can be significant reduced, to
almost half of that required in a single column carrying out the same separation.

Such a substantial reduction in both heating and cooling duties could lead to a
significant reduction in operating costs. The drawbacks of this configuration relate
firstly to the additional capital expenditure required—the two columns each process-
ing about half of the total feed are likely to cost significantly more than a single col-
umn separating the whole feed [5]. Furthermore, the use of high pressure in column
2 is likely to increase the heating and cooling duty per unit of feed processed and the
increase in temperature is likely to incur increased costs per unit of heating if a more
expensive hot utility is needed. Furthermore, the more extreme reboiler temperature
in column 2 is likely to reduce opportunities for heat integration with the associated
process.

In spite of these drawbacks, double-effect distillation is applied industrially, e.g.
for ethanol—water and acetic acid—water separations, as well as in multiple-stage
evaporation processes [26,27]. Figure 6.18(b) illustrates the double-effect concept
applied to the prefractionator arrangement discussed in Section 6.5.5.

6.5.3 Heat pumping (vapor recompression)

Heat rejected from the condenser of a simple distillation column cannot be reused
directly by the reboiler of the column, because the condenser is colder than the
reboiler. Heat pumping may be applied to upgrade the heat rejected by the condenser
to a temperature that allows its reuse in the reboiler of the column (or in another heat
sink). Vapor compression is the most widely used technology for heat pumping.
Other heat pumping technologies, including absorption and compression-resorption
heat pumps, are being developed with a view to improving energy efficiency [26].

Figure 6.19 illustrates a column using a compressor to upgrade heat, known as
“mechanical vapor recompression”. After compression, the temperature of the over-
head vapor is sufficiently high to allow heat to be rejected from the column
condenser to the reboiler. Therefore, the reboiler does not require a hot utility. (A
trim cooler may also be needed to reject heat that cannot be used in the reboiler.)
Once the compressed vapor has been cooled in this way and after the distillate prod-
uct has been withdrawn, the pressure is reduced and the cooled, condensed material
returns to the column as reflux.

The drawback of this arrangement is that the overhead vapor requires compres-
sion, involving expensive compression equipment and a costly form of energy: power.
These costs are least for a small temperature difference between the condenser and
reboiler, for example, for close-boiling mixtures [26]. However, if a refrigeration cycle
using compression would otherwise be needed to condense the overhead vapor, then
this “open loop” heat pumping arrangement may offer advantages because of the
inbuilt heat integration and the direct heat transfer in the condenser-reboiler exchanger
(compared to the separate evaporator and condenser of the refrigeration cycle). In
practice, this arrangement is applied in a wide range of separations, including those involving moderately light hydrocarbons (C₃ to C₈), aromatics, alcohols and high-boiling solvents from aqueous solutions, etc. [28–30].

6.5.4 Internally heat-integrated distillation

Three complex column configurations that can enhance energy efficiency of distillation have been discussed; all of these features are combined in an internally heat-integrated distillation column, known as a HIDiC. Firstly, intermediate heating and cooling in a distillation column, as discussed in Section 6.5.1, can increase energy efficiency and enhance opportunities for heat recovery because of the more moderate temperatures involved. Secondly, as described in Section 6.5.2, double-effect distillation exploits the effect of operating pressure on heating and cooling temperatures to facilitate recovery of heat rejected by a high-pressure column to a low-pressure column. Thirdly, heat pumping, discussed in Section 6.5.3, uses compression work to elevate the temperature of the heat rejected by a column condenser to allow heat recovery.

An internally heat-integrated distillation column, as illustrated in Figure 6.20(a), has a high-pressure rectifying section and a separate low-pressure stripping section. Temperatures in the rectifying section are higher than those in the stripping section, allowing heat transfer between the column sections.

Effectively, the rectifying section uses intermediate cooling along its length and the stripping section uses intermediate heating along its length. The overhead vapor of the stripping section is compressed to raise its pressure to that of the rectifying section. The reboiler and condenser duties can thus be significantly reduced, or even eliminated, especially if a feed preheater is used [31]. As illustrated in Figure 6.20(b), the varying reflux ratio within each column section lends itself to even distribution of mass.
transfer driving forces through the column, which leads to more thermodynamically reversible behavior and therefore lower energy requirements [32].

Challenges for the HIDiC technology related to conceptual design, mass transfer modeling, separation and heat transfer equipment design and process control, have been addressed by research programs and in pilot plant studies. A variety of hardware has been applied in these studies, including two concentric shells, plate-fin contactors, conventional trays with heat transfer plates and “thin plate” contactors [31].

Energy savings of 30—60% have been demonstrated in these studies at bench- and pilot-plant scale, e.g. a column 27 m high with a diameter of 1.4 m [31]. In spite of intensive research and development efforts, concentrated in Japan since the mid-1980s, internally heat-integrated distillation columns have not to date been applied commercially. Ongoing research addresses heat transfer equipment, batch distillation, azeotropic mixtures, column retrofit, hybrid membrane-HIDiC configurations, reactive HIDiC, etc. [31].

6.5.5 Energy efficiency in distillation sequences

6.5.5.1 Energy-efficient sequences of simple columns

When a multicomponent mixture is to be separated into three or more products, a sequence of two or more simple distillation columns is needed. Once the sequence of separations is selected, each simple column could be designed for energy efficiency, considering column operating conditions, heat integration, etc. Complex
column configurations carrying out more than two separations that promote energy efficiency are described further in the following sections.

6.5.5.2 **Columns with side-draw products**
A side-draw stream (or side-stream) may be withdrawn from a single column, where the purity of the side-draw depends on the composition profile within the column. It is sometimes possible to withdraw a stream that is enriched in an intermediate-boiling component, producing three products from a single column. The capital investment in a single column could be substantially less than that required for two simple columns performing the same separation. However, a relatively pure side-draw product typically requires high reflux ratios in the column, unless the intermediate-boiling components are abundant in the feed. A single column with a relatively pure side-draw product is unlikely to require less heating and cooling than a sequence of two columns producing three products [33].

6.5.5.3 **Thermal coupling (side-strippers and side-rectifiers)**
In a sequence of simple distillation columns, feed stage mixing gives rise to thermodynamic inefficiencies, which translate to low energy efficiency. The exception is the special, but industrially unimportant, case of columns carrying out so-called “preferred” splits, where the aim of the separation is to separate the lightest and heaviest components in the mixture; in this case, mixing effects of the feed stage and where liquid reflux and vapor boil-up reenter the column may be negligible [20].

These mixing effects can be partly avoided by using thermal coupling, where material streams flowing between two distillation columns couple the transfer of material with that of heat. As illustrated in Figure 6.21(a), the vapor stream leaving

![Figure 6.21 Thermal Coupling](image)

**FIGURE 6.21 Thermal Coupling**
(a) In thermally coupled indirect sequence of columns, the side-draw from the downstream column provides reflux to the upstream column; (b) The equivalent side-stripper configuration.

Adapted and reprinted with permission from N.A. Carlberg, A.W. Westerberg, Industrial and Engineering Chemistry Research 28 (9) 1379–1386, Copyright 1989, American Chemical Society.
overhead may be fed directly to a downstream column that also provides liquid reflux to the upstream column. By avoiding this thermodynamic inefficiency, the total heating and cooling duties of the thermally coupled column configuration may be less than that of the corresponding sequence of simple columns [34].

As shown in Figure 6.21(b), thermally coupled configurations can also be viewed as side-stream columns where the side-stream undergoes further separation and the remaining material is returned to the main column. In this sense, thermally coupled columns can produce, energy efficiently, a high-purity side-stream product.

Thermally coupled distillation configurations can reduce the quantity of heating and cooling required by a sequence. However, as there is only one reboiler providing heat for both columns, all the heat is needed at a high temperature. Similarly, if there is only one condenser, all the heat rejected is at the lowest temperature in the sequence. Thus, there is a trade-off between the quantity and the quality of the heating and cooling required in the sequence. Especially where two different temperature levels facilitate the use of a cheaper utility or the recovery of heat, thermal coupling may have a detrimental effect on energy efficiency.

Conceptual design of thermally coupled distillation columns is challenging, as these columns present additional degrees of freedom for design and are more complex to model [34]. Important industrial applications of thermal coupling are the use of side-strippers in refinery distillation columns, such as the atmospheric crude oil distillation unit [25] and side-rectifiers in air separation (to purify intermediate-boiling argon) [35,36]. Thermally coupled configurations present practical challenges related to withdrawing part of a vapor stream from a column and transferring a vapor stream from a lower-pressure downstream column to an upstream column at a high pressure [35]. In practice, two thermally coupled columns are constrained to operate at similar pressures; this constraint limits opportunities to manipulate the operating pressure for energy efficiency.

6.5.5.4 Prefractionation arrangements

Instead of introducing a feed directly to a distillation column, where mixing effects will inevitably increase energy requirements, it is possible to partially separate the feed, in a “prefractionation” column, before it undergoes further separation in the column. Furthermore, prefractionation may allow composition profiles to develop in the column that lend themselves to the withdrawal of a relatively pure side-draw product.

Figure 6.22(a) presents such a prefractionation arrangement. By appropriately designing both the prefractionator, which carries out the preliminary separation between the components to be separated and the main column, the feeds to the main column can be introduced with little remixing taking place; the composition profiles in both columns can develop in thermodynamically efficient ways [20]. The total heating and cooling required by the sequence of columns may be significantly lower (around 20—40%, depending on the feed mixture, composition and product specifications) than that of a sequence of simple columns carrying out the same separations [20,27].
The prefractionator may also be thermally coupled with the main column, as shown in Figure 6.22(b); this arrangement is also known as a Petlyuk column. The main benefit of introducing thermal coupling is that only a single reboiler and a single condenser are needed, compared to the uncoupled arrangement of Figure 6.22(a). There is little or no benefit in terms of heating and cooling duties, but there is a penalty with respect to quality of heating and cooling, as heating and cooling are required at the most extreme temperatures in the column. Furthermore, thermal coupling reduces opportunities to operate the two columns at different pressures. Operation and control are also more complicated than for the uncoupled arrangement.

The thermally coupled arrangement can be implemented in a single shell, as shown in Figure 6.22(c). The prefractionation is carried out on one side of a vertical wall, or partition, while the final separation takes place on the other side of the wall as well as above and below it. Liquid reflux from the overhead condenser is split across the top of the wall; reboiled vapor from the reboiler distributes across the wall according to the pressure drop on either side of the wall.

This dividing wall column configuration has essentially the same heating and cooling requirements as the Petlyuk column and is as constrained with respect to operating pressure. The use of a single column shell can bring significant advantages with respect to capital investment (around 30% [3]) because of the single shell and reduced costs of piping, foundations, plot space, etc. Specialized internals are needed to accommodate the dividing wall; both trays and packings have been successfully applied in these columns [37–40]. Liquid distribution arrangements include passive (e.g. weirs) and active (e.g. control valves) systems. A key challenge is to design the column and internals to achieve the correct vapor flows; these flows are dictated by the pressure drop on each side of the wall [3,37].

Dividing wall distillation is an established technology, although design and operation remain challenging [38]. These columns have been applied industrially,
initially in specialty chemicals production (in the late 1980s), but also in petroleum refining and many other areas [38–40]. The range of applications includes operating pressures of 0.2–1000 kPa and a wide range of product purities; column diameters are typically 0.6–4 m, but larger columns have been constructed [38]. Control strategies, equally, are well established [37,38]. Dividing wall columns have been employed in process revamps and for azeotropic, extractive and reactive distillation applications [37,38].

Various extensions of the Petlyuk and dividing wall column concepts have been developed, including the so-called Kaibel column with more than three products, columns with multiple dividing walls and side-stripper (and side-rectifier) arrangements, as shown in Figure 6.23. While these configurations are relatively complex, they can potentially bring additional advantages in terms of energy demand and capital costs [20].

![FIGURE 6.23 Extensions to Dividing Wall Columns](image)

(a) Kaibel column; (b) Column with multiple dividing walls [20]; (c) Side-stripper dividing wall column.

Reprinted from Journal of Natural Gas Science and Engineering, Copyright 2012, with permission from Elsevier.

6.5.5.5 Summary—energy-efficient distillation sequences

Many distillation sequences can carry out a required separation; column operating conditions can be selected for energy-efficient operation. The problem of selecting and designing the most energy-efficient sequence can be tackled by developing suitable models and applying optimization techniques [34].

Nonconventional column configurations, such as thermally coupled columns and prefractionation arrangements, can bring significant benefits in terms of energy demand, but at the expense of higher quality energy and with constraints relating to operating pressure. These columns, while complex and specialized, may also require less capital investment than the corresponding sequences of simple columns.
6.6 Energy-efficient distillation: evaluation of energy requirements

Process modeling and analysis tools that assist with the evaluation of distillation energy efficiency can be used to quantify distillation energy requirements. Models for individual columns are outlined; sequences can be evaluated considering each column in turn. However, interactions between the columns need to be captured by models for distillation sequences and complex column configurations. It is assumed in this discussion that the same separation is carried out in all the design alternatives; it is sometimes also relevant to consider interactions between product quality (e.g. purity) or yield and energy requirements.

6.6.1 Shortcut distillation column models

Shortcut distillation models, such as the Fenske, Underwood and Gilliland models, are well established and provide estimates of the minimum reflux ratio, minimum number of stages and actual reflux (or number of theoretical stages) of a column carrying out a specified separation, given feed data and the operating pressure \[14,17,41\]. These models are useful for developing conceptual column and sequence designs and for comparing them; when used together with simple models for capital and operating cost, the shortcut models allow economic evaluation of alternative design concepts.

The shortcut models assume constant relative volatility and constant molar overflow. The Underwood equations, to calculate the minimum reflux ratio, are of most relevance for evaluating energy requirements. However, care is needed when applying the method in multicomponent separations, especially when some components distribute significantly between the top and bottom products \[3\]. Also, the results may be sensitive to the relative volatility values selected to represent a column, e.g. whether based on the feed mixture at the feed conditions or on the feed stage, or based on the compositions of the products, where these compositions are themselves estimated using shortcut models \[3,17\].

The assumption of constant relative volatility may be unrealistic not only for obviously nonideal mixtures, e.g. those forming azeotropes but also for apparently well-behaved mixtures, e.g. light hydrocarbons such as ethene—ethane mixtures. The assumption of constant molar overflow is least valid for mixtures where components have dissimilar latent heats of vaporization; carrying out a separate enthalpy balance over the condenser can help to relate minimum vapor flow at the column pinch to that at the top of the column \[17\]. Other extensions to the shortcut models allow some complex column configurations to be evaluated, such as those with side-draw products and multiple feeds \[17\], but the application of the models is less straightforward in these cases.

The shortcut models, together with data related to the thermal and thermodynamic properties of the components and mixtures involved, allow energy requirements to be determined. In particular, the effects on heating and cooling duties
and the associated temperatures of design degrees of freedom, namely column operating pressure (and pressure drop), reflux ratio (or number of stages), feed condition and condenser type, can be estimated. However, as the models provide no detail of material or energy flows or temperature profiles within column sections, they cannot be used directly to evaluate intermediate heating and cooling options.

In spite of their shortcomings, shortcut models have been found useful when incorporated into a range of methodologies for conceptual design of columns and sequences (e.g. [34,42]), facilitated by the relatively low computational demands of the shortcut models.

### 6.6.2 More rigorous distillation models

Instead of applying shortcut distillation models, rigorous models that carry out stage-by-stage calculations of the material balance, phase equilibrium and enthalpy balance may be used to evaluate distillation column and sequence designs. These models involve a relatively large number of equations (depending on the number of components and number of equilibrium stages) and the countercurrent nature of the flows in the column complicates their solution [14,17,41].

Fortunately, robust and efficient algorithms for solving these equations simultaneously are well established and have been implemented in commercial process simulation software. The accuracy of these models is limited by the assumption that phase equilibrium is achieved on every theoretical stage; the use of stage efficiencies is a convenient, if rather empirical, way to represent mass transfer limitations. However, stage efficiencies are not straightforward to predict accurately and often are inadequate for representing mass transfer effects [41,43].

### 6.6.3 Distillation process performance indicators

Distillation models may be applied to predict and evaluate the performance of design alternatives. The separation specifications should be set appropriately, e.g. in terms of product purity or recovery, rather than in terms of product flow rate and reflux ratio, to permit valid comparisons. When evaluating design alternatives for distillation columns or sequences, it is reasonable to focus on energy demand, as operating costs typically dominate distillation process economics. In above-ambient processes, the cost of heating typically far outweighs the cost of cooling and in processes operating below ambient temperatures, the cost of refrigerated cooling is the overriding cost.

Various indicators have been applied for evaluating the performance of alternative distillation column or sequence designs; these are discussed following:

1. **Minimum or actual vapor load** [3,42]—The molar flow of vapor (known as vapor load) is readily calculated using shortcut distillation models and is correlated with actual heating and cooling duties. It does not account for energy quality, i.e. the temperatures at which heating and cooling are required, nor capital investment.

2. **Total heating or cooling duty** [20,44]—Reboiler, condenser and feed heater and cooler duties can be determined using shortcut or more rigorous models and
heat of vaporization data, but this measure of performance does not recognize
the impact of temperature on energy efficiency or capital—energy trade-offs.

3. **Ideal or actual power demand**—In distillation processes involving heat pumping,
such as those requiring subambient cooling, the dominant cost is that of
providing power to the refrigeration system. Eqn. (6.1) correlates ideal power
demand with the process cooling duty, the temperature at which cooling is
required and the temperature "lift" ($T_{\text{cond}} - T_{\text{evap}}$). The impact of heat integration
on power demand, where heat is rejected from a refrigeration cycle to a process
stream, can also be captured. Capital investment for the distillation and heat
pumping processes is neglected by this indicator.

4. **Operating cost**—The operating cost can be determined from the heating and
cooling duties and refrigeration power demand, using refrigeration process
models and given the cost of utilities, but it does not account for capital
investment.

5. **Total annualized cost**—Capital costs can be approximated using more or less
sophisticated approaches for capital cost estimation [5,11]. Capital cost models
are inevitably inaccurate and generally less established for novel technologies,
such as dividing wall columns. Trade-offs between capital costs and operating
costs can be taken into account by considering total annualized costs [5]. The
impact of changes in product yield or quality can be accounted for if product
revenue is also considered.

6. **Simple payback**—Payback is the ratio of annual income, profit, or savings to
capital investment and thus provides a measure of benefit to cost. It can
therefore compare processes with different throughputs or yields but cannot
reflect the scale of the investment.

### 6.6.4 Thermodynamic analysis of distillation columns

The concept of thermodynamically reversible distillation has proven useful for
analyzing and comparing distillation design alternatives [45,46]. In a reversible
distillation process, a simple column with infinitely many stages carries out a spec-
ified separation, as illustrated in Figure 6.24. The driving force is reduced to zero on
each stage by continuous addition of heat below the feed stage and continuous
removal of heat above the feed stage.

The total heat provided to a reversible distillation column is the minimum energy
demand for a specified separation at a given operating pressure. Minimum energy
demand is an important benchmark for evaluating energy efficiency of distillation col-
umns. The temperatures at which heating and cooling are required are also of interest.

#### 6.6.4.1 Reversible and near-reversible distillation profiles

Various methods have been developed to generate and interpret reversible and near-
reversible column profiles to quantify ideal heat flows to the column. These heat
flows relate to opportunities to apply side-reboilers and side-condensers at more
moderate temperatures than in the reboiler and condenser.
Reversible distillation is possible in theory for binary separations; in this case, the liquid composition profile is identical to the equilibrium curve. A stage-by-stage enthalpy balance can identify the amount of heat that needs to be added to each stage, as well as the associated temperature. Simulating the reversible column, to generate composition, temperature and enthalpy profiles, is relatively straightforward [45–47].

Real, irreversible distillation columns require more energy than thermodynamically ideal columns, because driving forces for separation and for heat transfer are non-negligible and because mixing streams of different composition or temperature generates entropy. To account for real columns, with a finite number of stages, “near-reversible” profiles can be constructed and interpreted to identify opportunities to increase energy efficiency by reducing the reflux ratio, modifying the feed condition and employing side-reboilers and side-condensers [47].

However, for multicomponent separations, only preferred splits can be carried out reversibly—in these separations, the lightest and heaviest components can be recovered to any degree to the overhead and bottom products, respectively; the remaining components distribute between the products. Most multicomponent splits of practical importance cannot be analyzed with respect to reversible operation. As the splits are not preferred splits, it is unavoidable that entropy is generated by feed-stage mixing. Several approaches have been developed to evaluate energy demand for multicomponent columns operating “as reversibly as possible” [46].

A well-established approach, known as “column targeting”, treats multicomponent separations as pseudo-binary separations to generate a temperature—enthalpy profile or “column grand composite curve” [24,47]. The procedure is straightforward,
but the validity of the analysis is severely restricted by several key assumptions: that reversible behavior is possible throughout the column, that minimum flows depend only the phase equilibrium behavior of the two key components (or groups of key components) and that molar enthalpies of the liquid and vapor streams are insensitive to composition [46]. Furthermore, the method does not compensate for the reduced reboil or reflux on some stages of a finite distillation column with a side-reboiler or side-condenser and the resulting penalty—to meet the separation specifications, either the total heating and cooling duty or the number of equilibrium stages must increase. The duties and temperatures of side-reboilers and side-condensers predicted using this temperature—enthalpy analysis are typically unrealistic with respect to both quantity and quality of heat flows [46].

An alternative approach uses the Underwood equations or rigorous models to identify the minimum vapor load in a distillation column separating a known feed [42]. The “Vmin diagram” identifies the minimum energy demand, corresponding to minimum reflux, for all separations of the feed and therefore for any particular separation. The energy demand of a reversible column is easily computed; the Vmin diagram is particularly powerful for determining the minimum energy demand of Petlyuk arrangements [20], where each subsection of the Petlyuk column carries out a preferred split.

### 6.6.4.2 Minimum driving force profiles

Other methodologies consider driving forces for separation within the column, recognizing that nonzero driving forces are needed in practice. Strictly speaking, chemical potential is the driving force for distillation, but defining an overall mass transfer driving force in terms of chemical potential is not straightforward [46]. Therefore, exergy, $Ex$, as defined in Eqn (6.3), has been proposed as a measure of irreversibility, as it accounts for both mass and heat transfer driving forces on a separation stage.

$$Ex = H - T_0S$$  \hspace{0.5cm} (6.3)

where $H$ is enthalpy and $S$ is entropy for a stream and $T_0$ is the ambient (reference) temperature.

The exergy loss of a stage is easily computed from converged simulation results, by applying an exergy balance over the stage [46].

This methodology considers design alternatives with side-reboilers and side-condensers of various duties and with additional stages, such that the total heat load is fixed and the exergy loss over any stage may not exceed a specified value. These designs can be systematically generated and evaluated with respect to exergy, operating cost, or capital—energy trade-offs [46].

### 6.6.4.3 Summary—evaluation of distillation energy requirements

To evaluate distillation design alternatives with respect to energy efficiency, quantitative models are needed to represent the distillation process and its energy requirements, considering energy quality and quantity. The validity of the evaluation
depends on the quality of the models employed and the measure of performance applied. Finding the most energy-efficient designs needs the set of design alternatives to be explored thoroughly; tools based on thermodynamic analysis of the columns can assist with this time-consuming task. Heat-integration opportunities should be considered simultaneously.

6.7 Conclusions

Distillation is an energy-intensive process, where heating and cooling needs (duty and temperature) dominate process economics, environmental impact and sustainability. A distillation column carrying out a specified separation has several degrees of freedom for design that affect energy efficiency. The column operating pressure is the most significant design variable, especially when it allows energy-intensive refrigeration to be avoided. Heat recovery between a distillation column and an associated process or refrigeration cycle can enhance energy efficiency; a challenge for distillation process design is to create and exploit opportunities for heat recovery. Well-designed distillation processes also need to be operated and controlled with energy efficiency in mind; generally, over-refluxing results in excessive energy demand and also constrains throughput.

Simple distillation columns are ubiquitous. Complex column configurations, such as dividing wall columns, are becoming more established and can offer substantial benefits with respect to energy efficiency as well as capital investment. Especially because these configurations cannot fully exploit the effects of operating pressure on energy efficiency, quantitative benefits are often diminished by qualitative penalties.

Although distillation is a mature technology, economic and environmental drivers continue to motivate further development of energy-efficient distillation processes that are not yet commercially proven, such as internally heat-integrated distillation columns. Research interest in hybrid distillation processes that can reduce energy consumption is vigorous; recent research publications address distillation—pervaporation, adsorption—distillation, liquid—liquid extraction—distillation and crystallization—distillation processes, as well as novel heat-pumped configurations and short-path distillation.

In order to design and operate energy-efficient column configurations, research in computer-aided process engineering continues to develop comprehensive, reasonably accurate and computationally robust tools for process synthesis, modeling, analysis and optimization.

References


