Chapter 7 Estimation of Capital Costs

In Chapter 1, the information provided on a process flow diagram, including a stream table and an equipment summary table, was presented. In the next four chapters, this information will be used as a basis for estimating

1. How much money (capital cost) it takes to build a new chemical plant
2. How much money (operating cost) it takes to operate a chemical plant
3. How to combine items 1 and 2 to provide several distinct types of composite values reflecting process profitability
4. How to select a "best process" from competing alternatives
5. How to estimate the economic value of making process changes and modifications to an existing processes
6. How to quantify uncertainty when evaluating the economic potential of a process

In this chapter, we concentrate on the estimation of capital costs. **Capital cost** pertains to the costs associated with construction of a new plant or modifications to an existing chemical manufacturing plant.

7.1 Classifications of Capital Cost Estimates

There are five generally accepted classifications of capital cost estimates that are most likely to be encountered in the process industries [1,2,3]:

1. Detailed estimate
2. Definitive estimate
3. Preliminary estimate
4. Study estimate
5. Order-of-magnitude estimate

The information required to perform each of these estimates is provided in Table 7.1.

Table 7.1 Summary of Capital Cost Estimating Classifications (References [1], [2], and [3])

Order-of-Magnitude (also known as Ratio or Feasibility) Estimate

Data: This type of estimate typically relies on cost information for a complete process taken from previously built plants. This cost information is then adjusted using appropriate scaling factors, for capacity, and for inflation, to provide the estimated capital cost.

Diagrams: Normally requires only a block flow diagram.

Study (also known as Major Equipment or Factored) Estimate

Data: This type of estimate utilizes a list of the major equipment found in the process. This includes all pumps, compressors and turbines, columns and vessels, fired heaters, and exchangers. Each piece of equipment is roughly sized and the approximate cost determined. The total cost of equipment is then factored to give the estimated capital cost.

Diagrams: Based on PFD as described in Chapter 1. Costs from generalized charts.

Note: Most individual student designs are in this category.

Preliminary Design (also known as Scope) Estimate

Data: This type of estimate requires more accurate sizing of equipment than used in the study estimate. In addition, approximate layout of equipment is made along with estimates of piping, instrumentation, and electrical requirements. Utilities are estimated.

Diagrams: Based on PFD as described in Chapter 1. Includes vessel sketches for major equipment, preliminary plot plan, and elevation diagram.

Note: Most large student group designs are in this category.

Definitive (also known as Project Control) Estimate

Data: This type of estimate requires preliminary specifications for all the equipment, utilities, instrumentation, electrical, and off-sites.

Diagrams: Final PFD, vessel sketches, plot plan, and elevation diagrams, utility balances, and a preliminary P&ID.

Detailed (also known as Firm or Contractor's) Estimate

Data: This type of estimate requires complete engineering of the process and all related off-sites and utilities. Vendor quotes for all expensive items will have been obtained. At the end of a detailed estimate, the plant is ready to go to the construction stage.

Diagrams: Final PFD and P&ID, vessel sketches, utility balances, plot plan and elevation diagrams, and piping isometrics. All diagrams are required to complete the construction of the plant if it is built.

The five classifications given in Table 7.1 roughly correspond to the five classes of estimate defined in the AACE Recommended Practice No. 17R-97 [4]. The accuracy range and the approximate cost for performing each class of estimate are given in Table 7.2.

Table 7.2 Classification of Cost Estimates

Class of Estimate	Level of Project Definition (as % of Complete Definition)	Typical Purpose of Estimate	Methodology (Estimating Method)	Expected Accuracy Range (+/- Range Relative to Best Index of 1)	Preparation Effort (Relative to Lowest Cost Index of 1)
Class 5	0% to 2%	Screening or Feasibility	Stochastic or Judgment	4 to 20	1
Class 4	1% to 15%	Concept Study or Feasibility	Primarily Stochastic	3 to 12	2 to 4
Class 3	10% to 40%	Budget, Authorization, or Control	Mixed but Primarily Stochastic	2 to 6	3 to 10
Class 2	30% to 70%	Control or Bid/Tender	Primarily Deterministic	1 to 3	5 to 20
Class 1	50% to 100%	Check Estimate or Bid/Tender	Deterministic	1	10 to 100

(From AACE Recommended Practice No. 17R-97 [4], reprinted with permission of AACE International, 209 Prairie Ave., Morgantown, WV; http://www.aacei.org)

In Table 7.2, the accuracy range associated with each class of estimate and the costs associated with carrying out the estimate are ranked relative to the most accurate class of estimate (Class 1). In order to use the information in Table 7.2, it is necessary to know the accuracy of a Class 1 estimate. For the cost estimation of a chemical plant, a Class 1 estimate (detailed estimate) is typically +6% to –4% accurate. This means that by doing such an estimate, the true cost of building the plant would likely be in the range of 6% higher than and 4% lower than the estimated price. Likewise, the effort to prepare a Class 5 estimate for a chemical process is typically in the range of 0.015% to 0.30% of the total installed cost of the plant [1,2].

The use of the information in Table 7.2, to estimate the accuracy and costs of performing estimates, is illustrated in Examples 7.1 and 7.2.

Example 7.1

The estimated capital cost for a chemical plant using the study estimate method (Class 4) was calculated to be \$2 million. If the plant were to be built, over what range would you expect the actual capital estimate to vary?

For a Class 4 estimate, from Table 7.2, the expected accuracy range is between 3 and 12 times that of a Class 1 estimate. As noted in the text, a Class 1 estimate can be expected to vary from +6% to –4%. We can evaluate the narrowest and broadest expected capital cost ranges as follows.

Lowest Expected Cost Range

High value for actual plant cost $(\$2.0 \times 10^6)[1 + (0.06)(3)] = \$2.36 \times 10^6$

Low value for actual plant cost $(\$2.0 \times 10^6)[1 - (0.04)(3)] = \$1.76 \times 10^6$

Highest Expected Cost Range

High value for actual plant cost $(\$2.0 \times 10^6)[1 + (0.06)(12)] = \$3.44 \times 10^6$

Low value for actual plant cost $(\$2.0 \times 10^6)[1 - (0.04)(12)] = \$1.04 \times 10^6$

The actual expected range would depend on the level of project definition and effort. If the effort and definition are at the high end, then the expected cost range would be between \$1.76 and \$2.36 million. If the effort and definition are at the low end, then the expected cost range would be between \$1.04 and \$3.44 million.

The primary reason that capital costs are underestimated stems from the failure to include all of the equipment needed in the process. Typically, as a design progresses, the need for additional equipment is uncovered, and the estimate accuracy improves. The different ranges of cost estimates are illustrated in Example 7.2.

Example 7.2

Compare the costs for performing an order-of-magnitude estimate and a detailed estimate for a plant that cost $\$5.0 \times 10^6$ to build.

For the order-of-magnitude estimate, the cost of the estimate is in the range of 0.015% to 0.3% of the final cost of the plant:

Highest Expected Value: $(\$5.0 \times 10^6)(0.003) = \$15{,}000$
Lowest Expected Value: $(\$5.0 \times 10^6)(0.00015) = \$750$

For the detailed estimate, the cost of the estimate is in the range of 10 to 100 times that of the order-of-magnitude estimate.

For the lowest expected cost range:

Highest Expected Value: $(\$5.0 \times 10^6)(0.03) = \$150{,}000$
Lowest Expected Value: $(\$5.0 \times 10^6)(0.0015) = \$7500$

For the highest expected cost range:

Highest Expected Value: $(\$5.0 \times 10^6)(0.3) = \$1{,}500{,}000$
Lowest Expected Value: $(\$5.0 \times 10^6)(0.015) = \$75{,}000$

Capital cost estimates are essentially paper-and-pencil studies. The cost of making an estimate indicates the personnel hours required in order to complete the estimate. From Table 7.2 and Examples 7.1 and 7.2, the trend between the accuracy of an estimate and the cost of the estimate is clear. If greater accuracy is required in the capital cost estimate, then more time and money must be expended in conducting the estimate. This is the direct result of the greater detail required for the more accurate estimating techniques.

What cost estimation technique is appropriate? At the beginning of Chapter 1, a short narrative was given that introduced the evolution of a chemical process leading to the final design and construction of a chemical plant. Cost estimates are performed at each stage of this evolution.

There are many tens to hundreds of process systems examined at the block diagram level for each process that makes it to the construction stage. Most of the processes initially considered are screened out before any detailed cost estimates are made. Two major areas dominate this screening process. To continue process development, the process must be both technically sound and economically attractive.

A typical series of cost estimates that would be carried out in the narrative presented in Chapter 1 is as follows.

- Preliminary feasibility estimates (order-of-magnitude or study estimates) are made to compare many process alternatives.
- More accurate estimates (preliminary or definitive estimates) are made for the most profitable processes identified in the feasibility study.
- Detailed estimates are then made for the more promising alternatives that remain after the preliminary estimates.
- Based on the results from the detailed estimate, a final decision is made whether to go ahead with the construction of a plant.

This text focuses on the preliminary and study estimation classification based on a PFD as presented in Chapter 1. This approach will provide estimates accurate in the range of +40% to –25%.

In this chapter, it is assumed that all processes considered are technically sound and attention is focused on the economic estimation of capital costs. The technical aspects of processes will be considered in later chapters.

7.2 Estimation of Purchased Equipment Costs

To obtain an estimate of the capital cost of a chemical plant, the costs associated with major plant equipment must be known. For the presentation in this chapter, it is assumed that a PFD for the process is available. This PFD is similar to the one discussed in detail in Chapter 1, which included material and energy balances with each major piece of equipment identified, materials of construction selected, and the size/capacity roughly estimated from conditions on the PFD. Additional PFDs and equipment summary tables are given for several processes in Appendix B.

The most accurate estimate of the purchased cost of a piece of major equipment is provided by a current price quote from a suitable vendor (a seller of equipment). The next best alternative is to use cost data on previously purchased equipment of the same type. Another technique, sufficiently accurate for study and preliminary cost estimates, utilizes summary graphs available for various types of common equipment. This last technique is used for study estimates emphasized in this text and is discussed in detail in Section 7.3. Any cost data must be adjusted for any difference in unit capacity (see Section 7.2.1) and also for any elapsed time since the cost data were generated (see Section 7.2.2).

7.2.1 Effect of Capacity on Purchased Equipment Cost

The most common simple relationship between the purchased cost and an attribute of the equipment related to units of capacity is given by Equation 7.1.

(7.1)

$$\frac{C_a}{C_b} = \left(\frac{A_a}{A_b}\right)^n$$

where A = Equipment cost attribute

C = Purchased cost

n = Cost exponent

Subscripts: a refers to equipment with the required attribute

b refers to equipment with the base attribute

The **equipment cost attribute** is the equipment parameter that is used to correlate capital costs. The equipment cost attribute is most often related to the unit capacity, and the term *capacity* is commonly used to describe and identify this attribute. Some typical values of cost exponents and unit capacities are given in Table 7.3. From Table 7.3, it can be seen that the following information is given:

Table 7.3 Typical Values of Cost Exponents for a Selection of Process Equipment

Range of Equipment Type	Correlation	Units of Capacity	Cost Exponent n
Reciprocating compressor with motor drive	0.75 to 1490	kW	0.84
Heat exchanger shell and tube carbon steel	1.9 to 1860	m^2	0.59
Vertical tank carbon steel	0.4 to 76	m^3	0.30
Centrifugal blower	0.24 – 71	std m^3/s	0.60
Jacketed kettle glass lined	0.2 to 3.8	m^3	0.48

(All data from Table 9-50, *Chemical Engineer's Handbook*, Perry, R.H., Green, D.W., and Maloney, J.O. (eds.), 7th ed, 1997. Reproduced by permission of The McGraw-Hill Companies, Inc., New York, NY.)

1. A description of the type of equipment used
2. The units in which the capacity is measured
3. The range of capacity over which the correlation is valid
4. The cost exponent (values shown for n vary between 0.30 and 0.84)

Equation 7.1 can be rearranged to give

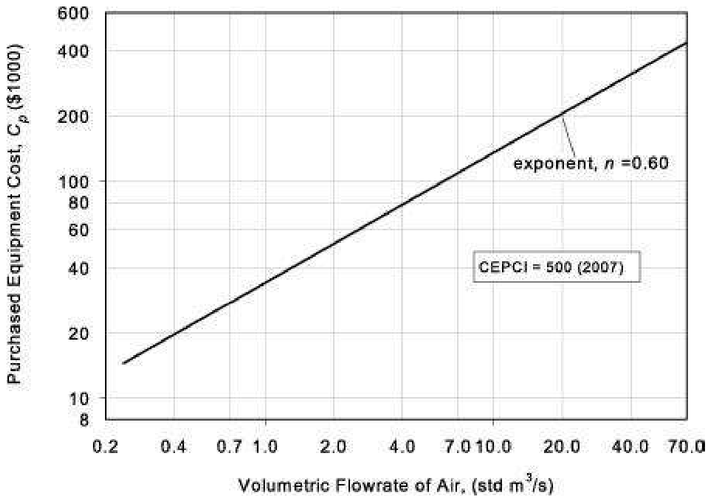
(7.2)

$$C_a = KA_a^n$$

where $K = C_b/A_b^n$

Equation 7.2 is a straight line with a slope of n when the log of C_a is plotted versus the log of A_a. To illustrate this relationship, the typical cost of a single-stage blower versus the capacity of the blower, given as the volumetric flowrate, is plotted in Figure 7.1. The value for the cost exponent, n, from this curve is 0.60.

Figure 7.1 Purchased Cost of a Centrifugal Air Blower (Data adapted from Reference [3])



The value of the cost exponent, n, used in Equations 7.1 and 7.2, varies depending on the class of equipment being represented. See Table 7.3. The value of n for different items of equipment is often around 0.6. Replacing n in Equation 7.1 and/or 5.2 by 0.6 provides the relationship referred to as the **six-tenths rule.** A problem using the six-tenths rule is given in Example 7.3.

Example 7.3

Use the six-tenths rule to estimate the percentage increase in purchased cost when the capacity of a piece of equipment is doubled.

Using Equation 7.1 with n = 0.6,

$$C_a/C_b = (2/1)^{0.6} = 1.52$$

$$\%\ \text{increase} = ((1.52 - 1.00)/1.00)(100) = 52\%$$

This simple example illustrates a concept referred to as the **economy of scale**. Even though the equipment capacity was doubled, the purchased cost of the equipment increased by only 52%. This leads to the following generalization.

The larger the equipment, the lower the cost of equipment per unit of capacity.

Special care must be taken in using the six-tenths rule for a single piece of equipment. The cost exponent may vary considerably from 0.6, as illustrated in Example 7.4. The use of this rule for a total chemical process is more reliable and is discussed in Section 7.3.

Example 7.4

Compare the error for the scale-up of a reciprocating compressor by a factor of 5 using the six-tenths rule in place of the cost exponent given in Table 7.3.

Using Equation 7.1,

Cost ratio using six-tenths rule (i.e., $n = 0.60$) = $5.0^{0.60} = 2.63$

Cost ratio using ($n = 0.84$) from Table 7.3 = $5.0^{0.84} = 3.86$

% Error = ((2.63 – 3.86)/3.86)(100) = –32 %

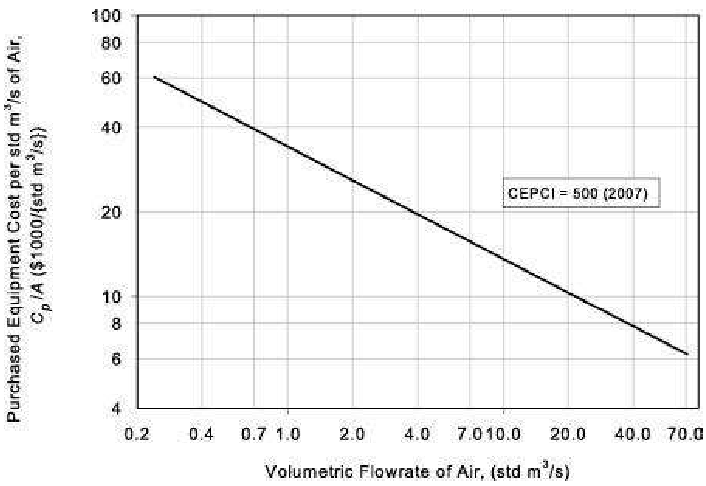
Another way to think of the economy of scale is to consider the purchased cost of equipment per unit capacity. Equation 7.2 can be rearranged to give the following relationship:

(7.3)

$$\frac{C}{A} = KA^{n-1}$$

If Equation 7.3 is plotted on log-log coordinates, the resulting curve will have a negative slope, as shown in Figure 7.2. The meaning of the negative slope is that as the capacity of a piece of equipment increases, the cost per unit of capacity decreases. This, of course, is a consequence of $n < 1$ but also shows clearly how the economy of scale works. As cost curves for equipment are introduced in the text, they will be presented in terms of cost per unit capacity as a function of capacity to illustrate better the idea of economy of scale. For many equipment types, the simple relationship in Equation 7.1 is not very accurate, and an equation that is second order in the attribute is used.

Figure 7.2 Purchased Cost per Unit of Flowrate of a Centrifugal Air Blower (Adapted from Reference [3])



In the last two examples, the relative costs of equipment of differing size were calculated. It is necessary to have cost information on the equipment at some "base case" in order to be able to determine the cost of other similar equipment. This base-case information must allow for the constant, K, in Equation 7.2, to be evaluated, as shown in Example 7.5. This base case cost information may be obtained from a current bid provided by a manufacturer for the needed equipment or from company records of prices paid for similar equipment.

Example 7.5

The purchased cost of a recently acquired heat exchanger with an area of 100 m^2 was $10,000.

Determine

- **a.** The constant K in Equation 7.2
- **b.** The cost of a new heat exchanger with area equal to 180 m^2

From Table 7.3: $n = 0.59$: for Equation 7.2:

- **a.** $K = C_b/(A_b)^n = 10{,}000/(100)^{0.59} = 661\ \{\$/(m^2)^{0.59}\}$
- **b.** $C_a = (661)(180)^{0.59} = \$14{,}100$

There are additional techniques that allow for the price of equipment to be estimated that do not require information from either of the sources given above. One of these techniques is discussed in Section 7.3.

7.2.2 Effect of Time on Purchased Equipment Cost

In Figures 7.1 and 7.2, the time at which the cost data were reported (2006) is given on the figure. This raises the question of how to convert this cost into one that is accurate for the present time. When one depends on past records or published correlations for price information, it is essential to be able to update these costs to take changing economic conditions (inflation) into account. This can be achieved by using the following expression:

(7.4)

$$C_2 = C_1 \left(\frac{I_2}{I_1} \right)$$

where C = Purchased cost

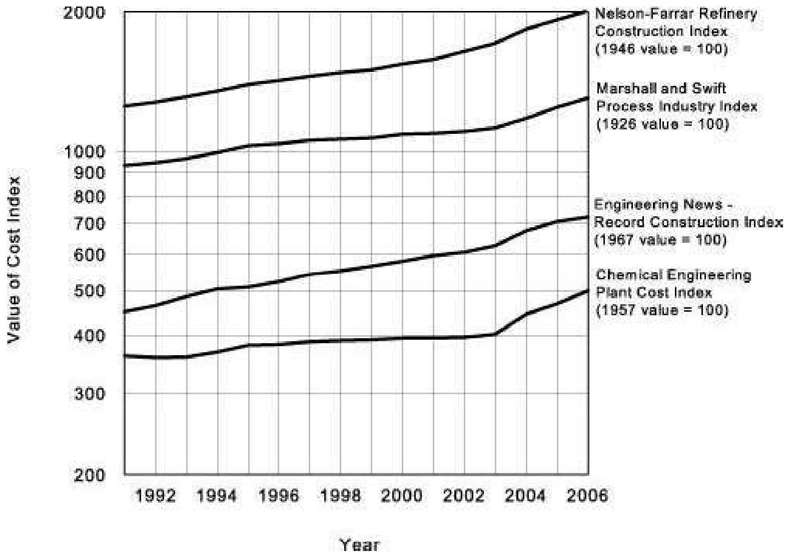
I = Cost index

Subscripts: 1 refers to base time when cost is known

2 refers to time when cost is desired

There are several cost indices used by the chemical industry to adjust for the effects of inflation. Several of these cost indices are plotted in Figure 7.3.

Figure 7.3 The Variations in Several Commonly Used Cost Indexes Over 15 Years (1992–2006)



All indices in Figure 7.3 show similar inflationary trends with time. The indices most generally accepted in the chemical industry and reported in the back page of every issue of *Chemical Engineering* are the Marshall and Swift Equipment Cost Index and the Chemical Engineering Plant Cost Index.

Table 7.4 provides values for both the Marshall and Swift Equipment Cost Index and the Chemical Engineering Plant Cost Index from 1991 to 2006.

Table 7.4 Values for the Chemical Engineering Plant Cost Index and the Marshall and Swift Equipment Cost Index from 1991 to 2006

Year	Marshall and Swift Equipment Cost Index	Chemical Engineering Plant Cost Index
1991	931	361
1992	943	358
1993	964	359
1994	993	368
1995	1028	381
1996	1039	382
1997	1057	387
1998	1062	390
1999	1068	391
2000	1089	394
2001	1094	394
2002	1104	396
2003	1124	402
2004	1179	444
2005	1245	468
2006	1302	500

Unless otherwise stated, the Chemical Engineering Plant Cost Index (CEPCI) will be used in this text to account for inflation. This is a composite index, and the items that are included in the index are listed in Table 7.5. A comparison between these two indices is given in Example 7.6.

Talbe 7.5 The Basis for the Chemical Engineering Plant Cost Index

Components of Index	Weighting of Component (%)	
Equipment, Machinery, and Supports		
(a) Fabricated equipment	37	
(b) Process machinery	14	
(c) Pipe, valves, and fittings	20	
(d) Process instruments and controls	7	
(e) Pumps and compressors	7	
(f) Electrical equipment and materials	5	
(g) Structural supports, insulation, and paint	10	
	100	61% of total
Erection and installation labor		22
Buildings, materials, and labor		7
Engineering and supervision		10
Total		100

Example 7.6

The purchased cost of a heat exchanger of 500 m^2 area in 1992 was $25,000.

- **a.** Estimate the cost of the same heat exchanger in 2006 using the two indices introduced above.
- **b.** Compare the results.

From Table 7.4	**1992**	**2006**
Marshal and Swift Index	943	1302

Chemical Engineering Plant Cost Index	358	500

a. Marshal and Swift: Cost = ($25,000)(1302/943) = $34,518
Chemical Engineering:
Cost = ($25,000)(500/358) = $34,916

b. Average Difference: (($34,518 – 34,916)/(($34,518 + 34,916)/2)(100) = –1.1%

7.3 Estimating the Total Capital Cost of a Plant

The capital cost for a chemical plant must take into consideration many costs other than the purchased cost of the equipment. As an analogy, consider the costs associated with building a new home.

> *The purchased cost of all the materials that are needed to build a home does not represent the cost of the home. The final cost reflects the cost of property, the cost for delivering materials, the cost of construction, the cost of a driveway, the cost for hooking up utilities, and so on.*

Table 7.6 presents a summary of the costs that must be considered in the evaluation of the total capital cost of a chemical plant.

Talbe 7.6 Factors Affecting the Costs Associated with Evaluation of Capital Cost of Chemical Plants (from References [2] and [5])

Factor Associated with the Installation of Equipment	Symbol	Comments
1. **Direct Project Expenses**		
a. Equipment f.o.b. cost (f.o.b. = free on board)	C_P	Purchased cost of equipment at manufacturer's site.
b. Materials required for installation	C_M	Includes all piping, insulation and fireproofing, foundations and structural supports, instrumentation and electrical, and painting associated with the equipment.
c. Labor to install equipment and material	C_L	Includes all labor associated with installing the equipment and materials mentioned in (a) and (b).
2. **Indirect Project Expenses**		
a. Freight, insurance, and taxes	C_{FIT}	Includes all transportation costs for shipping equipment and materials to the plant site, all insurance on the items shipped, and any purchase taxes that may be applicable.
b. Construction overhead	C_O	Includes all fringe benefits such as vacation, sick leave, retirement benefits, etc.; labor burden such as social security and unemployment insurance, etc.; and salaries and overhead for supervisory personnel.
c. Contractor engineering expenses	C_E	Includes salaries and overhead for the engineering, drafting, and project management personnel on the project.
3. **Contingency and Fee**		
a. Contingency	C_{Cont}	A factor to cover unforeseen circumstances. These may include loss of time due to storms and strikes, small changes in the design, and unpredicted price increases.
b. Contractor fee	C_{Fee}	This fee varies depending on the type of plant and a variety of other factors.
4. **Auxiliary Facilities**		
a. Site development	C_{Site}	Includes the purchase of land; grading and excavation of the site; installation and hookup of electrical, water, and sewer systems; and construction of all internal roads, walkways, and parking lots.
b. Auxiliary buildings	C_{Aux}	Includes administration offices, maintenance shop and control rooms, warehouses, and service buildings (e.g., cafeteria, dressing rooms, and medical facility).
c. Off-sites and utilities	C_{Off}	Includes raw material and final product storage; raw material and final product loading and unloading facilities; all equipment necessary to supply required process utilities (e.g., cooling water, steam generation, fuel distribution systems, etc.); central environmental control facilities (e.g., waste water treatment, incinerators, flares, etc.); and fire protection systems.

The estimating procedures to obtain the full capital cost of the plant are described in this section. If an estimate of the capital cost for a process plant is needed and access to a previous estimate for a similar plant with a different capacity is available, then the principles already introduced for the scaling of purchased costs of equipment can be used.

1. The six-tenths rule (Equation 7.1 with *n* set to 0.6) can be used to scale up or down to a new capacity.
2. The Chemical Engineering Plant Cost Index should be used to update the capital costs (Equation 7.4).

The six-tenths rule is more accurate in this application than it is for estimating the cost of a single piece of equipment. The increased accuracy results from the fact that multiple units are required in a processing plant. Some of the process units will have cost coefficients, *n*, less than 0.6. For this equipment the six-tenths rule overestimates the costs of these units. In a similar way, costs for process units having coefficients greater than 0.6 are underestimated. When the sum of the costs is determined, these differences tend to cancel each other out.

The Chemical Engineering Plant Cost Index (CEPCI) can be used to account for changes that result from inflation. The CEPCI values provided in Table 7.4 are composite values that reflect the inflation of a mix of goods and services associated with the chemical process industries (CPI).

You may be familiar with the more common consumer price index issued by the government. This represents a composite cost index that reflects the effect of inflation on the cost of living. This index considers the changing cost of a "basket" of goods composed of items used by the "average" person. For example, the price of housing, cost of basic foods, cost of clothes and transportation, and so on, are included and weighted appropriately to give a single number reflecting the average cost of these goods. By comparing this number over time, it is possible to get an indication of the rate of inflation as it affects the average person.

In a similar manner, the CEPCI represents a "basket" of items directly related to the costs associated with the construction of chemical plants. A breakdown of the items included in this index was given in Table 7.5. The index is directly related to the effect of inflation on the cost of an "average" chemical plant, as shown in Example 7.7.

Example 7.7

The capital cost of a 30,000 metric ton/year isopropanol plant in 1992 was estimated to be \$23 million. Estimate the capital cost of a new plant with a production rate of 50,000 metric tons/year in 2007 (assume CEPCI = 500).

Cost in 2007 = (Cost in 1992)(Capacity Correction)(Inflation Correction)

$$= (\$23{,}000{,}000)(50{,}000/30{,}000)^{0.6}(500/358)$$

= ($23,000,000)(1.359)(1.397) = $43,644,000

In most situations, cost information will not be available for the same process configuration; therefore, other estimating techniques must be used.

7.3.1 Lang Factor Technique

A simple technique to estimate the capital cost of a chemical plant is the Lang Factor method, due to Lang [6, 7, 8]. The cost determined from the Lang Factor represents the cost to build a major expansion to an existing chemical plant. The total cost is determined by multiplying the total purchased cost for all the major items of equipment by a constant. The major items of equipment are those shown in the process flow diagram. The constant multiplier is called the Lang Factor. Values for Lang Factors, F_{Lang}, are given in Table 7.7.

Talbe 7.7 Lang Factors for the Estimation of Capital Cost for Chemical Plant (from References [6, 7, 8])

Capital Cost = (Lang Factor) (Sum of Purchased Costs of All Major Equipment)	
Type of Chemical Plant	**Lang Factor = F_{Lang}**
Fluid processing plant	4.74
Solid-fluid processing plant	3.63
Solid processing plant	3.10

The capital cost calculation is determined using Equation 7.5.

(7.5)

$$C_{TM} = F_{Lang} \sum_{i=1}^{n} C_{p,i}$$

where C_{TM} is the capital cost (total module) of the plant

$C_{p,i}$ is the purchased cost for the major equipment units

n is the total number of individual units

F_{Lang} is the Lang Factor (from Table 7.7)

Plants processing only fluids have the largest Lang Factor, 4.74, and plants processing only solids have a factor of 3.10. Combination fluid-solid systems fall between these two values. The greater the Lang Factor, the less the purchased costs contribute to the plant costs. For all cases, the purchased cost of the equipment is less than one-third of the capital cost of the plant. The use of the Lang Factor is illustrated in

Example 7.8.

Example 7.8

Determine the capital cost for a major expansion to a fluid processing plant that has a total purchased equipment cost of $6,800,000.

Capital Costs = ($6,800,000)(4.74) = $32,232,000

This estimating technique is insensitive to changes in process configuration, especially between processes in the same broad categories shown in Table 7.7. It cannot accurately account for the common problems of special materials of construction and high operating pressures. A number of alternative techniques are available. All require more detailed calculations using specific price information for the individual units/equipment.

7.3.2 Module Costing Technique

The equipment module costing technique is a common technique to estimate the cost of a new chemical plant. It is generally accepted as the best for making preliminary cost estimates and is used extensively in this text. This approach, introduced by Guthrie [9, 10] in the late 1960s and early 1970s, forms the basis of many of the equipment module techniques in use today. This costing technique relates all costs back to the purchased cost of equipment evaluated for some base conditions. Deviations from these base conditions are handled by using multiplying factors that depend on the following:

1. The specific equipment type
2. The specific system pressure
3. The specific materials of construction

The material provided in the next section is based upon information in Guthrie [9, 10], Ulrich [5], and Navarrete [11]. The reader is encouraged to review these references for further information.

Equation 7.6 is used to calculate the bare module cost for each piece of equipment. The bare module cost is the sum of the direct and indirect costs shown in Table 7.6.

(7.6)

$$C_{BM} = C_p^o F_{BM}$$

where C_{BM} = bare module equipment cost: direct and indirect costs for each unit

F_{BM} = bare module cost factor: multiplication factor to account for the items in Table 7.6 plus the specific materials of construction and operating pressure

C_p^o = purchased cost for base conditions: equipment made of the most common material, usually carbon steel and operating at near ambient pressures

Because of the importance of this cost estimating technique, it is described below in detail.

7.3.3 Bare Module Cost for Equipment at Base Conditions

The bare module equipment cost represents the sum of direct and indirect costs shown in Table 7.6. The conditions specified for the base case are

1. Unit fabricated from most common material, usually carbon steel (CS)
2. Unit operated at near-ambient pressure

Equation 7.6 is used to obtain the bare module cost for the base conditions. For these base conditions, a superscript zero (0) is added to the bare module cost factor and the bare module equipment cost. Thus C^o_{BM} and F^o_{BM} refer to the base conditions.

Table 7.8 supplements Table 7.6 and provides the relationships and equations for the direct, indirect, contingency, and fee costs based on the purchased cost of the equipment. These equations are used to evaluate the bare module factor. The entries in Table 7.8 are described on page 202.

Talbe 7.8 Equations for Evaluating Direct, Indirect, Contingency, and Fee Costs

Factor	Basic Equation	Multiplying Factor to Be Used with Purchased Cost, C^o_p
1. Direct		
a. Equipment	$C^o_p = C^o_p$	1.0
b. Materials	$C_M = \alpha_M C^o_p$	α_M
c. Labor	$C_L = \alpha_L(C^o_p + C_M)$	$(1.0 + \alpha_M)\alpha_L$
Total Direct	$C_{DE} = C^o_p + C_M + C_L$	$(1.0 + \alpha_M)(1.0 + \alpha_L)$
2. Indirect		
a. Freight, etc.	$C_{FIT} = \alpha_{FIT}(C^o_p + C_M)$	$(1.0 + \alpha_M)\alpha_{FIT}$
b. Overhead	$C_O = \alpha_O C_L$	$(1.0 + \alpha_M)\alpha_L\alpha_O$
c. Engineering	$C_E = \alpha_E(C^o_p + C_M)$	$(1.0 + \alpha_M)\alpha_E$
Total Indirect	$C_{IDE} = C_{FIT} + C_O + C_E$	$(1.0 + \alpha_M)(\alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)$
Bare Module	$C^o_{BM} = C_{IDE} + C_{DE}$	$(1.0 + \alpha_M)(1.0 + \alpha_L + \alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)$
3. Contingency and Fee		
a. Contingency	$C_{Cont} = \alpha_{Cont} C^o_{BM}$	$(1.0 + \alpha_M)(1.0 + \alpha_L + \alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)\alpha_{Cont}$
b. Fee	$C_{Fee} = \alpha_{Fee} C^o_{BM}$	$(1.0 + \alpha_M)(1.0 + \alpha_L + \alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)\alpha_{Fee}$
Total Module	$C_{TM} = C^o_{BM} + C_{Cont} + C_{Fee}$	$(1.0 + \alpha_M)(1.0 + \alpha_L + \alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)(1.0 + \alpha_{Cont} + \alpha_{Fee})$

Column 1: Lists the factors given in Table 7.6.

Column 2: Lists equations used to evaluate each of the costs. These equations introduce multiplication cost factors, α_i. Each cost item, other than the purchased equipment cost, introduces a separate factor.

Column 3: For each factor, the cost is related to the purchased cost C^o_p by an equation of the form.

(7.7)

$$C_{XX} = C_p^o f(\alpha_{i,j,k...})$$

The function, $f(\alpha_{i,\, j,\, k...})$, is given in column 3 of Table 5.8.

From Table 7.8 and Equations 7.6 and 7.7, it can be seen that the bare module factor is given by

(7.8)

$$F_{BM}^o = [1 + \alpha_L + \alpha_{FIT} + \alpha_L \alpha_o + \alpha_E][1 + \alpha_M]$$

The values for the bare module cost multiplying factors vary between equipment modules. The calculations for the bare module factor and bare module cost for a carbon steel heat exchanger are given in Example 7.9.

Example 7.9

The purchased cost for a carbon steel heat exchanger operating at ambient pressure is $10,000. For a heat exchanger module, Guthrie [9, 10] provides the following cost information.

Item	% of Purchased Equipment Cost
Equipment	100.0
Materials	71.4
Labor	63.0
Freight	8.0
Overhead	63.4
Engineering	23.3

Using the information given above, determine the equivalent cost multipliers given in Table 7.8 and the following:

a. Bare module cost factor, F_{BM}^o
b. Bare module cost, C_{BM}^o

Item	% of Purchased Equipment Cost	Cost Multiplier (Table 7.8)	Value of Multiplier
Equipment	100.0	1.0	
Materials	71.4	α_M	0.714
Labor	63.0	α_L	0.63/(1 + 0.714) = 0.368
Freight	8.0	α_{FIT}	0.08/(1 + 0.714) = 0.047
Overhead	63.4	α_O	0.634/0.368/(1 + 0.714) = 1.005
Engineering	23.3	α_E	0.233/(1 + 0.714) = 0.136
Bare Module	**329.1**		

a. Using Equation 7.8,

$$F_{BM}^o = (1 + 0.368 + 0.047 + (1.005)(0.368) + 0.136)(1 + 0.714) = 3.291$$

b. From Equation 7.6,

$$C_{BM}^o = (3.291)(\$10{,}000) = \$32{,}910$$

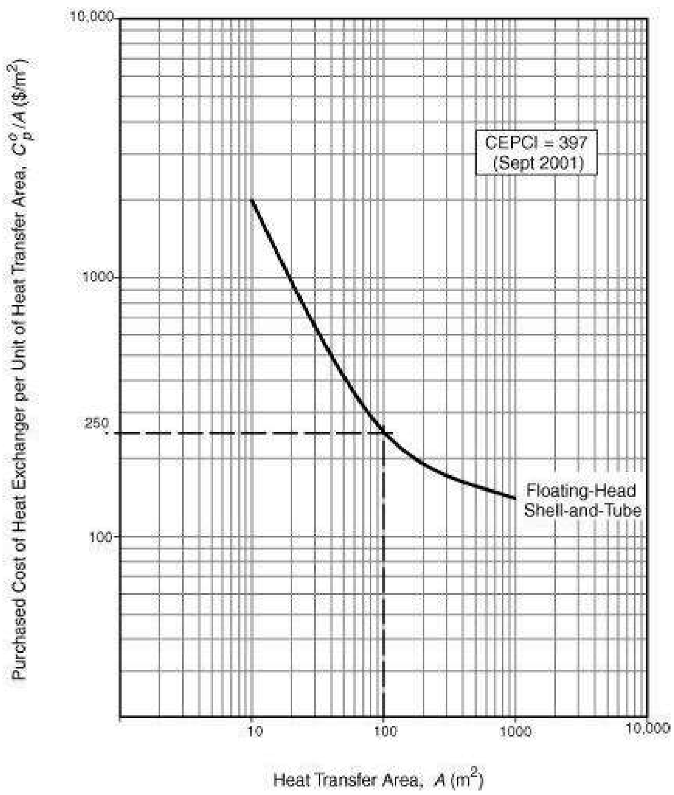
Fortunately, we do not have to repeat the procedure illustrated in Example 7.9 in order to estimate F^o_{BM} for every piece of equipment. This has already been done for a large number of equipment modules, and the results are given in Appendix A.

In order to estimate bare module costs for equipment, purchased costs for the equipment at base case conditions (ambient pressure using carbon steel) must be available along with the corresponding bare module factor and factors to account for different operating pressures and materials of construction. These data are made available for a variety of common gas/liquid processing equipment in Appendix A. These data were compiled during the summer of 2001 from information obtained from manufacturers and also from the R-Books software marketed by Richardson Engineering Services [12]. The method by which material and pressure factors are accounted for depends on the equipment type, and these are covered in the next section. The estimation of the bare module cost for a floating-head shell-and-tube heat exchanger is illustrated in Example 7.10 and in subsequent examples in this chapter.

Example 7.10

Find the bare module cost of a floating-head shell-and-tube heat exchanger with a heat transfer area of 100 m^2 at the end of 2006. The operating pressure of the equipment is 1.0 bar, with both shell-and-tube sides constructed of carbon steel. The cost curve for this heat exchanger is given in Appendix A, Figure A.5, and is repeated as Figure 7.4. It should be noted that unlike the examples shown in Figures 7.1 and 7.2, the log-log plot of cost per unit area versus area is nonlinear. In general this will be the case, and a second order polynomial is normally used to describe this relationship.

Figure 7.4 Purchased Costs for Floating-Head Shell-and-Tube Heat Exchangers



From Figure 7.4, C_p^o(2001) = ($ 250)(100) = $25,000 (the evaluation path is shown on Figure 7.4).

The bare module cost for shell-and-tube heat exchangers is given by Equation A.4.

(A.4)

$$C_{BM} = C_p^o\,[B_1 + B_2 F_p F_M]$$

The values of B_1 and B_2 for floating-head heat exchangers from Table A.4 are 1.63 and 1.66, respectively.

The pressure factor is obtained from Equation A.3.

(A.3)

$$\log_{10} F_p = C_1 + C_2 \log_{10} P + C_3 (\log_{10} P)^2$$

From Table A.2, for pressures <5 barg, $C_1 = C_2 = C_3 = 0$, and from Equation A.3, $F_p = 1$. Using data in Table A.3 for shell-and-tube heat exchangers with both shell and tubes made of carbon steel (Identification Number = 1) and Figure A.8, $F_M = 1$. Substituting this data into Equation A.4 gives

$$C^o_{BM}(2001) = C^o_p(2001)[1.63 + 1.66(F_p = 1)(F_M = 1)] = 3.29C^o_p = (3.29)(\$25{,}000) = \$82{,}300$$
$$C^o_{BM}(2006) = C^o_{BM}(2001)\,(500/394) = \$82{,}300\,(500/397) = \$103{,}590$$

A comparison of the value of bare module cost factor for Example 7.10 shows that it is the same as the value of 3.29 evaluated using the individual values for α_i, given in Example 7.9.

7.3.4 Bare Module Cost for Nonbase Case Conditions

For equipment made from other materials of construction and/or operating at nonambient pressure, the values for F_M and F_P are greater than 1.0. In the equipment module technique, these additional costs are incorporated into the bare module cost factor, F_{BM}. The bare module factor used for the base case, F^o_{BM}, is replaced with an actual bare module cost factor, F_{BM}, in Equation 7.6. The information needed to determine this actual bare module factor is provided in Appendix A. The effect of pressure on the cost of equipment is considered first.

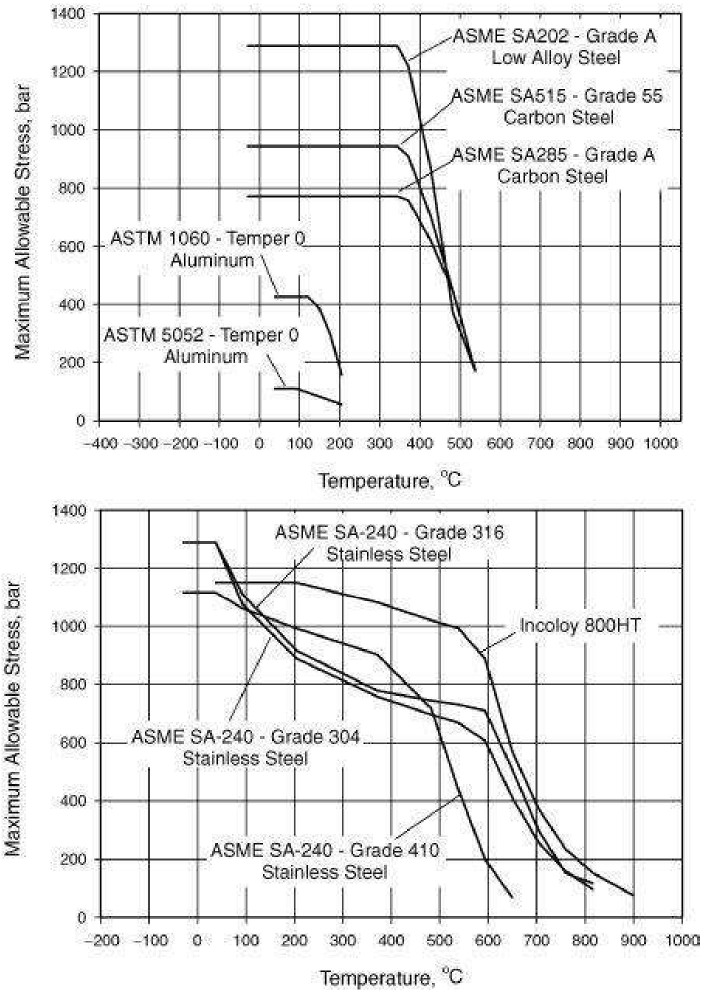
Pressure Factors. As the pressure at which a piece of equipment operates increases, the thickness of the walls of the equipment will also increase. For example, consider the design of a process vessel. Such vessels, when subjected to internal pressure (or external pressure when operating at vacuum) are subject to rigorous mechanical design procedures. For the simple case of a cylindrical vessel operating at greater than ambient pressure, the relationship between design pressure and wall thickness required to withstand the radial stress in the cylindrical portion of the vessel, as recommended by the ASME [13], is given as

(7.9)

$$t = \frac{PD}{2SE - 1.2P} + CA$$

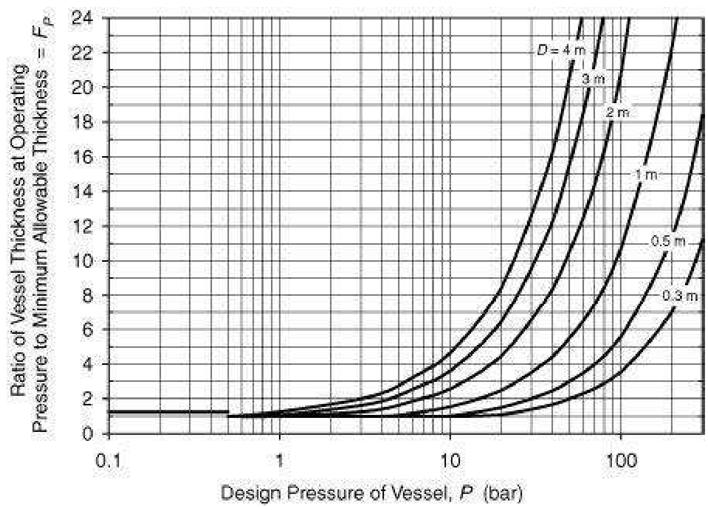
where t is the wall thickness in meters, P is the design pressure (bar), D is the diameter of the vessel (m), S is the maximum allowable working pressure (maximum allowable stress) of material (bar), E is a weld efficiency, and CA is the corrosion allowance (m). The weld efficiency is dependent on the type of weld and the degree of examination of the weld. Typical values are from 1.0 to 0.6. The corrosion allowance depends on the service, and typical values are from 3.15 to 6.3 mm (0.125 to 0.25 inches). However, for very aggressive environments, inert linings such as glass and graphite are often used to protect the structural metal. Finally, the maximum working pressure of the material of construction, S, is dependent not only on the material but also on the operating temperature. Some typical values of S are given for common materials of construction in Figure 7.5. From this figure, it is clear that for typical carbon steel the maximum allowable stress drops off rapidly after 350°C. However, for stainless steels (ASME SA-240) the decrease in maximum allowable stress with temperature is less steep, and operation up to 600–650°C is possible for some grades. For even higher temperatures and very corrosive environments, when the lining of vessels is not practical, more exotic alloys such as titanium and titanium-based alloys and nickel-based alloys may be used. For example, Hastelloy B has excellent resistance to alkali environments up to 850°C. Inconel 600, whose main constituents are Ni 72%, Cr 15%, and Fe 8%, has excellent corrosion resistance to oxidizing environments such as acids and can be used from cryogenic temperatures up to 1100°C. The maximum allowable working pressure for Incoloy 800HT, which also has excellent corrosion resistance in acidic environments, is shown as a function of temperature in Figure 7.5.

Figure 7.5 Maximum Allowable Stresses for Materials of Construction as a Function of Operating Temperature (Data from Perry et al. [3], Chapter 10, and Ref [15])



The relationship between cost of a vessel and its operating pressure is a complex one. However, with all other things being constant, the cost of the vessel is approximately proportional to the weight of the vessel, which in turn is proportional to the vessel thickness. From Equation 7.9, it is clear that as the operating pressure approaches 1.67*SE,* the required wall thickness, and hence cost, becomes infinite. Moreover, the thickness of the vessel for a given pressure will increase as the vessel diameter increases. The effect of pressure on the weight (and ultimately cost) of carbon steel vessel shells as a function of vessel diameter is shown in Figure 7.6. The *y*-axis of the figure shows the ratio of the vessel thickness at the design pressure to that at ambient pressure, and the *x*-axis is the design pressure. A corrosion allowance of 3.15 mm (1/8 inch) and a value of S = 944 bar (13,700 psi) are assumed. It is also assumed that the vessel is designed with a minimum wall thickness of 6.3 mm (1/4 inch). A minimum wall thickness is often required to ensure that the vessel does not buckle under its own weight or when being transported. In addition to these factors, the costs for the vessel supports, manholes, nozzles, instrument wells, the vessel head, and so on, all add to the overall weight and cost of the vessel. For the sake of simplification, it is assumed that the pressure factor (F_P) for vertical and horizontal process vessels is equal to the value given on the *y*-axis of Figure 7.6. This, clearly, is a simplification but should be valid for the expected accuracy of this technique. Hence, the equation for F_P for process vessels is given by Equation 7.10.

Figure 7.6 Pressure Factors for Carbon Steel Vessels



(7.10)

$$F_{P,vessel} = \begin{cases} = 1 \quad \text{for } t < t_{min} \text{ and } P > -0.5 \text{ barg} \\ = \dfrac{\dfrac{(P+1)D}{(2)(944)(0.9) - 1.2(P+1)} + CA}{t_{min}} \quad \text{for } t > t_{min} \text{ and } P > \ -0.5 \text{ barg} \\ = 1.25 \quad \text{for } P < -0.5 \text{ barg} \end{cases}$$

where D is the vessel diameter in m, P is the operating pressure in barg, CA is the corrosion allowance (assumed to be 0.00315 m), and t_{min} is the minimum allowable vessel thickness (assumed to be 0.0063 m). A value of S = 944 bar has been assumed for carbon steel. As the operating temperature increases, the value of S decreases (see Figure 7.5) and the accuracy of F_p drops. For operating pressures less than –0.5 barg, the vessel must be designed to withstand full vacuum, that is, 1 bar of external pressure. For such operations, strengthening rings must be installed into the vessels to stop the vessel walls from buckling. A pressure factor of 1.25 should be used for such conditions, and this is shown in Figure 7.6.

Pressure factors for different equipment are given in Appendix A, Equation A.3, and Table A.2. These pressure factors are presented in the general form given by Equation A.3:

(A.3)

$$\log_{10} F_p = C_1 + C_2 \log_{10} P + C_3(\log_{10} P)^2$$

This equation is clearly different from Equation 7.10 for process vessels. Moreover, the value predicted by this equation (using the appropriate constants) gives values of F_P much smaller than those for vessels at the same pressure. This difference arises from the fact that for other equipment, the internals of the equipment make up the major portion of the cost. Therefore, the cost of a thicker outer shell is a much smaller fraction of the equipment cost than for a process vessel, which is strongly dependent on the weight of the metal. Example 7.11 considers the effect of pressure on a shell-and-tube heat exchanger.

Example 7.11

- **a.** Repeat Example 7.10 except consider the case when the operating pressures in both the shell- and the tube-side are 100 barg.
- **b.** Explain why the pressure factor for the heat exchanger is much smaller than for any of the process vessels shown in Figure 7.6.

Solution

- **a.** From Example 7.10, C_P^o(2001) = \$25,000, $F_P = 1$
  From Table A.2, for $5<P<140$ barg, $C_1 = 0.03881$, $C_2 = -0.11272$, $C_3 = 0.08183$
  Using Equation A.3 and substituting for P = 100 barg and the above constants,
  $\log_{10}F_P = 0.03881 - 0.11272\log_{10}(100) + 0.08183[\log_{10}(100)]^2 = 0.1407$
  $$F_P = 10^{0.1407} = 1.383$$

From Equation A.4:

$$C_{BM}(2001) = C_p^o(2001)[B_1 + B_2F_PF_M] = \$25{,}000[1.63 + 1.66(1.383)(1.0)] = \$98{,}100$$
$$C_{BM}(2006) = \$98{,}100\,(500/397) = \$123{,}590$$

- **b.** Compared with Figure 7.6, this pressure factor (1.383) is much less than any of the vessels at P = 100 barg. Why?

The answer lies in the fact that much of the cost of a shell-and-tube heat exchanger is associated with the cost of the tubes that constitute the heat exchange surface area. Tubing is sold in standard sizes based on the BWG (Birmingham wire gauge) standard. Tubes for heat exchangers are typically between 19.1 and 31.8 mm (3/4 and 1-1/4 inch) in diameter and between 2.1 and 0.9 mm (0.083 and 0.035 inch) thick, corresponding to BWGs of 14 to 20, respectively. Using Equation 7.9, the maximum operating pressure of a 25.4 mm (1 inch) carbon steel tube can be estimated (assume that CA is zero), the results are as follows.

BWG	Thickness (t) (mm)	P (from Eq. 7.9) (barg)
20	0.889	59.1
18	1.244	81.8
16	1.651	106.9
14	2.108	134.1

From the table, it is evident that even the thinnest tube normally used for heat exchangers is capable of withstanding pressures much greater than atmospheric. Therefore, the most costly portion of a shell-and-tube heat exchanger (the cost of the tubes) is relatively insensitive to pressure. Hence, it makes sense that the pressure factors for this type of equipment are much smaller than those for process vessels at the same pressure.

The purchased cost of the equipment for the heat exchanger in Example 7.11 would be C_P(2006) = (\$25,000)(1.383) (500/394) = \$43,880. If this equipment cost were multiplied by the bare module factor for the base case, the cost would become C_{BM} = (\$43,880)(3.29) = \$144,360. This is 16% greater than the \$124,490 calculated in Example 7.11. The difference between these two costs results from assuming, in the latter case, that all costs increase in direct proportion to the increase in material cost. This is far from the truth. Some costs, such as insulation, show small changes with the cost of materials, whereas other costs, such as installation materials, freight, labor, and so on, are impacted to varying extents. The method of equipment module costing accounts for these variations in the bare module factor.

Finally, some equipment is unaffected by pressure. Examples are tower trays and packing. This "equipment" is not subjected to significant differential pressure because it is surrounded by process fluid. Therefore, in Equation A.3, use $C_1 = C_2 = C_3 = 0$. Some other equipment also has zero for these constants. For example, compressor drives are not exposed to the process fluid and so are not significantly affected by operating pressure. Other equipment, such as compressors, do not have pressure corrections because such data were not available. Use of these cost correlations for equipment outside the pressure range shown in Table A.2 should be done with extreme caution.

Materials of Construction (MOCs). The choice of what MOC to use depends on the chemicals that will contact the walls of the equipment. As a guide, Table 7.9, excerpted from Sandler and Luckiewicz [14], may be used for preliminary MOC selection. However, the interaction between process streams and MOCs can be very complex and the compatibility of the MOC with the process stream must be investigated fully before the final design is completed.

Talbe 7.9 Corrosion Characteristics for Some Materials of Construction

Chemical Component	Carbon Steel	304 Stainless Steel	316 Stainless Steel	Aluminum	Copper	Brass	Monel	Hastelloy C	Titanium	TFE	Graphite
Acetaldehyde	N		A			C		A	A	A	A
Acetic acid, glacial	N		A	A	A	C	B	A	A	A	A
Acetic acid, 20%	N	A	A	A	A	C	B	A	A	A	A
Acetic anhydride	N	A	B	A	A	C		A	A	A	A
Acetone	A	A	A	A	A	A	A	A	A	A	
Ammonia, 10%	C	A	A	C	N	N	N	A	A	A	A
Aniline	A	A	A	N	N	N	A	A	A	A	A
Aqua regia	N	N	N	N	N	N	N	C	A	A	
Benzaldehyde		A	A	A	A	A	A	A	A	A	A
Benzene	A	A	A	A	A	A	A	A	A	A	A
Benzoic acid		C	A					A	B	A	A
Furfural	A	C	C	A	A	A	A	A	A	A	A
Gasoline	C	A	A	A	A	A	A	A	N	A	A
Heptane	A	A	A	A	A	A	A	A	A	A	A
Hexane		A	A	A			A	A	A	A	A
HCl, 0–25%	N	N	N	N	C	N	C	C	C	A	A
HCl, 25–37%	N	N	N	N	C	N	C	C	C	A	A
HF, 30%	N	B	B	N	N	N	A	A	N	A	A
HF, 60%	N	B	B	N	N	N	A	A	N	A	A
H_2O_2, 30%	C	C	A	C	C	N	C	A	A	A	A
H_2O_2, 90%	C	C	A	C	C	N		A	A	A	A
H_2S, aqueous	C	C	A	A	N	N	N	A	A	A	A
Maleic acid		A	A			A		A	A	A	A
Methanol		A	A	A	A		A	A	A	A	A
Methyl chloride		A	A	N			A	A	A	A	A
Methyl ethyl ketone	A	A	A	A	A		A	A	A	A	A
Methylene chloride		A	A		N		N	A	A	A	A
Napthalene		A	A	A			A	A	A	A	A
Nitric acid, 10%	N	A	A	B	N		N	A	A	A	A
Nitric acid, 50%	N	C	C	B	N		N	A	A	A	N
Oleic acid	C	A	A	A	C		A	A	A	A	A
Oxalic acid	C	C	B	C	C		A	A	A	A	A
Phenol	N	C	C	B	N		A	A	A	A	A
Phosphoric acid, 0–50%	C	C	C	N	C		C	A	B	A	A
Phosphoric acid, 51–100%	C	C	C	N	C		C	A	B	A	A
Propyl alcohol		A	A	A	A			A	A	A	
Sodium hydroxide, 20%	A	A	A	N	C	N	A	A	A	A	A
Sodium hydroxide, 50%	A	A	A	N	C	N	A	A	A	A	A
Stearic acid		A	A	A	A		B	A	A	A	A
Sulfuric acid, 0–10%	N	N	N	N	N		C	A	B	A	A
Sulfuric acid, 10–75%	N	N	N	N	N		C	A	C	A	A
Sulfuric acid, 75–100%	N	N	N	N	N		C	C	N	A	A
Tataric acid		A	A	A	A		C	A	A	A	A
Toluene	A	A	A	A	A		A	A	A	A	
Urea		A	A	A				A	A	A	A
Xylene		A	A					A	A	A	A

A = acceptable; B = acceptable up to 30°C; C = caution, use under limited conditions; N = not recommended; no entry = information is not available. (Reproduced from Sandler and Luckiewicz, *Practical Process Engineering, a Working Approach to Plant Design*, with permission of XIMIX, Inc. Philadelphia, 1987.)

Many *polymeric* compounds are nonreactive in both acidic and alkaline environments. However, polymers generally lack the structural strength and resilience of metals. Nevertheless, for operations at less than about 120°C in corrosive environments the use of polymers as liners for steel equipment or incorporated into fiberglass structures (at moderate operating pressures) often gives the most economical solution. The most common MOCs are still *ferrous* alloys, in particular carbon steel. *Carbon steels* are distinguished from other ferrous alloys such as wrought and cast iron by the amount of carbon in them.

Carbon steel has less than 1.5 wt% carbon, can be given varying amounts of hardness or ductility, is easy to weld, and is cheap. It is still the material of choice in the CPI when corrosion is not a concern.

- *Low-alloy steels* are produced in the same way as carbon steel except that amounts of chromium and molybdenum are added (chromium between 4 and 9 wt%). The molybdenum increases the strength of the steel at high temperatures, and the addition of chromium makes the steel resistant to mildly acidic and oxidizing atmospheres and to sulfur-containing streams.
- *Stainless steels* are so-called high-alloy steels containing greater than 12 wt% chromium and possessing a corrosion-resistant surface coating, also known as a passive coating. At these chromium levels, the corrosion of steel to rusting is reduced by more than a factor of 10. Chemical resistance is also increased dramatically.
- *Nonferrous alloys* are characterized by higher cost and difficulty in machining. Nevertheless, they possess improved corrosion resistance.

Aluminum and its alloys have a high strength-to-weight ratio and are easy to machine and cast, but in some cases are difficult to weld. The addition of small amounts of other metals—for example, magnesium, zinc, silicon, and copper—can improve the weldability of aluminum. Generally, corrosion resistance is very good due to the formation of a passive oxide layer, and aluminum has been used extensively in cryogenic (low-temperature) operations.

Copper and its alloys are often used when high thermal conductivity is required. Resistance to seawater and nonoxidizing acids such as acetic acid is very good, but copper alloys should not be used for services that contact ammonium ions (NH_4^-) or oxidizing acids. Common alloys of copper include *brasses* (containing 5–45 wt% zinc) and *bronzes* (containing tin, aluminum, and/or silicon).

- *Nickel and its alloys* are alloys in which nickel is the major component.

Nickel-copper alloys are known by the name Monel, a trademark of the International Nickel Corp. These alloys have excellent resistance to sulfuric and hydrochloric acids, salt water, and some caustic environments.

Nickel-chromium alloys are known by the name Inconel, a trademark of the International Nickel Corp. These alloys have excellent chemical resistance at high temperatures. They are also capable of withstanding attack from hot concentrated aqueous solutions containing chloride ions.

Nickel-chromium-iron alloys are known by the name Incoloy, a trademark of the International Nickel Corp. These alloys have characteristics similar to Inconel but with slightly less resistance to oxidizing agents.

Nickel-molybdenum alloys are known by the name Hastelloy, a trademark of the Cabot Corp. These alloys have very good resistance to concentrated oxidizing agents.

- *Titanium and its alloys* have good strength-to-weight ratios and very good corrosion resistance to oxidizing agents. However, it is attacked by reducing agents, it is relatively expensive, and it is difficult to weld.

As previously shown, the combination of operating temperature and operating pressure will also affect the choice of MOC. From Table 7.9, it is evident that the number of MOCs available is very large and that the correct choice of materials requires input from a trained metallurgist.

Moreover, information about the cost of materials presented in this text is limited to a few different MOCs. The approximate relative cost of some common metals is given in Table 7.10. As a very approximate rule, if the metal of interest does not appear in Appendix A, then Table 7.10 can be used to find a metal that has approximately the same cost. As the metallurgy becomes more “exotic,” the margin for error becomes larger, and the data provided in this text will lead to larger errors in estimating the plant cost than for a plant constructed of carbon steel or stainless steel.

Talbe 7.10 Relative Costs of Metals Using Carbon Steel as the Base Case

Material	Relative Cost
Carbon steel	Base case (lowest)
Low-alloy steel	Low to moderate
Stainless steel	Moderate
Aluminum and aluminum alloys	Moderate
Copper and copper alloys	Moderate
Titanium and titanium-based alloys	High
Nickel and nickel-based alloys	High

To account for the cost of different materials of construction, it is necessary to use the appropriate material factor, F_M, in the bare module factor. This material factor is *not* simply the relative cost of the material of interest to that of carbon steel. The reason is that the cost to produce a piece of equipment is not directly proportional to the cost of the raw materials. For example, consider the cost of a process vessel as discussed in the previous section. Just as the bare module cost was broken down into factors relating to the purchased cost of the equipment (Tables 7.6 and 7.8), the purchased cost (or at least the manufacturing cost) can be broken down into factors relating to the cost of manufacturing the equipment. Many of these costs will be related to the size of the vessel that is in turn related to the vessel's weight, W_{vessel}. An example of these costs is given in Table 7.11.

Talbe 7.11 Costs Associated with the Manufacture of a Process Vessel

Factors Associated with the Manufacturing Cost of a Vessel	Relationship relating Cost to vessel weight, W_{vessel}
Direct Expenses	
Cost of raw materials	$\beta_{RM} W_{vessel}$
Machining costs	$\beta_{MC} W_{vessel}$
Labor costs	$\beta_L W_{vessel}$
Indirect Costs	
Overhead	$\beta_{OH} \beta_L W_{vessel}$
Engineering expenses	$\beta_E (\beta_{RM} + \beta_{MC}) W_{vessel}$
Contingencies	$\beta_{Cont} W_{vessel}$
Total manufacturing cost	$[\beta_{RM} + \beta_{MC} + \beta_L + \beta_{OH} \beta_{LI} + \beta_E (\beta_{RM} + \beta_{MC}) + \beta_{Cont}] W_{vessel}$

From Table 7.11, it is clear that the cost of the vessel is proportional to its weight. Therefore, the cost will be proportional to the vessel thickness, and thus the pressure factor derived in the previous section is valid (or at least is a reasonably good approximation). The effect of different MOCs is connected to the factor β_{RM}. Clearly, as the raw material costs increase, the total manufacturing costs will not increase proportionally to β_{RM}. In other words, if material *X* is 10 times as expensive as carbon steel, a vessel made from material *X* will be less than 10 times the cost of a similar vessel made from carbon steel. For example, over the last 15 years, the cost of stainless steel has varied between 4.7 and 7.0 times the cost of carbon steel [16]. However, the cost of a stainless steel process vessel has varied in the approximate range of 2.3 to 3.5 times the cost of a carbon steel vessel for similar service.

Materials factors for the process equipment considered in this text are given in Appendix A, Tables A.3–A.6, and Figures A.18 and A.19. These figures are constructed using averaged data from the following sources: Peters and Timmerhaus [2], Guthrie [9, 10], Ulrich [5], Navarrete [11], and Perry et al. [3]. Example 7.12 illustrates the use of these figures and tables.

Example 7.12

Find the bare module cost of a floating-head shell-and-tube heat exchanger with a heat transfer area of 100 m^2 for the following cases.

a. The operating pressure of the equipment is 1 barg on both shell and tube sides, and the MOC of the shell and tubes is stainless steel.

b. The operating pressure of the equipment is 100 barg on both shell and tube sides, and the MOC of the shell and tubes is stainless steel.

From Example 7.10, $C_P^o(2001)$ = \$25,000 and $C_P^o(2006)$ = \$25,000 (500/397) = \$31,490.

a. From Example 7.10, at 1 barg, $F_P = 1$

From Table A.3 for a shell-and-tube heat exchanger made of SS, Identification No. = 5 and using Figure A.8, $F_M = 2.73$

From Equation A.4,

$C_{BM}(2006) = C_P^o[B_1 + B_2F_PF_M] = \$31{,}490[1.63 + 1.66(1.0)(2.73)] = \$194{,}000$

b. From Example 7.11 for P = 100 barg, $F_P = 1.383$

From (a) above, $F_M = 2.73$

Substituting these values in to Equation A.4,

$C_{BM}(2006) = C_P^o[B_1 + B_2F_PF_M] = \$31{,}490[1.63 + 1.66(1.383)(2.73)] = \$248{,}600$

The last three examples all considered the same size heat exchanger made with different materials of construction and operating pressure. The results are summarized below.

Example	Pressure	Materials	F_{BM}	Cost
7.10	ambient	CS tubes/shell	3.29	\$103,590
7.11	100 barg	CS tubes/shell	3.93	\$123,590
7.12a	ambient	SS tubes/shell	6.16	\$194,000
7.12b	100 barg	SS tubes/shell	7.90	\$248,600

These results reemphasize the point that the cost of the equipment is strongly dependent on the materials of construction and the pressure of operation.

7.3.5 Combination of Pressure and MOC Information to Give the Bare Module Factor, F_{BM}, and Bare Module Cost, C_{BM}

In Examples 7.10–7.12, the bare module factors and costs were calculated using Equation A.4. The form of this equation is not obvious, and its derivation is based on the approach used by Ulrich [5]:

(7.11)

$$\text{Cost of Equipment} = C_p^o F_P F_M$$

This is the equipment cost at operating conditions:

(7.12)

$$\text{Cost for Equipment Installation (for base conditions)} = C_p^o(F_{BM}^o - 1)$$

This cost is calculated by taking the bare module cost, at base conditions, and subtracting the cost of the equipment at the base conditions.

The incremental cost of equipment installation due to nonbase case conditions is

(7.13)

$$= C_p^o(F_P F_M - 1) f_{P\&I}$$

This cost is based on the incremental cost of equipment due to nonbase conditions multiplied by a factor, ($f_{P\&I}$), that accounts for the fraction of the installation cost that is associated with piping and instrumentation. The values of $f_{P\&I}$ are modified from Guthrie [9, 10] to account for an increase in the level and cost of instrumentation that modern chemical plants enjoy compared with that at the time of Guthrie's work.

Equations 7.11 through 7.13 can be combined to give the following relationship:

(7.14)

$$\text{Bare module cost, } C_{BM} = C_p^o F_P F_M + C_p^o(F_{BM}^o - 1) + C_p^o(F_P F_M - 1) f_{P\&I}$$

$$= C_p^o[F_P F_M(1 + f_{P\&I}) + F_{BM}^o - 1 - f_{P\&I}] = C_p^o[B_1 + B_2 F_P F_M]$$

Equation 7.13 is the same as Equation A.4, with $B_1 = F_{BM}^o - 1 - f_{P\&I}$ and $B_2 = 1 + f_{P\&I}$.

7.3.6 Algorithm for Calculating Bare Module Costs

The following six-step algorithm is used to estimate actual bare module costs for equipment from the figures in Appendix A.

1. Using the correct figure in Appendix A (Figures A.1–A.17), or the data in Table A.1, obtain C_p^o for the desired piece of equipment. This is the purchased equipment cost for the base case (carbon steel construction and near ambient pressure).
2. Find the correct relationship for the bare module factor. For exchangers, pumps, and vessels, use

Equation A.4 and the data in Table A.4. For other equipment, the form of the equation is given in Table A.5.

3. For exchangers, pumps, and vessels, find the pressure factor, F_P, Table A.2 and Equation A.2 or A.3, and the material of construction factor, F_M, Equation A.4, Table A.3, and Figure A.18. Use Equation A.4 to calculate the bare module factor, F_{BM}.
4. For other equipment find the bare module factor, F_{BM}, using Table A.6 and Figure A.19.
5. Calculate the bare module cost of equipment, C_{BM}, from Equation 7.6.
6. Update the cost from 2001 (CEPCI – 397) to the present by using Equation 7.4.

Example 7.13 illustrates the six-step algorithm for the case of a distillation column with associated trays.

Example 7.13

Find the bare module cost (in 2006) of a stainless steel tower 3 m in diameter and 30 m tall. The tower has 40 stainless steel sieve trays and operates at 20 barg.

The costs of the tower and trays are calculated separately and then added together to obtain the total cost.

For the tower,

a. Volume = $\pi D^2 L/4 = (3.14159)(3)^2(30)/4 = 212.1\ \text{m}^3$

From Equation A.1,

$$\log_{10} C_p^o(2001) = 3.4974 + 0.4485\log_{10}(212.1) + 0.1074\{\log_{10}(212.1)\}^2 = 5.1222$$

$$C_p^o(2001) = 10^{5.1222} = \$132{,}500$$
$$C_p^o(2006) = \$132{,}500(500/397) = \$166{,}880$$

b. From Equation A.3 and Table A.4, $F_{BM} = 2.25 + 1.82\ F_M F_P$

c. From Equation 7.10 with P = 20 barg and D = 3 m,

$$F_{P,vessels} = \frac{\dfrac{(20+1)3}{(2)[(944)(0.9) - 0.6(20+1)]} + 0.00315}{0.0063} = 6.47$$

From Table A.3, identification number for stainless steel vertical vessel = 20; from Figure A.8, $F_M = 3.11$

$$F_{BM} = 2.25 + 1.82(6.47)(3.11) = 38.87$$

d. $C_{BM}(2006) = (166{,}880)(38.87) = \$6{,}486{,}000$

For the trays,

a. Tray (tower) area = $\pi D^2/4 = 7.0686$

From Equation A.1,

$$\log_{10} C_p^o(2001) = 2.9949 + 0.4465\log_{10}(7.0686) + 0.3961\{\log_{10}(7.0686)\}^2 = 3.6599$$

$$C_p^o(2001) = 10^{3.6599} = \$4570$$
$$C_p^o(2006) = \$4{,}570(500/397) = \$5{,}756$$

From Table A.5,

$$C_{BM} = C_p N F_{BM} f_q$$

$N = 40$

$f_q = 1.0$ (since number of trays > 20, Table A.5)

From Table A.6, SS sieve trays identification number = 61; from Figure A.9, $F_{BM} = 1.83$

$$C_{BM,trays}(2006) = (\$5,756)(40)(1.83)(1.0) = \$421,300$$

For the tower plus trays,

$$C_{BM,tower+trays}(2006) = \$6,486,000 + \$\ 421,300 = \$\ 6,908,300$$

7.3.7 Grass Roots and Total Module Costs

The term *grass roots* refers to a completely new facility in which we start the construction on essentially undeveloped land, a grass field. The term *total module cost* refers to the cost of making small-to-moderate expansions or alterations to an existing facility.

To estimate these costs, it is necessary to account for other costs in addition to the direct and indirect costs. These additional costs were presented in Table 7.6 and can be divided into two groups.

Group 1: Contingency and Fee Costs: The contingency cost varies depending on the reliability of the cost data and completeness of the process flowsheet available. This factor is included in the evaluation of the cost as a protection against oversights and faulty information. Unless otherwise stated, values of 15% and 3% of the bare module cost are assumed for contingency costs and fees, respectively. These are appropriate for systems that are well understood. Adding these costs to the bare module cost provides the *total module cost*.

Group 2: Auxiliary Facilities Costs: These include costs for site development, auxiliary buildings, and off-sites and utilities. These terms are generally unaffected by the materials of construction or the operating pressure of the process. A review of costs for these auxiliary facilities by Miller [17] gives a range of approximately 20% to more than 100% of the bare module cost. Unless otherwise stated, these costs are assumed to be equal to 50% of the bare module costs for the base case conditions. Adding these costs to the total module cost provides the *grassroots cost*.

The total module cost can be evaluated from

(7.15)

$$C_{TM} = \sum_{i=1}^{n} C_{TM,i} = 1.18 \sum_{i=1}^{n} C_{Bm,i}$$

and the grassroots cost can be evaluated from

(7.16)

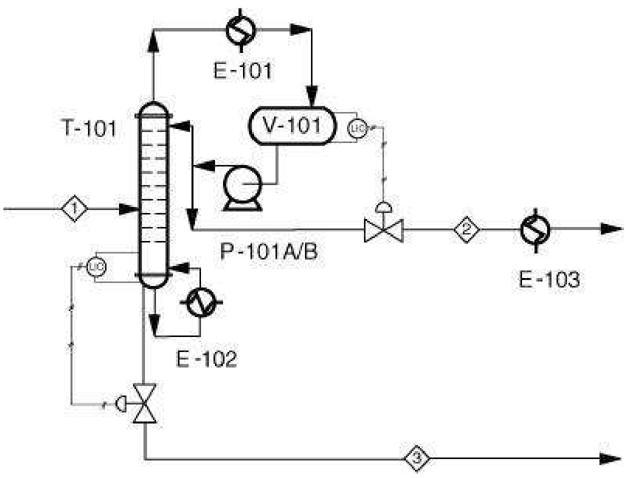
$$C_{GR} = C_{TM} + 0.50\sum_{i=1}^{n} C^{o}_{BM,i}$$

where n represents the total number of pieces of equipment. The use of these equations is shown in Example 7.14.

Example 7.14

A small expansion to an existing chemical facility is being investigated, and a preliminary PFD of the process is shown in Figure E7.14.

Figure E7.14 PFD for Example 7.14



The expansion involves the installation of a new distillation column with a reboiler, condenser, pumps, and other associated equipment. A list of the equipment, sizes, materials of construction, and operating pressures is given in Table E7.14(a). Using the information in Appendix A, calculate the total module cost for this expansion in 2006.

Table E7.14(a) Information on Equipment Required for the Plant Expansion Described in Example 7.14

Equipment No.	Capacity / Size	Material of Construction*	Operating Pressure (barg†
E-101 Overhead condenser	Area = 170 m^2 Shell and tube (floating-head)	Tube - CS Shell - CS	Tube = 5.0 Shell = 5.0
E-102 Reboiler	Area = 205 m^2 Shell and tube (floating-head)	Tube - SS Shell - CS	Tube = 18.0 Shell = 6.0
E-103 Product cooler	Area = 10 m^2 (double pipe)	All CS construction	Inner = 5.0 Outer = 5.0
P-101A/B Reflux pumps	$Power_{shaft}$ = 5 kW Centrifugal	CS	Discharge = 5.0
T-101 Aromatics column	Diameter = 2.1 m Height = 23 m	Vessel - CS	Column = 5.0
	32 sieve trays	Trays - SS	
V-101 Reflux drum	Diameter = 1.8 m Length = 6 m Horizontal	Vessel - CS	Vessel = 5.0

*CS = Carbon steel; SS = Stainless steel
†barg = bar gauge, thus 0.0 barg = 1.0 bar

The same algorithm presented above is used to estimate bare module costs for all equipment. This information is listed in Table E7.14(b), along with purchased equipment cost, pressure factors, material factors, and bare module factors.

Table E7.14(b) Results of Capital Cost Estimate for Example 7.14

Equipment	F_P	F_M	F_{BM}	C_p^o (2001) ($)	C_{BM} (2001) ($)	C_{BM}^o (2001) ($)
E-101	1.0	1.0	3.29	33,000	108,500	108,500
E-102	1.023	1.81	4.70	36,900	177,900	121,300
E-103	1.0	1.0	3.29	3700	12,300	12,300
P-101A/B	1.0	1.55	3.98	(2)(3200)	(2)(12,600)	(2)(10,300)
T-101	1.681	1.0	5.31	54,700	290,700	222,800
32 trays		1.83	1.83	(32)(2200)	131,200	71,700
V-101	1.513	1.0	3.79	13,500	51,200	40,600
Totals				219,900	797,000	597,800
CEPCI = 397						

The substitutions from Table E7.14(b) are made into Equations 7.15 and 7.16 to determine the total module cost and the grassroots cost.

$$\text{total module cost } (C_{TM}) = 1.18\sum_{i=1}^{n} C_{BM,i} = 1.18(\$797{,}000)(500/397) = \$1{,}184{,}000$$

$$\text{grassroots cost } (C_{GR}) = C_{TM} + 0.50\sum_{i=1}^{n} C_{BM,i}^o = \$1{,}184{,}00 + 0.50(\$597{,}800)(500/397)$$
$$= \$1{,}561{,}000$$

Although the grassroots cost is not appropriate here (because we have only a small expansion to an existing facility), it is shown for completeness.

7.3.8 A Computer Program (CAPCOST) for Capital Cost Estimation Using the Equipment Module

Approach

For processes involving only a few pieces of equipment, estimating the capital cost of the plant by hand is relatively easy. For complex processes with many pieces of equipment, these calculations become tedious. To make this process easier, a computer program has been developed that allows the user to enter data interactively and obtain cost estimates in a fraction of the time required by hand calculations with less chance for error. The program (CAPCOST_2008.xls) is programmed in Microsoft Excel, and a template copy of the program is supplied on the CD that accompanies this book.

The program is written in the Microsoft Windows programming environment. The program requires the user to input information about the equipment—for example, the capacity, operating pressure, and materials of construction. The cost data can be adjusted for inflation by entering the current value of the CEPCI. Other information such as output file names and the number of the unit (100, 200, etc.) is also required.

The equipment options available to the user are given below.

- Blenders
- Centrifuges
- Compressors and blowers without drives
- Conveyors
- Crystallizers
- Drives for compressors, blowers, and pumps
- Dryers
- Dust collectors
- Evaporators and vaporizers
- Fans with electric drives
- Filters
- Fired heaters, thermal fluid heaters, and packaged steam boilers
- Furnaces
- Heat exchangers
- Mixers
- Process vessels with/without internals
- Power recovery equipment
- Pumps with electric drives
- Reactors
- Screens
- Storage vessels (fixed roof and floating roof)
- Towers
- User-added modules

The type of equipment required can be entered by using the mouse-activated buttons provided on the first worksheet. The user will then be asked a series of questions that appear on the screen. The user will be required to identify or enter the same information as would be needed to do the calculations by hand—that is, operating pressure, materials of construction, and the size of the equipment. The same information as contained in the cost charts and tables in Appendix A is embedded in the program, and the program should give the same results as hand calculations using these charts.

When the data for equipment are entered, a list of the costs on the first worksheet is updated. The use of the spreadsheet is explained in the CAPCOST.avi help files contained on the CD, and the reader is encouraged to view the file prior to using the software. You are strongly advised to verify the results of Example E7.14 for yourself prior to using the program to solve problems in the back of this chapter.

7.4 Summary

In this chapter, the different types of capital cost estimating techniques that are available were reviewed. The accuracy of the different estimates was shown to increase significantly with the time involved in completion and the amount of data required. The information required to make an equipment module estimate based on data from the major process equipment was also covered. The effects of operating pressure and materials of construction on the bare module cost of equipment were reviewed. Several examples were given to show how the installed cost of equipment is significantly greater than the purchased cost and how the installed cost increases with increased pressure and materials of construction. The use of cost indices to adjust for the effects of inflation on equipment costs was considered, and the Chemical Engineering Plant Cost Index (CEPCI) was adopted for all inflation adjustments. The concepts of grass roots and total module costs were introduced in order to make estimates of the total capital required to build a brand new plant or make an expansion to an existing facility. To ease the calculation of the various costs, a computer program for cost estimation was introduced. This chapter contains the basic approach to estimating capital costs for new chemical plants and expansions to existing plants, and mastery of this material is assumed in the remaining chapters.

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Short Answer Questions

1. What are the three main factors that determine the capital cost of a piece of equipment such as a heat exchanger at a given time?

2. What is the Chemical Engineering Plant Cost Index (*CEPCI*) used for, and what does it measure?

3. What is the difference between the total module cost and the grassroots cost of a chemical process?

4. When would you use a cost exponent of 0.6?

5. What is meant by the economy of scale?

6. What is a Lang Factor?

The pressure factor F_p for a shell-and-tube heat exchanger is significantly smaller than for a vessel

7. over the same pressure range. Why is this so?

Problems

8. The cost* of a recent plant in Alberta, Canada, to produce 1.27 million tonne/y of polyethylene was \$540 million. Estimate what the range of cost estimates would likely have been for a Class 5, a Class 3, and a Class 1 estimate.

9. In Appendix A, Figures A.1–A.17, the purchased costs for various types of equipment are given. The *y*-axis is given as the cost of the equipment per unit of capacity, and the *x*-axis is given as the capacity. The capacity is simply the relevant sizing parameter for the equipment. Identify all equipment that does not conform to the principle of the economy of scale.

10. A process vessel was purchased in the United Kingdom for our plant in the United States in 1993. A similar vessel, but of different capacity, was purchased in 1998. From the data given below, estimate the cost in U.S.\$ of a vessel of 120 m^3 capacity purchased today (assume the current CEPCI = 500).

Date	Vessel Capacity (m^3)	Purchased Cost (Pounds Sterling = £)	Exchange Rate
1993	75	£ 7,800	\$1.40/£
1998	155	£ 13,800	\$1.65/£
2007	120		\$2.00/£

11. You have been hired as a consultant to a legal firm. Part of your assignment is to determine the size of a storage tank purchased in 1978 (CEPCI = 219), before computerization of records. Many records from this era were destroyed in a fire (not in the plant, but in a distant office building). The tank was replaced every 10 years, and the sizes have changed due to plant capacity changes. You have the information in the table below. Estimate the original capacity of this vessel.

Date	Tank Capacity (1000 gal)	Purchased Cost
1978	?	\$35,400
1988	105	\$45,300
1998	85	\$45,500

12. In your role as a consultant to a legal firm, you have been requested to determine whether calculations submitted in a legal action are valid. The problem is to determine what year a compressor was placed into service. The information in the table is available. It is claimed that the compressor was placed into service in 1976. History suggests that during the period from 1976 to 1985 there was significant inflation. Do you believe the information submitted is correct? If not, what year do you believe the compressor to be placed into service? Use *CEPCI* = 500 for 2006.

Date	Compressor Power (kW)	Total Module Cost (in 10^3\$)
???	1000	645.93
2000	500	500.00
2006	775	811.68

Note: CEPCI (1986) = 318, CEPCI (1981) = 297, CEPCI (1976) = 192.

13. When designing equipment for high-temperature and high-pressure service, the maximum allowable stress as a function of temperature of the material of construction is of great importance. Consider a cylindrical vessel shell that is to be designed for pressure of 150 bar (design pressure). The diameter of the vessel is 3.2 m, it is 15 m long, and a corrosion allowance of 6.35 mm (1/4″) is to be used. Construct a table that shows the thickness of the vessel walls in the temperature range of 300 to 500°C (in 20°C increments) if the materials of construction are (a) ASME SA515-grade carbon steel and (b) ASME SA-240-grade 316 stainless steel.

14. Using the results of Problem 13, determine the relative costs of the vessel using the two materials of construction (CS and 316 SS) over the temperature range. You may assume that the cost of the vessel is directly proportional to the weight of the vessel and that the 316 SS costs 3.0 times that of CS. Based on these results, which material of construction is favored over the temperature range 300–500°C for this vessel?

The following problems may be solved either by using hand calculations or by using CAPCOST (use a value of CEPCI = 500).

15. Determine the bare module cost of a 1-shell pass, 2-tube pass (1-2) heat exchanger designed for the following operating conditions:

Maximum operating pressure (tube side) = 30 barg

Maximum operating pressure (shell side) = 5 barg

Process fluid in tubes requires stainless steel MOC

Shell-side utility (cooling water) requires carbon steel MOC

Heat exchange area = 160 m^2

16. Repeat Problem 15, except reverse the shell-side and tube-side fluids. Are your results consistent with the heuristics for heat exchangers given in Chapter 11? Which heuristic is relevant?

17. In Chapter 15, the concepts of heat-exchanger networks and pinch technology are discussed. When designing these networks to recover process heat, it is often necessary to have a close temperature approach between process streams, which leads to large heat exchangers with multiple shells. Multiple-shell heat exchangers are often constructed from sets of 1-2 shell and tube exchangers stacked together. For costing considerations, the cost of the multiple-shell heat exchanger is best estimated as a number of smaller 1-2 exchangers. Consider a heat exchanger constructed of carbon steel and designed to withstand a pressure of 20 barg in both the shell and tube sides. This equipment has a heat exchange area of 400 m^2. Do the following.

a. Determine the bare module cost of this 4-shell and 8-tube pass heat exchanger as four, 1-2 exchangers, each with a heat-exchange area of 100 m^2.
b. Determine the bare module cost of the same exchanger as if it had a single shell.
c. What is the name of the principle given in this chapter that explains the difference between the two answers in (a) and (b)?

18. A distillation column is initially designed to separate a mixture of toluene and xylene at around ambient temperature (say, 100°C) and pressure (say, 1 barg). The column has 20 stainless steel valve trays and is 2 m in diameter and 14 m tall. Determine the purchased cost and the bare module cost using a *CEPCI* = 500.

19. A column with similar dimensions, number of trays, and operating at the same conditions as given in Problem 18 is to be used to separate a mixture containing the following chemicals. For each case determine the bare module cost using a *CEPCI* = 500.

    a. 10% nitric acid solution
    b. 50% sodium hydroxide solution
    c. 10% sulfuric acid solution
    d. 98% sulfuric acid solution

    Hint: you may need to look for the relevant MOC for part (d) on the Internet or another resource.

It is recommended that the following problems be solved using CAPCOST (use a value of CEPCI = 500).

Determine the bare module, total module, and grassroots cost of the following.

20. Toluene hydrodealkylation plant described in Chapter 1 (see Figures 1.3 and 1.5 and Tables 1.5 and 1.7).

21. Ethylbenzene plant described in Appendix B, project B.2.

22. Styrene plant described in Appendix B, project B.3.

23. Drying oil plant described in Appendix B, project B.4.

24. Maleic anhydride plant described in Appendix B, project B.5.

25. Ethylene oxide plant described in Appendix B, project B.6.

26. Formalin plant described in Appendix B, project B.7.

*http://www.chemicals-technology.com/projects/joffre/

Chapter 8 Estimation of Manufacturing Costs

The costs associated with the day-to-day operation of a chemical plant must be estimated before the economic feasibility of a proposed process can be assessed. This chapter introduces the important factors affecting the manufacturing cost and provides methods to estimate each factor. In order to estimate the manufacturing cost, we need process information provided on the PFD, an estimate of the fixed capital investment, and an estimate of the number of operators required to operate the plant. The fixed capital investment is the same as either the total module cost or the grassroots cost defined in Chapter 7. Manufacturing costs are expressed in units of dollars per unit time, in contrast to the capital costs, which are expressed in dollars. How we treat these two costs, expressed in different units, to judge the economic merit of a process is covered in Chapters 9 and 10.

8.1 Factors Affecting the Cost of Manufacturing a Chemical Product

There are many elements that influence the cost of manufacturing chemicals. A list of the important costs involved, including a brief explanation of each cost, is given in Table 8.1.

Table 8.1 Factors Affecting the Cost of Manufacturing (*COM*) for a Chemical Product (from References [1, 2, and 3])

Factor	Description of Factor
1. Direct Costs	**Factors that vary with the rate of production**
A. Raw materials	Costs of chemical feed stocks required by the process. Flowrates obtained from the PFD.
B. Waste treatment	Costs of waste treatment to protect environment.
C. Utilities	Costs of utility streams required by process. Includes but not limited to a. Fuel gas, oil, and/or coal b. Electric power c. Steam (all pressures) d. Cooling water e. Process water f. Boiler feed water g. Instrument air h. Inert gas (nitrogen, etc.) i. Refrigeration Flowrates for utilities found on the PFD/PIDs.
D. Operating labor	Costs of personnel required for plant operations.
E. Direct supervisory and clerical labor	Cost of administrative, engineering, and support personnel.
F. Maintenance and repairs	Costs of labor and materials associated with maintenance.
G. Operating supplies	Costs of miscellaneous supplies that support daily operation not considered to be raw materials. Examples include chart paper, lubricants, miscellaneous chemicals, filters, respirators and protective clothing for operators, etc.
H. Laboratory charges	Costs of routine and special laboratory tests required for product quality control and troubleshooting.
I. Patents and royalties	Cost of using patented or licensed technology.
2. Fixed Costs	**Factors not affected by the level of production**
A. Depreciation	Costs associated with the physical plant (buildings, equipment, etc.). Legal operating expense for tax purposes.
B. Local taxes and insurance	Costs associated with property taxes and liability insurance. Based on plant location and severity of the process.
C. Plant overhead costs (sometimes referred to as factory expenses)	Catch-all costs associated with operations of auxiliary facilities supporting the manufacturing process. Costs involve payroll and accounting services, fire protection and safety services, medical services, cafeteria and any recreation facilities, payroll overhead and employee benefits, general engineering, etc.
3. General Expenses	**Costs associated with management level and administrative activities not directly related to the manufacturing process**
A. Administration costs	Costs for administration. Includes salaries, other administration, buildings, and other related activities.
B. Distribution and selling costs	Costs of sales and marketing required to sell chemical products. Includes salaries and other miscellaneous costs.
C. Research and development	Costs of research activities related to the process and product. Includes salaries and funds for research-related equipment and supplies, etc.

The cost information provided in Table 8.1 is divided into three categories:

1. **Direct manufacturing costs:** These costs represent operating expenses that vary with production

rate. When product demand drops, production rate is reduced to less than the design capacity. At this lower rate, we would expect a reduction in the factors making up the direct manufacturing costs. These reductions may be directly proportional to the production rate, as for raw materials, or might be reduced slightly—for example, maintenance costs or operating labor.

2. **Fixed manufacturing costs:** These costs are independent of changes in production rate. They include property taxes, insurance, and depreciation, which are charged at constant rates even when the plant is not in operation.
3. **General expenses:** These costs represent an overhead burden that is necessary to carry out business functions. They include management, sales, financing, and research functions. General expenses seldom vary with production level. However, items such as research and development and distribution and selling costs may decrease if extended periods of low production levels occur.

The equation used to evaluate the cost of manufacture using these costs becomes:

Cost of Manufacture (*COM*) = Direct Manufacturing Costs (*DMC*) +
Fixed Manufacturing Costs (*FMC*) + General Expenses (*GE*)

The approach we provide in this chapter is similar to that presented in other chemical engineering design texts [1, 2, 3].

The cost of manufacturing, *COM*, can be determined when the following costs are known or can be estimated:

1. Fixed capital investment (*FCI*): (C_{TM} or C_{GR})
2. Cost of operating labor (C_{OL})
3. Cost of utilities (C_{UT})
4. Cost of waste treatment (C_{WT})
5. Cost of raw materials (C_{RM})

Table 8.2 gives data to estimate the individual cost items identified in Table 8.1 (both tables carry the same identification of individual cost terms). With the exception of the cost of raw materials, waste treatment, utilities, and operating labor (all parts of the direct manufacturing costs), Table 8.2 presents equations that can be used to estimate each individual item. With each equation, a typical range for the constants (multiplication factors) to estimate an individual cost item is presented. If no other information is available, the midpoint value for each of these ranges is used to estimate the costs involved. It should be noted that the best information that is available should always be used to establish these constants. The method presented here should be used only when no other information on these costs is available.

Table 8.2 Multiplication Factors for Estimating Manufacturing Cost* (See Also Table 8.1)

Cost Item from Table 8.1	Typical Range of Multiplying Factors	Value Used in Text
1. Direct Manufacturing Costs		
a. Raw materials	C_{RM}[†]	
b. Waste treatment	C_{WT}[†]	
c. Utilities	C_{UT}[†]	
d. Operating labor	C_{OL}	C_{OL}
e. Direct supervisory and clerical labor	$(0.1 - 0.25)C_{OL}$	$0.18C_{OL}$
f. Maintenance and repairs	$(0.02 - 0.1)FCI$	$0.06FCI$
g. Operating supplies	$(0.1 - 0.2)$(Line 1.F)	$0.009FCI$
h. Laboratory charges	$(0.1 - 0.2)C_{OL}$	$0.15C_{OL}$
i. Patents and royalties	$(0 - 0.06)COM$	$0.03COM$
Total Direct Manufacturing Costs	$C_{RM} + C_{WT} + C_{UT} + 1.33C_{OL} + 0.03COM + 0.069FCI$	
2. Fixed Manufacturing Costs		
a. Depreciation	$0.1FCI$[‡]	$0.1FCI$[‡]
b. Local taxes and insurance	$(0.014 - 0.05)FCI$	$0.032FCI$
c. Plant overhead costs	$(0.50 - 0.7)$(Line 1.D. + Line 1.E + Line 1.F)	$0.708C_{OL} + 0.036FCI$
Total Fixed Manufacturing Costs	$0.708C_{OL} + 0.068FCI$ + depreciation	
3. General Manufacturing Expenses		
a. Administration costs	0.15(Line 1.D. + Line 1.E.+ Line 1.F.)	$0.177C_{OL} + 0.009FCI$
b. Distribution and selling costs	$(0.02 - 0.2)COM$	$0.11COM$
c. Research and development	$0.05COM$	$0.05COM$
Total General Manufacturing Costs	$0.177C_{OL} + 0.009FCI + 0.16COM$	
Total Costs	$C_{RM} + C_{WT} + C_{UT} + 2.215C_{OL} + 0.190COM + 0.146FCI$ + depreciation	

*Costs are given in dollars per unit time (usually per year).
†Costs are evaluated from information given on the PFD and the unit cost.
‡Depreciation costs are covered separately in Chapter 9. The use of 10% of *FCI* is a crude approximation at best.
From references [1], [2], and [3].

By using the midpoint values given in Table 8.2, column 2, the resulting equations for the individual items are calculated in column 3. The cost items for each of the three categories are added together to provide the total cost for each category. The equations for estimating the costs for each of the categories are as follows:

$$DMC = C_{RM} + C_{WT} + C_{UT} + 1.33C_{OL} + 0.069FCI + 0.03COM$$

$$FMC = 0.708C_{OL} + 0.068FCI + \text{depreciation}$$

$$GE = 0.177C_{OL} + 0.009FCI + 0.16COM$$

We can obtain the total manufacturing cost by adding these three cost categories together and solving for the total manufacturing cost, *COM*. The result is

(8.1)

$$COM = 0.280FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM})$$

In Equation (8.1), the depreciation allowance of $0.10FCI$ is added separately.

The cost of manufacture without depreciation, COM_d, is

(8.2)

$$COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM})$$

The calculation of manufacturing costs and expenses is given in Example 8.1.

Example 8.1

The following cost information was obtained from a design for a 92,000 tonne/year nitric acid plant.

Fixed Capital Investment	$11,000,000
Raw Materials Cost	$ 7,950,000/yr
Waste Treatment Cost	$ 1,000,000/yr
Utilities	$ 356,000/yr
Direct Labor Cost	$ 300,000/yr
Fixed Costs	$ 1,500,000/yr

Determine

- **a.** The manufacturing cost in $/yr and $/tonne of nitric acid
- **b.** The percentage of manufacturing costs resulting from each cost category given in Tables 8.1 and 8.2

Using Equation 8.2,

COM_d = (0.180)($11,000,000) + (2.73)($300,000) +
(1.23)($356,000 + $1,000,000 + $7,950,000) = $14,245,000/yr
($14,245,000/yr)/(92,000 tonne/yr) = $155/tonne

From the relationships given in Table 8.2,

Direct Manufacturing Costs = $7,950,000 + $1,000,000 + $356,000 + (1.33)($300,000) + (0.069)($11,000,000) + (0.03)($14,245,000) = $10,891,000

Percentage of manufacturing cost = (100)(10.891)/14.25 = 76%

Fixed Manufacturing Costs = (0.708)($300,000) + (0.068)($11,000,000) = $960,000

Percentage of manufacturing cost = (100)(0.960)/14.25 = 7%

General Expenses = (0.177)($300,000) + (0.009)($11,000,000) + (0.16)($14,245,000) = $2,431,000

Percentage of manufacturing cost = (100)(2.431)/14.25 = 17%

In Example 8.1, the direct costs were shown to dominate the manufacturing costs, accounting for about 76% of the manufacturing costs. Of these direct costs, the raw materials cost, the waste treatment cost, and the cost of utilities accounted for more than $9 million of the $10.9 million direct costs. These three cost contributions are not dependent on any of the estimating factors provided in Table 8.2. Therefore, the manufacturing cost is generally insensitive to the estimating factors provided in Table 8.2. The use of the midrange values is acceptable for this situation.

8.2 Cost of Operating Labor

The technique used to estimate operating labor requirements is based on data obtained from five chemical companies and correlated by Alkayat and Gerrard [4]. According to this method, the operating labor requirement for chemical processing plants is given by Equation 8.3:

(8.3)

$$N_{OL} = (6.29 + 31.7P^2 + 0.23N_{np})^{0.5}$$

where N_{OL} is the number of operators per shift, P is the number of processing steps involving the handling of particulate solids—for example, transportation and distribution, particulate size control, and particulate removal. N_{np} is the number of nonparticulate processing steps and includes compression, heating and cooling, mixing, and reaction. In general, for the processes considered in this text, the value of P is zero, and the value of N_{np} is given by

(8.4)

$$N_{np} = \sum Equipment$$

compressors
towers
reactors
heaters
exchangers

Equation 8.3 was derived for processes with, at most, two solid handling steps. For processes with a greater number of solid handling operations, this equation should not be used.

The value of N_{OL} in Equation 8.3 is the number of operators required to run the process unit per shift. A single operator works on the average 49 weeks a year (3 weeks' time off for vacation and sick leave), five 8-hour shifts a week. This amounts to (49 weeks/year × 5 shifts/week) 245 shifts per operator per year. A chemical plant normally operates 24 hours/day. This requires (365 days/year × 3 shifts/day) 1095 operating shifts per year. The number of operators needed to provide this number of shifts is [(1095 shifts/yr)/(245 shifts/operator/yr)] or approximately 4.5 operators. Four and one-half operators are hired for each operator needed in the plant at any time. This provides the needed operating labor but does not include any support or supervisory staff.

To estimate the cost of operating labor, the average hourly wage of an operator is required. Chemical plant operators are relatively highly paid, and data from the Bureau of Labor and Statistics [5] give the hourly rate for miscellaneous plant and system operators in the Gulf Coast region at $26.48 in May 2006. This corresponds to $52,900 for a 2000-hour year. The cost of labor depends considerably on the location of the plant, and significant variations from the above figure may be expected. Historically, wage levels for chemical plant operators have grown slightly faster than the other cost indexes for process plant equipment given in Chapter 7. *The Oil and Gas Journal* and *Engineering News Record* provide appropriate indices to correct labor costs for inflation, or reference [5] can be consulted. The estimation of operating costs is illustrated in Example 8.2.

Example 8.2

Estimate the operating labor requirement and costs for the toluene hydrodealkylation facility shown in Figures 1.3 and 1.5.

From the PFD in Figure 1.5, the number and type of equipment are determined.

Using Equation (8.4), an estimate of the number of operators required per shift is made. This information is shown in Table E8.2.

Table E8.2 Results for the Estimation of Operating Labor Requirements for the Toluene Hydrodealkylation Process Using the Equipment Module Approach

Equipment Type	Number of Equipment	N_{np}
Compressors	1	1
Exchangers	7	7
Heaters/Furnaces	1	1
Pumps*	2	—
Reactors	1	1
Towers	1	1
Vessels*	4	—
	Total	11

*Pumps and vessels are not counted in evaluating N_{np} in Equation 8.4.

$$N_{OL} = [6.29 + (0)^{0.1} + (0.23)(11)]^{0.5} = [8.82]^{0.5} = 2.97$$

The number of operators required per shift = 2.97.

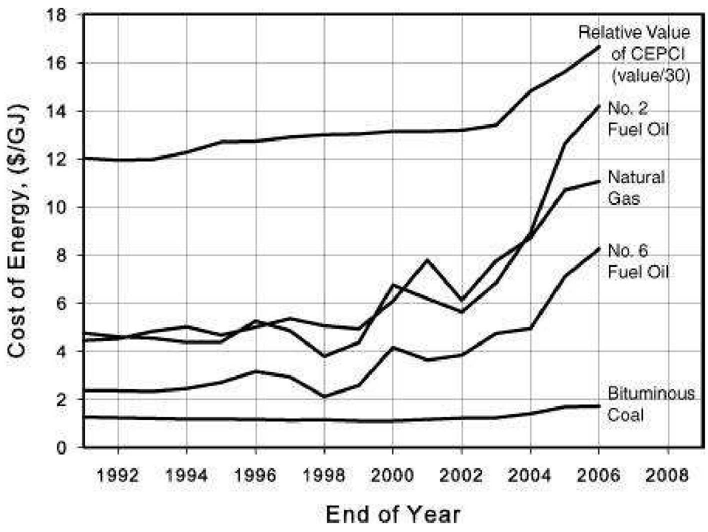
Operating Labor = (4.5)(2.97) = 13.4 (rounding up to the nearest integer yields 14 operators)

Labor Costs (2001) = 14 × $52,900 = $740,600/yr

8.3 Utility Costs

The costs of utilities are directly influenced by the cost of fuel. Specific difficulties emerge when estimating the cost of fuel, which directly impact the price of utilities such as electricity, steam, and thermal fluids. Figure 8.1 shows the general trends for fossil fuel costs from 1991 to 2006. The costs presented represent average values and are not site specific. These costs do not reflect the wide variability of cost and availability of various fuels throughout the United States.

Figure 8.1 Changes in Fuel Prices from 1991 to 2006 (Information taken from Energy Information Administration [6])



8.3.1 Background Information on Utilities

As seen from Figure 8.1, coal represents the lowest-cost fossil fuel on an energy basis. Most coal is consumed near the "mine mouth" in large power plants to produce electricity. The electricity is transported by power lines to the consumer. At locations remote from mines, both the availability and cost of transportation reduce and/or eliminate much of the cost advantage of coal. Coal suffers further from its negative environmental impact—for example, relatively high sulfur content and relatively high ratio of CO_2 produced per unit of energy.

After no. 6 fuel oil (a heavy oil with a relatively high sulfur content), the next lowest cost fuel source shown in Figure 8.1 is natural gas. Natural gas fuel is the least damaging fossil fuel energy supply with respect to the environment. It is transported by pipelines throughout much of the country. The cost is more uniform than coal throughout different regions of the country. There remain, however, regions in the country that are not yet serviced by the natural gas distribution system. In these regions, the use of natural gas is not an option that can be considered. Although natural gas is a mixture of several light hydrocarbons, it consists predominantly of methane. For the calculations used in this text, it is assumed that methane and natural gas are equivalent.

No. 2 fuel oil is the final fossil fuel that is commonly used as an energy source in the chemical industry. Until recently, it has been the highest-cost fossil fuel source. It is most readily available near coastal regions where oil enters the country and refining takes place. The United States has become increasingly dependent on imported oil, which may be subject to large upsets in cost and domestic availability. Uncertainties in the availability of supplies, high storage costs, and large fluctuations in cost make this source of energy least attractive in many situations. However, recently the cost of natural gas has increased substantially to the point that No. 2 fuel oil is now a viable alternative to natural gas in many plants.

Figure 8.1 shows that fuel costs have increased somewhat more rapidly and in a much more chaotic fashion than the cost index (CEPCI) that we have used previously to correct costs for inflation. As a result of the regional variations in the availability and costs of fossil fuels, along with the inability of the cost index to represent energy costs, we take the position that site-specific cost and availability information must be provided for a valid estimation of energy costs. We assume, in this text, that natural gas is the fuel of choice unless otherwise stated.

The PFD for the toluene hydrodealkylation process (Figure 1.5) represents the “battery-limits” plant. The equipment necessary to produce the various service or utility streams, which are used in the process and are necessary for the plant to operate, are not shown on the PFD. However, the utility streams such as cooling water and steam for heating are shown on the PFD. These streams, termed *utilities*, are necessary for the control of stream temperatures as required by the process. These utilities can be supplied in a number of ways.

1. **Purchasing from a Public or Private Utility:** In this situation no capital cost is involved, and the utility rates charged are based upon consumption. In addition the utility is delivered to the battery limits at known conditions.
2. **Supplied by the Company:** A comprehensive off-site facility provides the utility needs for many processes at a common location. In this case, the rates charged to a process unit reflect the fixed capital and the operating costs required to produce the utility.
3. **Self-Generated and Used by a Single Process Unit:** In this situation the capital cost for purchase and installation becomes part of the fixed capital cost of the process unit. Likewise the related operating costs for producing that particular utility are directly charged to the process unit.

Utilities that would likely be provided in a comprehensive chemical plant complex are shown in Table 8.3.

Table 8.3 Utilities Provided by Off-Sites for a Plant with Multiple Process Units (Costs Represent Charges for Utilities Delivered to the Battery Limit of a Process)

Utility	Description	Cost $/GJ	Cost $/Common Unit
Air Supply	Pressurized and dried air (add 20% for instrument air)		
	a. 6 barg (90 psig)		$0.49/100 std m^{3*}
	b. 3.3 barg (50 psig)		$0.35/100 std m^{3*}
Steam from Boilers	Process steam: latent heat only		
	a. Low pressure (5 barg, 160°C) from HP steam		
	With credit for power	13.28	$27.70/1000 kg
	Without credit for power	14.05	$29.29/1000 kg
	b. Medium pressure (10 barg, 184°C) from HP steam		
	With credit for power	14.19	$28.31/1000 kg
	Without credit for power	14.83	$29.59/1000 kg
	c. High pressure (41 barg, 254°C)	17.70	$29.97/1000 kg
Steam Generated from Process	Estimate savings as avoided cost of burning natural gas in boiler	12.33	
Cooling Tower Water	Processes cooling water: 30°C to 40°C or 45°C	0.354	$14.8/1000 m^{3†}
Other Water	High-purity water for		
	a. Process use		$0.067/1000 kg
	b. Boiler feed water (available at 115°C)‡		$2.45/1000 kg
	c. Potable (drinking)		$0.26/1000 kg
	d. Deionized water		$1.00/1000 kg
Electrical Substation	Electric Distribution	16.8	$0.06/kWh
	a. 110 V		
	b. 220 V		
	c. 440 V		
Fuels	a. Fuel oil (no. 2)	14.2	$549/m^3
	b. Natural gas	11.1§	$0.42/std m^{3*}
	c. Coal (f.o.b. mine mouth)	1.72	$41.4/tonne
Refrigeration	a. Moderately low temperature Refrigerated water in at T = 5°C and returned at 15°C	4.43	$0.185/1000kg
	b. Low temperature Refrigerant available at T = −20°C	7.89	
	c. Very low temperature Refrigerant available at T = −50°C	13.11	Based on process cooling duty
Thermal Systems	Cost based on thermal efficiency of fired heater using natural gas		
	a. 90% efficient	12.33	Based on process heating duty
	b. 80% efficient	13.88	
Waste Disposal (Solid and Liquid)	a. Nonhazardous		$36/tonne
	b. Hazardous		$200–2000/tonne°
Waste Water Treatment	a. Primary (filtration)		$41/1000 m^3
	b. Secondary (filtration + activated sludge)		$43/1000 m^3
	c. Tertiary (filtration, activated sludge, and chemical processing)		$56/1000 m^3

*Standard conditions are 1.013 bar and 15°C.

†Based on $\Delta T_{cooling\ water}$ = 10°C. Cooling water return temperatures should not exceed 45°C due to excess scaling at higher temperatures.

‡Approximately equal credit is given for condensate returned from exchangers using steam.

§Based on lower heating value of natural gas.

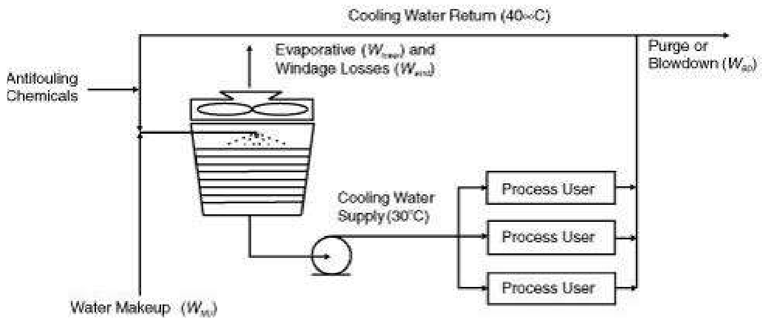
°For hazardous waste, the cost of disposal varies widely. Chemical analyses are required for all materials that cannot be thoroughly identified. This does not include radioactive waste.

8.3.2 Calculation of Utility Costs

The calculation of utility costs can be quite complicated, and the true cost of such streams is often difficult to estimate in a large facility. For estimating operating costs associated with supplying utilities to different processes, the approach taken here is to assume that the capital investment required to build a facility to supply the utility—for example, a cooling tower, a steam boiler, and so forth—has already been made. This would be the case when a grassroots cost has been used for the fixed capital investment. The costs associated with supplying a given utility are then obtained by calculating the operating costs to generate the utility. These are the costs that have been presented in Table 8.3, and the following sections show how these cost estimates were obtained for the major utilities given in the table.

Cooling Tower Water. In most large chemical, petrochemical, and refinery plants, cooling water is supplied to process units from a central facility. This facility consists of a cooling tower (or many towers), water makeup, chemical injection, and the cooling water feed pumps. A typical cooling water facility is shown in Figure 8.2.

Figure 8.2 Schematic Diagram of Cooling Water Loop



The cooling of the water occurs in the cooling tower where some of the water is evaporated. Adding makeup water to the circulating cooling water stream makes up this loss. Because essentially pure water is evaporated, there is a tendency for inorganic material to accumulate in the circulating loop; therefore, there is a water purge or blowdown from the system. The makeup water stream also accounts for windage or spray losses from the tower and also the water purge. Chemicals are added to reduce the tendency of the water to foul heat-exchanger surfaces within the processes. For a detailed discussion and further information regarding the conditioning of water for cooling towers, the reader is referred to Hile et al. [7] and Gibson [8]. From Figure 8.2, we can estimate the cost to supply process users with cooling water if the following are known:

- Total heat load and circulation rate required for process users
- Composition and saturation compositions of inorganic chemicals in the feed water
- Required chemical addition rate
- Desired supply and return temperatures (shown earlier to be 30°C and 40°C, respectively)
- Cost of cooling tower and cooling water pumps
- Costs of supply chemicals, electricity for pumps and cooling tower fans, and makeup water

The estimation of operating costs associated with a typical cooling water system is illustrated in Example

8.3.

Example 8.3

Estimate the utility cost for producing a circulating cooling water stream using a mechanical draft cooling tower. Consider a basis of 1 GJ/h of energy removal from the process units. Flow of cooling water required to remove this energy $= \dot{m}$ kg/h kg/h.

An energy balance gives

$$\dot{m}c_p\Delta T = 1 \times 10^9 \Rightarrow (\dot{m})(4180)(40 - 30) = 41{,}800\,\dot{m} = 1 \times 10^9 \text{ J/h}$$

Therefore, $\dot{m} = \dfrac{1 \times 10^9}{41{,}800} = 23{,}923 \text{ kg/h}$

Latent heat of water at average temperature of 35°C = 2417 kJ/kg

Amount of water evaporated from tower, W_{tower}

$$W_{tower} = \frac{Heat\ Load}{\Delta H_{vap}} = \frac{1 \times 10^9}{2417 \times 10^3} = 413.7 \text{ kg/h}$$

This is (413.7)(100)/(23,923) = 1.73% of the circulating water flowrate.

Typical windage losses from mechanical draft towers are between 0.1 and 0.3% [9, 10]; use 0.3%.

To calculate the blowdown, we must know the maximum allowable salt (inorganics) concentration factor, S, of the circulating water compared with the makeup water. The definition of S is given in the following equation:

$$S = \frac{\text{concentration salts in cooling water loop}}{\text{concentration salts in makeup water}} = \frac{S_{loop}}{S_{in}}$$

Typical values are between 3 and 7 [9]. Here a value of 5 is assumed. By performing a water and salt balance on the loop shown in Figure 8.2, the following results are obtained:

$$W_{MU} = W_{tower} + W_{wind} + W_{BD}$$

$$s_{in}W_{MU} = s_{loop}W_{wind} + s_{loop}W_{BD}$$

Because $s_{loop} = 5s_{in}$, it follows that

$$s_{in}(W_{tower} + W_{wind} + W_{BD}) = s_{loop}W_{wind} + s_{loop}W_{BD}$$

$$W_{BD} = \frac{s_{in}W_{tower} + W_{wind}(s_{in} - s_{loop})}{s_{loop} - s_{in}} = \frac{s_{in}W_{tower}}{s_{loop} - s_{in}} - W_{wind} = \frac{W_{tower}}{4} - W_{wind} = \frac{1.73\%}{4} - 0.3\% = 0.133\%$$

$$W_{MU} = 1.73 + 0.3 + 0.133 = 2.163\% = 517\text{kg/h}$$

Pressure drop around the cooling water loop is estimated as follows: ΔP_{loop} = 15 psi (pipe losses) + 5 psi (exchanger losses) + 10 psi (control valve loss) + 8.7 psi of static head (because water must be pumped to top of cooling water tower, estimated to be 20 ft above pump inlet) = 38.7 psi = 266.7 kPa.

Power required for cooling water pumps with a volumetric flow rate $\dot{V}$, assuming an overall efficiency of 75%, is

$$\text{Pump Power} = \frac{1}{\varepsilon}\dot{V}\Delta P = \frac{1}{(0.75)}\frac{(23{,}923)}{(1000)(3600)}(266.7) = 2.36\text{ kW}$$

Power required for fans: From reference [11], the required surface area in the tower = 0.5 ft^2/gpm (this assumes that the design wet-bulb air temperature is 26.7°C [80°F]). From the same reference, the fan horsepower per square foot of tower area is 0.041 hp/ft^2.

$$\text{Power for fan} = \frac{(23{,}923)(2.2048)}{(60)(8.337)}(0.5)(0.041) = (2.16)(0.746) = 1.61\text{ kW}$$

From a survey of vendors, the cost of chemicals is \$0.156/1000 kg of makeup water.

Using an electricity cost of \$0.06/kWh and a process water cost of \$0.067/1000 kg, the overall cost of the cooling water is given by

Cost of cooling water = cost of electricity + cost of chemicals for makeup water + cost of makeup water

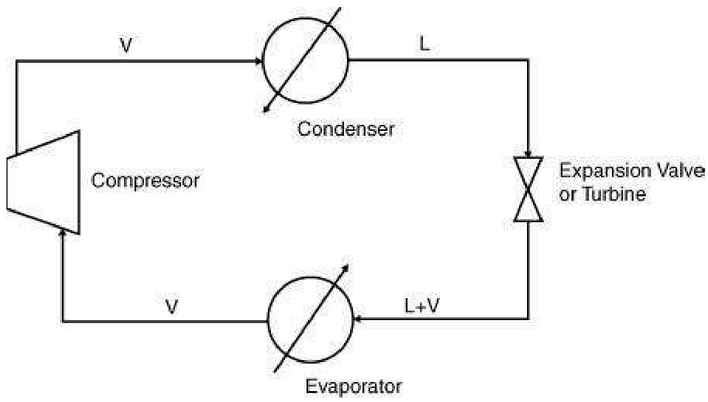
Using the cost values for electricity and process water given in Table 8.3,

$$\text{Cooling water cost} = (0.06)(2.36 + 1.61) + \frac{(517.3)(0.156)}{1000} + \frac{(517.3)(0.067)}{1000}$$
$$= \$0.354/\text{hr} = \$0.354/\text{GJ}$$

Clearly, this cost will change depending on the cost of electricity, the cost of chemicals, and the cost of process water.

Refrigeration. The basic refrigeration cycle consists of circulating a working fluid around a loop consisting of a compressor, evaporator, expansion valve or turbine, and condenser. This cycle is shown in Figure 8.3. The phases of the working fluid (L-liquid and V-vapor) are shown on the diagram.

Figure 8.3 Process Flow Diagram for a Simple Refrigeration Cycle



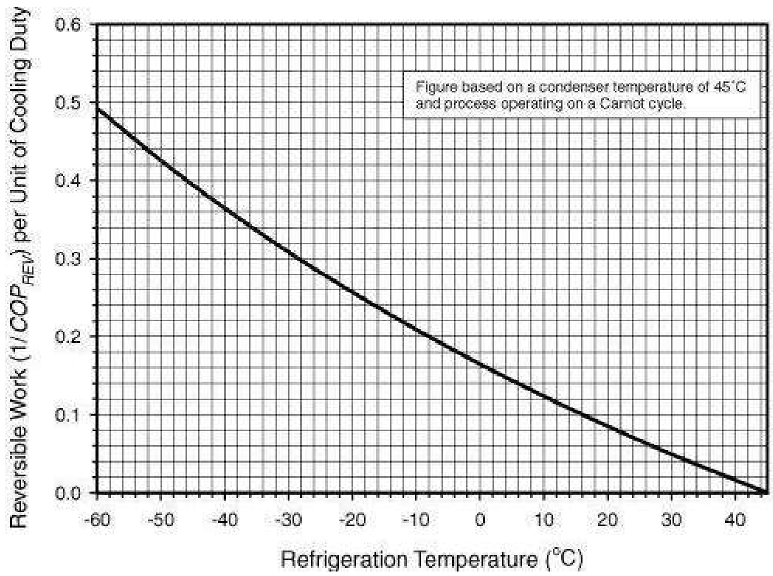
The Carnot efficiency of a mechanical refrigeration system can be expressed by the reversible coefficient of performance, COP_{REV}:

$$COP_{REV} = \frac{\text{evaporator temperature } (T_1)}{\text{temperature difference between condenser and evaporator } (T_2 - T_1)}$$

$$COP \cong \frac{\text{evaporator heat load}}{\text{work required}} \quad \text{or} \quad \text{work required} = \frac{\text{evaporator heat load}}{\text{COP}}$$

Because all the processes for a Carnot engine must be reversible, the COP_{REV} gives the best theoretical performance of a refrigeration system. Thus the net required power (compressor-expansion turbine) will always be greater than that predicted by the equation above using COP_{REV}. Nevertheless, it is clear that as the temperature difference between the evaporator and condenser increases then the work required per unit of energy removed in the evaporator (refrigerator) increases. Therefore, the operating costs for refrigeration will increase as the temperature at which the refrigeration is required decreases. The condensation of the working fluid will most often be achieved using cooling water, so a reasonable condensing temperature would be 45°C (giving a 5°C approach in the condensing exchanger). Figure 8.4 illustrates the effect of the evaporator temperature on the reversible work required for a given cooling load. This figure gives an approximate guide to the relative cost of refrigeration. The relative costs of refrigeration at different temperatures are explored in Example 8.4.

Figure 8.4 Ideal Work for Refrigeration Cycles as a Function of Refrigeration Temperature



Example 8.4

Using Figure 8.4, calculate the relative costs of providing refrigeration at 5°C, –20°C, and –50°C. From the figure, the ordinate values are given as follows.

Temperature	1/COP_{REV}
5°C	0.144
–20°C	0.257
–50°C	0.426

Therefore, compared with cooling at 5°C, cooling to –20°C is 0.257/0.144 times as expensive, and cooling to –50°C is 0.426/0.144 times as expensive. This analysis assumes that the two refrigeration systems operate equally efficiently with respect to the reversible limit and that the major cost is the power to run the compressors.

In Example 8.5, a real refrigeration system is considered and operating costs are estimated.

Example 8.5

Obtain a cost estimate for a refrigerated cooling utility operating at 5°C.

Consider a single-stage refrigeration system to provide refrigeration at 5°C, using 1,1 difluoroethane (R-152a) as the refrigerant. The process flow diagram and operating conditions are given in Figure E8.5 and Table E8.5, respectively.

Figure E8.5 Process Flow Diagram for Simple Refrigeration Cycle of Example 8.5

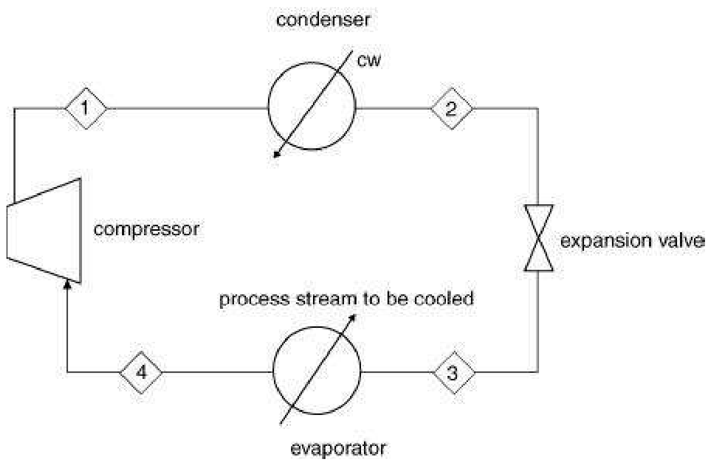


Table E8.5 Stream Conditions for Figure E8.5				
	Stream Number			
Condition	**1**	**2**	**3**	**4**
Pressure (bar)	10.9	10.9	3.21	3.21
Temperature (°C)	68.7	45.0	5.0	5.0
Vapor Fraction	1.0	0.0	0.2492	1.0

For the simulation shown, pressure drops across piping and heat exchangers have not been considered. When the circulation rate of R-152a is 65.3 kmol/h, the duty of the evaporator is 1 GJ/h. The compressor is assumed to be 75% efficient and the loads on the equipment are as follows:

Compressor Power = 66.5 kW (at 75% efficiency)

Condenser Duty = 1.24 GJ/h

Evaporator Duty = 1.00 GJ/h

Compressor work per unit of cooling = (66.5)/(1,000,000/3600) = 0.2394

This value compares with 0.144 for the Carnot cycle. The main differences are due to the inefficiencies in the compressor and the use of a throttling valve instead of a turbine.

The cost of refrigeration at 5°C = (66.5)(0.06) + (1.24)(0.354) = 3.99 + 0.44 = 4.43 $/h = 4.43 $/GJ

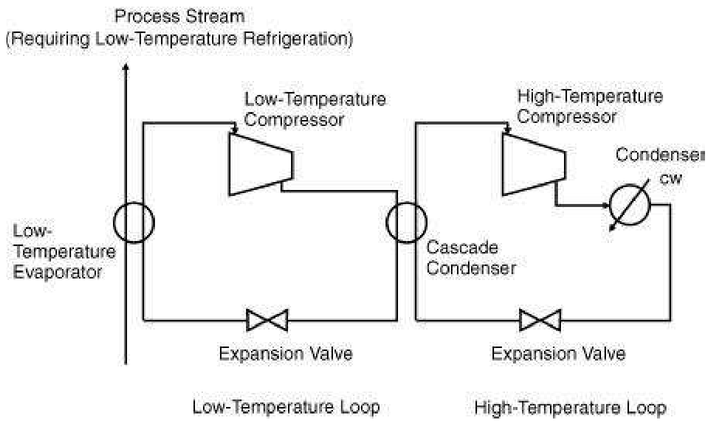
Using the results of Example 6.4, we can predict the cost of refrigeration at –20°C and –50°C as

The cost of refrigeration at –20°C = (4.43)(1.78) = $7.89/GJ

The cost of refrigeration at –50°C = (4.43)(2.96) = $13.11/GJ

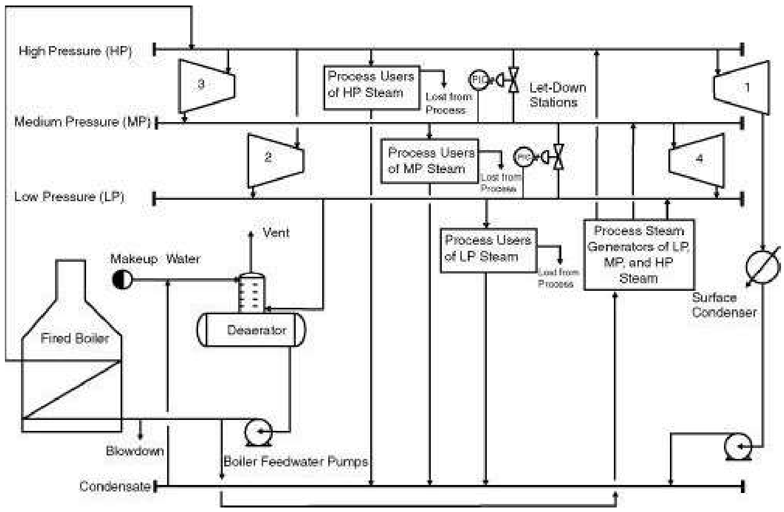
For refrigeration systems operating at less than temperatures of approximately –60°C, the simple refrigeration cycle shown in Figures 8.4 and E8.5 is no longer applicable. The main reason for this is that there are no common refrigerants that can be liquified at 45°C under reasonable pressures (not excessively high) and still give the desired low temperature in the condenser also at reasonable pressures (not excessively low). For these low-temperature systems, some form of cascaded refrigeration system is required. In such systems, two working fluids are used. The primary fluid provides cooling to the process (at the lowest temperature) and rejects heat to the secondary working fluid that rejects its heat to cooling water at 45°C. A simplified diagram of a cascaded refrigeration system is shown in Figure 8.5.

Figure 8.5 Schematic Diagram of a Simple Cascaded Refrigeration System



Steam Production. Steam is produced by the evaporation and superheating of specially treated water. The fuel that is used to supply the energy to produce steam is by far the major operating expense. However, water treatment costs can be substantial depending on the supply water composition and the degree of recovery of condensed steam in process heat exchangers. As shown in Table 8.3, for large chemical plants, steam is often required at several different pressure levels. However, it is often generated at the highest level and then let down to the lower pressure levels through turbines. These turbines produce electricity used in the plant. A typical steam generating facility is shown in Figure 8.6. Because there are losses of steam in the system due to leaks and, more important, due to process users not returning condensate, there is a need to add makeup water. This water is filtered to remove particulates and then treated to reduce the hardness. The latter can be achieved by the addition of chemicals to precipitate magnesium and calcium salts followed by filtration. These salts have **reverse solubility** characteristics and therefore precipitate at high temperatures. Alternatively, an ion-exchange system can be employed. The solids-free, "soft" water is now fed to the steam generating system. The thorough treatment of the water is necessary, because any contaminants entering with the water will ultimately deposit on heat-exchanger surfaces and boiler tubes and cause fouling and other damage. Another important issue is the dissolved oxygen and carbon dioxide that enter with the makeup water. These dissolved gases must be removed in order to eliminate (reduce) corrosion of metal surfaces in the plant. The removal occurs in the deaerator, in which the makeup water is scrubbed with steam to de-gas the water. Oxygen scavengers are also added to the circulating condensate to remove any trace amounts of oxygen in the system. Amines may also be added to the water in order to neutralize any residual carbonic acid formed from dissolved carbon dioxide. Finally, blowdown of water from the water storage tank (situated near the boiler) is necessary to remove any heavy sludge and light solids that are picked up as steam and condensate circulate through the system [12]. The problems associated with the buildup of chemicals become even more troublesome in high-pressure (>66 bar) boilers, and several solutions are discussed by Wolfe [13].

Figure 8.6 Typical Steam Producing System for a Large Chemical Facility



In order to estimate accurately steam generation costs, it is necessary to complete a steam balance on the plant. An algorithm for carrying out a steam balance for a new facility is listed below.

1. Determine the pressure levels for the steam in the plant. These are usually set at around 41.0 barg (600 psig), between 10.0 barg (150 psig) and 15.5 barg (225 psig), and between 3.4 barg (50 psig) and 6.1 barg (90 psig).
2. Determine the total number of process users of the different levels of steam. These numbers become the basis for the steam balance.
3. Determine which of the above users will return condensate to the boiler feed water (BFW) system. Note: If live steam injection is required for the process, there will be no condensate returned from this service. In addition, for some small users, condensate return may not be economical.
4. Determine the condensate-return header pressure.
5. Estimate the blowdown losses.
6. Complete a balance on the steam and condensate, and determine the required water makeup to the steam system.
7. Determine the steam generating capacity of the steam boiler. The logic used here is that all steam will be generated at the highest-pressure level and will be let down either through turbines or let-down stations (valves) to the medium- and low-pressure headers. The high-pressure steam is often generated at 44.3 barg (650 psig) to allow for frictional losses and superheated to 400°C (752°F) to produce more efficient power production in the turbines.
8. Additional power generation may be accomplished by running turbines using the high-pressure steam, by using surface condensers (operating at the cooling water temperature), and by running turbines between the medium- and low-pressure steam headers. All these options are shown in Figure 8.6. In order to balance a plant's electrical and steam needs, the determination of the correct amount of steam to generate is an iterative process.

Clearly the algorithm can become quite complicated. In order to determine a reasonable value or cost for the different steam levels, the approach used here is to assume that all the steam will be generated at the highest pressure level and then let down to the appropriate pressure level through turbines or let-down stations (valves). In the former case, credit will be taken for generating power; in the latter case, credit will not be taken. The procedure for calculating the cost of steam at different pressure levels is given in Example 8.6.

Example 8.6

Determine the cost of producing high-, medium-, and low-pressure steam using a natural gas fuel source. For medium- and low-pressure steam production, assume that steam is produced at the highest pressure level, and consider both the case when this steam is sent through a turbine to make electricity and when it is simply throttled through a valve.

Again the approach taken here is to assume that the fixed capital investment associated with the initial purchase of the steam generation facilities has been accounted for elsewhere. The analysis given below accounts only for the operating costs associated with steam (and power) production. The source of fuel is assumed to be natural gas that costs $11.10/GJ. See Table 8.3.

High-Pressure Steam (41.0 barg)

Basis is 1000 kg of HP steam generated at 45.3 bar and 400°C $\Rightarrow h_{44.3\ barg,\ 400°C}$ = 3204.3 kJ/kg.

Conditions at the header are 41 bar saturated (T_{sat} = 254°C). Note that the steam is generated at a higher pressure and superheated for more efficient expansion, but that desuperheating will be assumed at the process user.

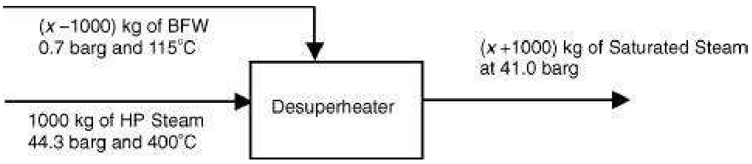
Assume boiler feed water comes from a deaerator that operates at exhaust steam header pressure of 0.7 barg and T_{sat} = 115°C (10 psig) $\Rightarrow h_{BFW}$ = 483.0 kJ/kg.

$$\Delta H_{BFW\text{-}HP\ Steam} = (3204.3 - 483.0) = 2721.3 \text{ kJ/kg}$$

$$\text{Energy required to produce HP steam} = (2721.3)(1000) = 2.721 \text{ GJ}$$

Because this HP steam is superheated, we can produce more than 1000 kg of saturated steam from it. In order to desuperheat this steam, BFW is added to produce saturated steam at 41.0 barg (h = 2797.6 kJ/kg). See Figure E8.6.

Figure E8.6 Sketch of Desuperheating Process for HP Steam



Let x be the amount of saturated HP steam produced from superheated HP steam; an enthalpy balance gives $(1000)(3204.3) + (x - 1000)(483) = (x)(2797.6) \Rightarrow x = 1175.7$ kg.

The cost of natural gas to produce 1000 kg of sat HP steam (assuming a 90% boiler efficiency) is given by

$$\text{Cost} = \frac{(2.721)}{(0.9)}\frac{(1000)(11.1)}{(1175.7)} = \$28.54$$

Treatment costs for circulating boiler feed water = \$0.15/1000 kg for oxygen scavengers, and so on (average from several vendors).

Boiler feed water cost is based on the assumption that 10% makeup is required.

Cost of electricity to power air blowers supplying combustion air to boiler:

Natural gas usage = 2.721/0.9/1.1757 = 2.572 GJ = (2.572)(6)/(0.23)= 67.1 std m^3 = 67.1/22.4 = 2.99 kmol

Oxygen usage (based on 3% excess over stoichiometric) = (2.99)(2)(1.03) = 6.17 kmol oxygen

This comes from (6.17)/(0.21) = 29.38 kmol of air.

Assume that this air must be raised 0.5 bar to overcome frictional losses in boiler and stack, and assuming that the blower is 60% efficient. Therefore,

The electrical usage for blower is 14 kWh/1000 kg of steam produced, giving an electricity cost = (14)(0.06) = \$0.84.

The cost of BFW is based on the water makeup, treatment chemicals, and the thermal energy in the stream.

For a basis of 1000 kg of BFW,

Cost of makeup water = \$0.067

Cost of chemicals for treatment = \$0.15

Energy in BFW $= \dot{m}c_p\Delta T = (1000)(4.18)(115 - 25) = 0.376\ \text{GJ}$

Value of energy = (\$11.10)(0.376) = \$ 4.17

BFW cost = 4.17 + 0.067 + 0.15 = \$4.39/1000 kg

Cost of BFW makeup = (0.1)(4.39) = \$0.439

Total cost of HP steam = \$28.54 + \$0.15 + \$0.84 + \$0.439 = \$29.97/1000 kg

Medium-Pressure Steam (10.0 barg)

It is assumed that letting steam down through a turbine from the high-pressure header to the medium-pressure header will generate electrical power.

The theoretical steam requirement (kg steam/kWh) for this situation is found by assuming an isentropic expansion of the steam from the HP condition to the medium-pressure level. From the steam tables, we have the following information:

$$h_{44.3\ barg,\ 400°C} = 3204.3 \text{kJ/kg and } s_{44.3\ barg,\ 400°C} = 6.6953 \text{ kJ/kg K}$$

By interpolating at constant specific entropy, we get that the outlet temperature is 212°C and the outlet enthalpy = 2851.0 kJ/kg.

$$\Delta h = (3204.3 - 2851.0) = 353.3 \text{ kJ/kg} = \text{theoretical work}$$

Therefore, 1000 kg of HP steam produces 353.3 MJ or 98.14 kWh of electricity. Assuming a turbine efficiency of 75%, the output power is (0.75)(98.14) = 73.6 kWh.

$$\text{Credit for electricity} = (73.6)(0.06) = \$4.42$$

The actual outlet enthalpy of the steam is 3,204.3 – (353.3)(0.75) = 2939.3 kJ/kg. This is still superheated steam. Desuperheating the steam to 10.3 barg and saturated conditions (h = 2779.1 kJ/kg) will generate x kg of MP steam from the 1000 kg of HP steam, where

$$(1000)(2939.3) + (x - 1000)(483.0) = (x)(2779.1) \Rightarrow x = 1069.8 \text{ kg}$$

Therefore, the cost of natural gas to produce 1,000 kg of MP steam (assuming a 90% boiler efficiency) is

$$\text{Cost} = \frac{(2.721)}{(0.9)} \frac{(1000)(11.10)}{(1069.8)} = \$31.37$$

We can find the cost of electricity for the blower by using the ratio of the natural gas usage from the high-pressure steam case. Therefore,

$$\text{cost of electricity} = \frac{(31.37)}{(28.54)}(0.84) = \$0.92$$

Total cost of MP steam (with power production) = \$31.37 – \$4.42 + \$0.92 + \$0.439 = \$28.31/1000 kg

For the case when power production is not implemented, the HP steam is throttled to the pressure of the MP header through a let-down station, which is essentially an irreversible, isentropic process through a valve. The superheated steam is then desuperheated at the process user.

Enthalpy of HP steam (at 44.8 barg and 400°C) = 3204.3 kJ/kg

Enthalpy of saturated MP steam = 2779.1 kJ/kg

Enthalpy of BFW = 483.0 kJ/kg

If x is the amount of saturated MP steam obtained by desuperheating, then an enthalpy balance gives

$$(1000)(3204.3) + (x - 1000)(483) = (x)(2779.1) \Rightarrow x = 1185.2 \text{ kg}$$

Cost of natural gas to produce 1000 kg of sat HP steam (assuming a 90% boiler efficiency) is

$$\text{Cost} = \frac{(2.721)}{(0.9)}\frac{(1000)(11.10)}{(1185.2)} = \$28.32$$

$$\text{Cost of electricity for the air blower} = \frac{(28.32)}{(28.54)}(0.84) = \$0.83$$

Total cost of MP steam (without power production) = \$28.32 + \$0.83 + \$0.439
= \$29.59/1000 kg

Note: This is almost identical to the cost for HP steam.

Low-Pressure Steam (5.2 barg)

The calculation procedures for evaluating the cost of low-pressure steam are identical to those given above for medium-pressure steam and the results are given below.

Total cost of LP steam (with power production) = \$32.25 – \$5.94 + \$0.95 + \$0.439
= \$27.70/1000 kg

Total cost of LP steam (without power production) = \$28.03 + \$0.82 + \$0.439 = \$29.29/1000 kg

Waste Heat Boilers

When steam is generated from within the process—in a waste heat boiler, for example—the savings to the process are usually calculated from the avoided cost of using an equivalent amount of natural gas in the boiler system. If we assume that the boiler efficiency is 90%, then for every GJ of energy saved by producing steam within a process unit, the boiler facility saves (\$11.1)/(0.9) = \$12.33 in natural gas costs.

Hot Circulating Heat-Transfer Fluids. Again, the greatest cost for these systems is the fuel that is burned to heat the circulating heat-transfer fluid. Typical efficiencies (based on the lower heating value, LHV, of the fuel) for these heaters range from 60% to 82% [1]. With air preheating economizers, the efficiency can be as high as 90%. Example 8.7 illustrates the use of efficiencies in fired heaters.

Example 8.7

Estimate the utility cost of a heat-transfer medium heated in a fired heater using natural gas as the fuel.

Assuming that the heat-transfer medium is heated in a process heater that is 80% efficient and uses natural gas at \$11.10/GJ as the fuel source, we get

$$\text{Cost of 1 GJ of energy} = (1)(11.10)/(0.80) = \$\ 13.88/\text{GJ}$$

Assuming a 90% efficient heater, we get

$$\text{Cost of 1 GJ of energy} = (1)(11.10)/(0.9) = \$\ 12.33/\text{GJ}$$

8.4 Raw Material Costs

The cost of raw materials can be estimated by using the current price listed in such publications as the *Chemical Market Reporter* (*CMR*) [14]. A list of common chemicals and their selling price, as of August 2006, are given in Table 8.4. Current raw material and product chemical prices may be obtained from the current issue of the *CMR* [14]. To locate costs for individual items, it is not sufficient to look solely at the current issue, because not all chemicals are listed in each issue. It is necessary to explore several of the most recent issues. In addition, for certain chemicals large seasonal price fluctuations may exist, and it may be advisable to look at the average price over a period of several months.

Table 8.4: Costs of Some Common Chemicals*

Chemical	Cost ($/kg)	Typical Shipping Capacity or Basis for Price
Acetaldehyde	1.003	Railroad Tank Cars
Acetic Acid	1.090 (2004)	Railroad Tank Cars
Acetone (MMA grade)	0.948	Railroad Tank Cars
Acrylic Acid	1.929	Railroad Tank Cars
Allyl Chloride	1.80**	F.O.B. Gulf Coast
Benzene	0.657	Barge, Gulf Coast
Chlorine	0.375	Railroad Tank Car
Dimethyl Ether	0.948† (Jan. 2000)	Railroad Tank Car
Ethanol (190 Proof)	0.937	Railroad Tank Car
Ethylbenzene	1.069	Railroad Tank Car, Gulf Coast
Ethylene	1.202	Contract
Ethylene Oxide	1.764	Railroad Tank Car
Formaladehyde/Formalin (37 wt%)		
No-inhibitor	0.838	Railroad Tank Car, Gulf Coast
7% Methanol Inhibitor	0.463	Railroad Tank Car, Gulf Coast
Hydrochloric Acid (23YBe)	0.095	Railroad Tank Car, Works
Iso-Butylene	0.706	F.O.B. Works
Iso-Propanol (99%)	1.378	Railroad Tank Car
Maleic Anhydride	1.543	Railroad Tank Car
Methanol	0.294	F.O.B. Gulf Coast
Methyl Ethyl Ketone	1.598	Railroad Tank Car
MTBE	0.687	Barge, Gulf Coast
Propylene		
(Polymer Grade)	1.014	F.O.B. Gulf Coast
(Chemical Grade)	0.981	F.O.B. Gulf Coast
Styrene	1.543	F.O.B. Works
Sulfur (Crude)	0.043	Railroad Car
Sulfuric Acid (virgin)	0.090	Railroad Tank Car, Gulf Coast
Toluene	0.648	Barge, Gulf Coast
Mixed Xylenes	0.608	Barge, Gulf Coast
Ortho-Xylene	0.805	Railroad Tank Cars
Para-Xylene	1.135	Railroad Tank Cars
Meta-Xylene	2.910	Railroad Tank Cars

*Unless stated otherwise these are average values from http://www.icis.com/StaticPages/a-e.htm#top, August 2006.
**Vendor quote.
†From *CMR*, January 2000.

Another factor that is sometimes overlooked is that often companies will lock onto a selling or purchase price through a short- or long-term contract. Such contracts will often yield prices that are significantly lower than those given in the *CMR*. In addition, in doing economic evaluations for different chemical processes, the purchase and selling price for chemicals will not always be available from the *CMR*. For example, in January 2001, *CMR* stopped publishing the price of dimethyl ether. Likewise, prices for allyl alcohol have not been published for several years. The prices shown in Table 8.4 were obtained from manufacturers' quotes. When doing economic evaluations for new, existing, or future plants, it is advisable to establish the true selling or purchase price for all raw materials and products. Because the largest operating cost is nearly always the cost of raw materials, it is important to obtain accurate prices if realistic economic evaluations are to be obtained.

8.5 Yearly Costs and Stream Factors

Manufacturing and associated costs are most often reported in terms of $/yr. Information on a PFD is most often reported in terms of kg or kmol per hour or per second. In order to calculate the yearly cost of raw materials or utilities, the fraction of time that the plant is operating in a year must be known. This fraction is known as the **stream factor** (SF), where

(8.5)

$$\text{Stream Factor (SF)} = \frac{\text{Number of Days Plant Operates per Year}}{365}$$

Typical values of the stream factor are in the range of 0.96 to 0.90. Even the most reliable and well-managed plants will typically shut down for two weeks a year for scheduled maintenance, giving an SF = 0.96. Less reliable processes may require more downtime and hence lower SF values. The stream factor represents the fraction of time that the process unit is on-line and operating at design capacity. When estimating the size of equipment, care must be taken to use the design flowrate for a typical stream day and not a calendar day. Example 8.8 illustrates the use of the stream factor.

Example 8.8

- **a.** Determine the yearly cost of toluene for the process given in Chapter 1.
- **b.** What is the yearly consumption of toluene?
- **c.** What is the yearly revenue from the sale of benzene?

Assume a stream factor of 0.95, and note that the flowrates given on the PFD are in kilograms per stream hour.

From Table 1.5, flowrate of toluene = 10,000 kg/h (Stream 1)
From Table 1.5, flowrate of benzene = 8210 kg/h (Stream 15)
From Table 8.5, cost of toluene = $0.648/kg

Table 8.5 Theoretical Steam Requirements (kg steam/kWh)

	Inlet Pressure of Steam (barg) (Superheat in °C)							
	10.0 (sat'd)	13.8 (sat'd)	17.2 50	27.6 170	41.4 145	41.4 185	58.6 165	58.6 205
Exhaust Pressure								
2″ Hg abs	4.77	4.54	4.11	3.34	3.22	3.07	2.98	2.85
4″ Hg abs	5.33	5.04	4.54	3.62	3.47	3.30	3.20	3.05
0 barg	8.79	7.94	6.88	5.08	4.72	4.45	4.22	4.00
0.69 barg	10.87	9.57	8.11	5.77	5.28	4.97	4.67	4.40
2.07 barg	15.24	12.72	10.40	6.91	6.18	5.78	5.35	5.02
3.45 barg	20.86	16.32	12.79	7.97	6.97	6.49	5.93	5.54
4.14 barg	24.45	18.32	14.11	8.50	7.34	6.83	6.20	5.78
4.82 barg	28.80	20.68	15.47	9.05	7.71	7.16	6.45	6.01

From Perry, R. H., and D. W. Green, *Perry's Chemical Engineering Handbook*, 6th ed., McGraw-Hill, New York, NY, 1984. Reprinted by permission of The McGraw-Hill Companies.

From Table 8.5, cost of benzene = $0.657/kg

a. Yearly cost of toluene = (24)(365)(10,000)(0.648)(0.95) = $53,927,000/yr.
b. Yearly consumption of toluene = (24)(365)(10,000)(0.95) / 1000 = 83,200 tonne/yr.
c. Yearly revenue from benzene sales = (24)(365)(8210)(0.657)(0.95) = $44,889,000/yr.

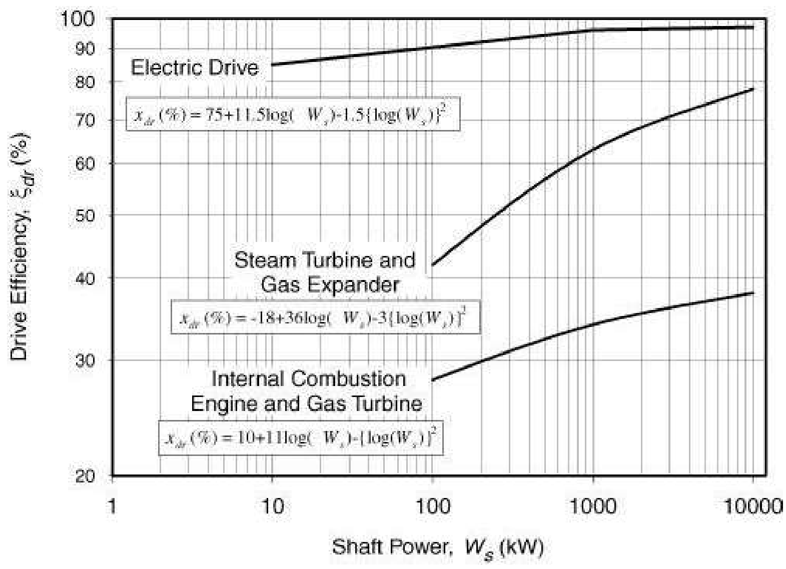
Comparing the results from Parts (a) and (c), we can see that with the current prices for these two chemicals it is not economical to produce benzene from toluene. Historically, the price differential between benzene and toluene has been greater than the $0.009/kg shown in Table 8.4, and this is the reason why this process has been used, and is currently being used, to produce benzene. Clearly, if this low price differential were to exist for a long period of time, this process might have to be shut down.

8.6 Estimating Utility Costs from the PFD

Most often, utilities do not directly contact process streams. Instead, they exchange heat energy (fuel gas, steam, cooling water, and boiler feed water) in equipment such as heat exchangers and process heaters, or they supply work (electric power or steam) to pumps, compressors, and other rotating equipment. In most cases, the flowrate can be found either by inspection or by doing a simple heat balance around the equipment.

Steam can be used to drive a piece of rotating equipment such as a compressor. In this case, both the theoretical steam requirement and efficiency are required. Table 8.5 provides the theoretical steam requirements as a function of the steam inlet pressure and the exhaust pressure for steam turbine drives. The mechanical efficiencies of different drives are shown in Figure 8.7, using data from Walas [9].

Figure 8.7 Efficiencies for Pumps and Compressor Drives (Data from Walas [8], Chapter 4)



To illustrate the techniques used to estimate the utility flowrates and utility costs for various types of equipment, see Example 8.9.

Example 8.9

Estimate the quantities and yearly costs of the appropriate utilities for the following pieces of equipment on the toluene hydrodealkylation PFD (Figure 1.5). It is assumed that the stream factor is 0.95 and that all the numbers on the PFD are on a stream time basis. The duty on all of the units can be found in Table 1.7.

- **a.** E-101, Feed Preheater
- **b.** E-102, Reactor Effluent Cooler
- **c.** H-101, Heater
- **d.** C-101, Recycle Gas Compressor, assuming electric drive
- **e.** C-101, Recycle Gas Compressor, assuming steam drive using 10 barg steam discharging to atmospheric pressure.
- **f.** P-101, Toluene Feed Pump

Solution

a. E-101: Duty is 15.19 GJ/h. From Table 8.3, Cost of High-Pressure Steam = \$17.70/GJ

$$\text{Energy Balance: } Q = 15.19\ \text{GJ/h} = (\dot{m}_{steam})(\Delta H_{vap}) = (\dot{m}_{steam})(1699.3)\ \text{kJ/kg}$$

$$\dot{m}_{steam} = 8939\ \text{kg/h} = 2.48\ \text{kg/s}$$

Yearly Cost = $(Q)(C_{steam})(t)$ = (15.19 GJ/h)(\$17.70/GJ)(24)(365)(0.95) = \$ 2,237,000/yr

Alternatively, Yearly Cost = (Yearly flowrate)(Cost per unit mass)

Yearly Cost = (2.48)(3600)(24)(365)(0.95)(29.97/1000) = \$2,227,000/yr
(same as above within round-off error)

b. E-102: Duty is 46.66 GJ/h. From Table 8.3, Cost of Cooling Water = \$0.354/GJ

$$Q = 46.66\ \text{GJ/h} = (\dot{m}_{cw})(C_{pcw})(\Delta T_{cw}) = (\dot{m}_{cw})(4.18)(10) = 41.8\,\dot{m}_{cw}$$

$$\dot{m}_{cw} = (46.66)(10^9/41.8)(10^3) = 1{,}116{,}270\ \text{kg/h} = 310\ \text{kg/s}$$

Yearly Cost = (46.66 GJ/h)(24)(365)(0.95)(\$0.354/GJ) = \$137,000/yr

c. H-101: Duty is 27 GJ/h (7510 kW). Assume that an indirect, nonreactive process heater has a thermal efficiency (ξ_{th}) of 90%. From Table 8.3, natural gas costs \$11.10/GJ, and the heating value is 0.0377 GJ/m^3.

$$Q = 27\ \text{GJ/h} = (\dot{v}_{gas})(\Delta H_{natural\ gas})(\text{efficiency}) = (\dot{v}_{gas})(0.0377)(0.9)$$

$$\dot{v}_{gas} = 796\ \text{std m}^3/\text{h}\ (0.22\ \text{std m}^3/\text{sec})$$

Yearly Cost = (27)(11.10)(24)(365)(0.95)/(0.90) = \$2,771,000/yr

d. C-101: Shaft power is 49.1 kW, and from Figure 8.7 the efficiency of an electric drive (ξ_{dr}) is 90%.

$$\text{Electric Power} = P_{dr} = \text{Output power}/\xi_{dr} = (49.1)/(0.90) = 54.6 \text{ kW}$$

$$\text{Yearly Cost} = (54.6)(0.06)(24)(365)(0.95) = \$27{,}200/\text{yr}$$

e. Same as Part (d) with steam driven compressor. For 10 barg steam with exhaust at 0 barg, Table 8.5 provides a steam requirement of 8.79 kg steam/kWh of power. The shaft efficiency is about 35% (extrapolating from Figure 8.7).

$$\text{Steam required by drive} = (49.1)(8.79/0.35) = 1233 \text{ kg/h } (0.34 \text{ kg/s})$$

$$\text{Cost of Steam} = (1233)(24)(365)(0.95)(28.32 \times 10^{-3}) = \$290{,}600/\text{yr}$$

f. **P-101:** Shaft power is 14.2 kW. From Figure 8.7 the efficiency of an electric drive is about 86%.

$$\text{Electric Power} = 14.2/0.86 = 16.5 \text{ kW}$$

$$\text{Yearly Cost} = (16.5)(0.06)(24)(365)(0.95) = \$8240/\text{yr}$$

Note: The cost of using steam to power the compressor is much greater than the cost of electricity even though the cost per unit energy is much lower for the steam. The reasons for this are (1) the thermodynamic efficiency is low, and (2) the efficiency of the drive is low for a small compressor. Usually steam drives are used only for compressor duties greater than 100 kW.

8.7 Cost of Treating Liquid and Solid Waste Streams

As environmental regulations continue to tighten, the problems and costs associated with the treatment of waste chemical streams will increase. In recent years there has been a trend to try to reduce or eliminate the volume of these streams through waste minimization strategies. Such strategies involve utilizing alternative process technology or using additional recovery steps in order to reduce or eliminate waste streams. Although waste minimization will become increasingly important in the future, the need to treat waste streams will continue. Some typical costs associated with this treatment are given in Table 8.3, and flowrates can be obtained from the PFD. It is worth noting that the costs associated with the disposal of solid waste streams, especially hazardous wastes, have grown immensely in the past few years, and the values given in Table 8.3 are only approximate average numbers. Escalation of these costs should be done with extreme caution.

8.8 Evaluation of Cost of Manufacture for the Production of Benzene via the Hydrodealkylation of Toluene

The cost of manufacture for the production of benzene via the toluene HDA process is given in Example 8.10.

Example 8.10

Calculate the cost of manufacture without depreciation (COM_d) for the toluene hydrodealkylation process using the PFD in Figure 1.5 and the flow table given in Table 1.5.

A utility summary for all the equipment is given in Table E8.10, from which we find the total yearly utility costs for this process:

Steam = $ 3,412,000/yr

Cooling Water = $ 165,000/yr

Fuel Gas = $2,771,000/yr

Electricity = $37,400/yr

Total Utilities = $6,385,000/yr

Raw Material Costs from the PFD, Table 8.4, and Example 8.8 are

Toluene = $53,927,000/yr

Hydrogen = $6,622,000/yr (based on a value of $0.118/std m^3)

Total Raw Materials = $60,549,000/yr

There are no waste streams shown on the PFD, so

Waste Treatment = $0.0/yr

From Example 8.2 the cost of operating labor is

Table E8.10 Summary of Utility Requirements for the Equipment in the Toluene Hydrodealkylation Process

Equipment	Electric Power (kW)	Steam High-Pressure (kg/s)	Steam Med-Pressure (kg/s)	Steam Low-Pressure (kg/s)	Cooling Water (m^3/s)	Fuel Gas (std m^3/s)
E-101	—	2.48	—	—	—	—
E-102	—	—	—	—	0.31	—
E-103	—	—	—	0.14	—	—
E-104	—	—	—	—	0.055	—
E-105	—	—	—	—	0.007	—
E-106	—	—	1.26	—	—	—
H-101	—	—	—	—	—	0.22
C-101	54.5	—	—	—	—	—
P-101	16.5	—	—	—	—	—
P-102	4.0	—	—	—	—	—
Totals	75.0	2.48	1.26	0.14	0.372	0.22
Total yearly cost $/yr	37,400	2,227,000	1,069,000	116,000	165,000	2,771,000

Data from Figure 1.5, Table 1.7, and Example 8.9.

$\boldsymbol{C_{OL}}$ **= (14)(52,900) = $741,000/yr**

From Problem 7.21 (using CAPCOST), we find that the fixed capital investment (C_{GR}) for the process is $ 11.7 × 10^6.

FCI = $ 11.7 × $\mathbf{10^6}$

Finally, using Equation 8.2, the total manufacturing cost is estimated to be

$COM_d = 0.180FCI_L + 2.73C_{OL} + 1.23(\text{Utilities} + \text{Raw Materials} + \text{Waste Treatment})$

$COM_d = (0.180)(11.7 \times 10^6) + 2.73\ (741{,}000) + 1.23\ (6{,}385{,}000 + 60{,}549{,}000 + 0)$

$COM_d = \$\ 86.46 \times 10^6/\text{yr}$

8.9 Summary

In this chapter, the cost of manufacturing for a chemical process was shown to depend on the fixed capital investment, the cost of operating labor, the cost of utilities, the cost of waste treatment, and the cost of raw materials. In most cases, the cost of raw materials is the biggest cost. Methods to evaluate these different costs were discussed. Specifically, the amount of the raw materials and utilities can be obtained directly from the PFD. The cost of operating labor can be estimated from the number of pieces of equipment given on the PFD. Finally, the fixed capital investment can again be estimated from the PFD using the techniques given in Chapter 7.

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Short Answer Questions

1. In the general equation for determining the cost of manufacturing (COM_d), Equation (8.2), one of the terms is 0.180*FCI*, where *FCI* is the fixed capital investment of the plant. “This term is included to cover the interest payment on the loan for the plant.” Is this statement true or false? Explain your answer.

2. Why is the number of operators per shift multiplied by approximately 4.5 to obtain the total number of operators required to run the plant?

3. What is a stream factor?

4. When estimating the cost of manufacturing (COM_d) for a chemical process, the overall COM_d may be estimated using only five individual costs. Lists these five costs.

5. Cooling water is priced on an energy basis: $/GJ. We usually assume the temperature rise to be 10°C. Does the cooling water cost change if the return temperature changes? Are there any limitations to the return temperature? Explain.

6. In Equation (8.2), the cost of raw materials, C_{RM}, is multiplied by a factor of 1.23. The reason for this is that, in general, we expect that the estimated cost of raw materials will be about 20% low and we add a correction factor of 1.23 to adjust for this. Do you agree with this explanation? If you do not,

give another reason for using the factor of 1.23.

7. In Equation (8.2), the cost of operating labor, C_{OL}, is multiplied by a factor of 2.73. One reason for this is that the value of C_{OL} includes only plant operators and not supervisory and clerical labor costs. Is this statement true or false? What other factors (if any) account for the multiplication factor of 2.73?

8. Explain the difference between direct costs, fixed costs, and general expenses. Give two examples of each.

Problems

9. You are employed at a chemical company and have recently been transferred from a plant that manufactures synthetic dyes to a new facility that makes specialty additives for the polymer resin industry.

   a. You have been asked to estimate the cost of manufacturing at this new facility. Would you
      i. Use Equation (8.2) to estimate COM_d?
      ii. Use data from the old plant where you worked, because you are very familiar with all the aspects of manufacturing for that process?
      iii. Dig up information on the new process and use these figures?

   b. When would you use a relationship such as Equation (8.2)?

10. When a chemical plant needs steam at multiple pressure levels, it is often economical to generate all the steam at a high pressure and then to let the steam down through pressure-reducing turbines to the desired pressure. This principle is illustrated in Figure 8.6. The downside of this approach is that as the exhaust pressure of the turbine increases, the theoretical (and actual) steam requirements increase, meaning that less energy is extracted. To illustrate this point, do the following.

   a. Estimate the amount of energy extracted from 10,000 kg/h of 58.6 barg steam superheated by 165°C when connected to the following turbines (each 80% efficient).
      i. Exhaust pressure is 4” Hg absolute.
      ii. Exhaust pressure is 4.82 barg.

   b. Estimate the amount of energy extracted from 10,000 kg/h of saturated, 10.0 barg steam when connected to a turbine (80% efficient) exhausting at 4.82 barg.

   c. Identify the locations of each of the three turbines described above on Figure 8.6.

11. What are the operating costs associated with a typical cooling water system? Based on the example given in this chapter, answer the following.

   a. What percentage of the operating costs is the makeup water?

   b. By how much would the cost of cooling water increase if the cost of power (electricity) were to double?

c. By how much would the cost of cooling water increase if the cost of make-up water were to double?

12. Determine the cost of producing a refrigerant stream at –50°C using propane as the working fluid in a noncascaded system. You may wish to refer to Example 8.5 to do this problem. The steps you should follow are as follows.

a. Determine the pressure at which propane can be condensed at 45°C, which assumes that cooling water with a 5°C temperature approach will be used as the condensing medium.

b. Determine the pressure to which the propane must be throttled in order to liquefy it at –50°C.

c. Use the results of Parts (a) and (b) to set the approximate pressure levels in the condenser and evaporator in the refrigeration system.

d. Determine the amount of propane necessary to extract 1 GJ of heat in the evaporator.

e. Assuming a 5kPa pressure drop in both heat exchangers and that a single-stage compressor is used with an efficiency of 75%, determine the cost of electricity to run the compressor, determine the cooling water cost, and from this determine the cost of providing refrigeration at –50°C using propane as the working fluid.

13. Repeat the process described in Problem 12 using a simple refrigeration loop to determine the cost of providing 1 GJ of refrigeration at –50°C using the following working fluids:

a. Propylene

b. Ethane

c. Ammonia

Determine whether any of the working fluids given above cannot be used in a simple (noncascaded) refrigeration loop. For these fluids, would using a cascaded refrigeration system to provide –50°C refrigerant make sense? Explain carefully your answers to these questions.

14. Estimate the cost of operating labor (C_{OL}), the cost of utilities (C_{UT}), and the cost of manufacturing (COM_d) for the ethylbenzene process given in Project B.2 of Appendix B. You must do Problem 7.21 in order to estimate COM_d.

15. Estimate the cost of operating labor (C_{OL}), the cost of utilities (C_{UT}), and the cost of manufacturing (COM_d) for the styrene process given in Project B.3 of Appendix B. You must do Problem 7.22 in order to estimate COM_d.

16. Estimate the cost of operating labor (C_{OL}), the cost of utilities (C_{UT}), and the cost of manufacturing (COM_d) for the drying oil process given in Project B.4 of Appendix B. You must do Problem 7.23 in

order to estimate COM_d.

17. Estimate the cost of operating labor (C_{OL}), the cost of utilities (C_{UT}), and the cost of manufacturing (COM_d) for the maleic anhydride process given in Project B.5 of Appendix B. You must do Problem 7.24 in order to estimate COM_d.

18. Estimate the cost of operating labor (C_{OL}), the cost of utilities (C_{UT}), and the cost of manufacturing (COM_d) for the ethylene oxide process given in Project B.6 of Appendix B. You must do Problem 7.25 in order to estimate COM_d.

19. Estimate the cost of operating labor (C_{OL}), the cost of utilities (C_{UT}), and the cost of manufacturing (COM_d) for the formalin process given in Project B.7 of Appendix B. You must do Problem 7.26 in order to estimate COM_d.

Appendix A Cost Equations and Curves for the CAPCOST Program

The purpose of this appendix is to present the equations and figures that describe the relationships used in the capital equipment-costing program CAPCOST introduced in Chapter 7 and used throughout the text. The program is based on the module factor approach to costing that was originally introduced by Guthrie [1, 2] and modified by Ulrich [3].

A.1 Purchased Equipment Costs

All the data for the purchased cost of equipment for the second edition of this book were obtained from a survey of equipment manufacturers during the period May to September of 2001, so an average value of the CEPCI of 397 over this period should be used when accounting for inflation.

Additional process equipment has been added to the third edition and is listed below:

- Conveyors
- Crystallizers
- Dryers
- Dust Collectors
- Filters
- Mixers
- Reactors
- Screens

The purchased costs for these types of equipment were obtained in 2003 but the costs given here have been normalized to 2001. For this new equipment, bare module factors were not available, nor were pressure factors or materials of construction factors. In general, these units are generally bought as a package, and installation in the plant is not expensive. The bare module factors for these units are taken to be the field installation factors given by Guthrie [1, 2].

Data for the purchased cost of the equipment, at ambient operating pressure and using carbon steel construction, C_p^o, were fitted to the following equation:

(A.1)

$$\log_{10} C_p^o = K_1 + K_2 \log_{10}(A) + K_3[\log_{10}(A)]^2$$

where A is the capacity or size parameter for the equipment. The data for K_1, K_2, and K_3, along with the maximum and minimum values used in the correlation, are given in Table A.1. These data are also presented in the form of graphs in Figures A.1–A.17. It should be noted that in these figures, the data are plotted as C_p^o/A as a function of size attribute, A. This form of the graph clearly illustrates the decreasing cost per unit of capacity as the size of the equipment increases.

Table A.1 Equipment Cost Data to Be Used with Equation A.1

Equipment Type	Equipment Description	K_1	K_2	K_3	Capacity, Units	Min Size	Max Size
Blenders	Kneader	5.0141	−0.4133	0.3224	Volume, m^3	0.14	3
	Ribbon	4.1366	−0.4928	0.0070	Volume, m^3	0.7	11
	Rotary	4.1366	−0.4928	0.0070	Volume, m^3	0.7	11
Centrifuges	Auto batch separator	4.7681	−0.0260	0.0240	Diameter, m	0.5	1.7
	Centrifugal separator	4.3612	−0.1236	−0.0049	Diameter, m	0.5	1
	Oscillating screen	4.8600	−0.6660	0.1063	Diameter, m	0.5	1.1
	Solid bowl w/o motor	4.9697	0.1689	0.0038	Diameter, m	0.3	2
Compressors	Centrifugal, axial, and reciprocating	2.2897	1.3604	−0.1027	Fluid power, kW	450	3000
	Rotary	5.0355	−1.8002	0.8253	Fluid power, kW	18	950
Conveyors	Apron	3.9255	−0.4961	0.1506	Area, m^2	1.0	15
	Belt	4.0637	−0.7416	0.1550	Area, m^2	0.5	325
	Pneumatic	4.6616	−0.6795	0.0638	Area, m^2	0.75	65
	Screw	3.6062	−0.7341	0.1982	Area, m^2	0.5	30
Crystallizers	Batch	4.5097	−0.8269	0.1344	Volume, m^3	1.5	30
Drives	Gas turbine	−21.7702	13.2175	−1.5279	Shaft power, kW	7500	23,000
	Intern comb. engine	2.7635	0.8574	−0.0098	Shaft power, kW	10	10,000
	Steam turbine	2.6259	1.4398	−0.1776	Shaft power, kW	70	7500
	Electric—explosion-proof	2.4604	1.4191	−0.1798	Shaft power, kW	75	2600
	Electric—totally enclosed	1.9560	1.7142	−0.2282	Shaft power, kW	75	2600
	Electric—open/drip-proof	2.9508	1.0688	−0.1315	Shaft power, kW	75	2600
Dryers	Drum	4.5472	−0.7269	0.1340	Area, m^2	0.5	50
	Rotary, gas fired	3.5645	0.1118	−0.0777	Area, m^2	5	100
	Tray	3.6951	−0.4558	−0.1248	Area, m^2	1.8	20
Dust Collectors	Baghouse	4.5007	−0.5818	0.0813	Volume, m^3	0.08	350
	Cyclone scrubbers	3.6298	−0.4991	0.0411	Volume, m^3	0.06	200
	Electrostatic precipitator	3.6298	−0.4991	0.0411	Volume, m^3	0.06	200
	Venturi scrubber	3.6298	−0.4991	0.0411	Volume, m^3	0.06	200
Evaporators	Forced circulation (pumped)	5.0238	0.3475	0.0703	Area, m^2	5	1000
	Falling film	3.9119	0.8627	−0.0088	Area, m^2	50	500
	Agitated film (scraped wall)	5.0000	0.1490	−0.0134	Area, m^2	0.5	5
	Short tube	5.2366	−0.6572	0.3500	Area, m^2	10	100
	Long tube	4.6420	0.3698	0.0025	Area, m^2	100	10,000
Fans	Centrifugal radial	3.5391	−0.3533	0.4477	Gas flowrate, m^3/s	1	100
	Backward curve	3.3471	−0.0734	0.3090	Gas flowrate, m^3/s	1	100
	Axial vane	3.1761	−0.1373	0.3414	Gas flowrate, m^3/s	1	100
	Axial tube	3.0414	−0.3375	0.4722	Gas flowrate, m^3/s	1	100
Filters	Bent	5.1055	−0.5001	0.0001	Area, m^2	0.9	115
	Cartridge	3.2107	−0.2403	0.0027	Area, m^2	15	200
	Disc and drum	4.8123	−0.7142	0.0420	Area, m^2	0.9	300
	Gravity	4.2756	−0.6480	0.0714	Area, m^2	0.5	80
	Leaf	3.8187	−0.3765	0.0176	Area, m^2	0.6	235
	Pan	4.8123	−0.7142	0.0420	Area, m^2	0.9	300
	Plate and frame	4.2756	−0.6480	0.0714	Area, m^2	0.5	80
	Table	5.1055	−0.5001	0.0001	Area, m^2	0.9	115
	Tube	5.1055	−0.5001	0.0001	Area, m^2	0.9	115
Furnaces	Reformer furnace	3.0680	0.6597	0.0194	Duty, kW	3000	100,000
	Pyrolysis furnace	2.3859	0.9721	−0.0206	Duty, kW	3000	100,000
	Nonreactive fired heater	7.3488	−1.1666	0.2028	Duty, kW	1000	100,000

(continued)

Equipment Type	Equipment Description	K_1	K_2	K_3	Capacity, Units	Min Size	Max Size
Heat exchangers	Scraped wall	3.7803	0.8569	0.0349	Area, m^2	2	20
	Teflon tube	3.8062	0.8924	−0.1671	Area, m^2	1	10
	Bayonet	4.2768	−0.0495	0.1431	Area, m^2	10	1000
	Floating head	4.8306	−0.8509	0.3187	Area, m^2	10	1000
	Fixed tube	4.3247	−0.3030	0.1634	Area, m^2	10	1000
	U-tube	4.1884	−0.2503	0.1974	Area, m^2	10	1000
	Kettle reboiler	4.4646	−0.5277	0.3955	Area, m^2	10	100
	Double pipe	3.3444	0.2745	−0.0472	Area, m^2	1	10
	Multiple pipe	2.7652	0.7282	0.0783	Area, m^2	10	100
	Flat plate	4.6656	−0.1557	0.1547	Area, m^2	10	1000
	Spiral plate	4.6561	−0.2947	0.2207	Area, m^2	1	100
	Air cooler	4.0336	0.2341	0.0497	Area, m^2	10	10000
	Spiral tube	3.9912	0.0668	0.2430	Area, m^2	1	100
Heaters	Diphenyl heater	2.2628	0.8581	0.0003	Duty, kW	650	10750
	Molten salt heater	1.1979	1.4782	−0.0958	Duty, kW	650	10750
	Hot water heater	2.0829	0.9074	−0.0243	Duty, kW	650	10750
	Steam boiler	6.9617	−1.4800	0.3161	Duty, kW	1200	9400
Mixers	Impeller	3.8511	−0.2991	−0.0003	Power, kW	5	150
	Propeller	4.3207	−0.9641	0.1346	Power, kW	5	500
	Turbine	3.4092	−0.5104	0.0030	Power, kW	5	150
Packing	Loose (for towers)	2.4493	0.9744	0.0055	Volume, m^3	0.03	628
Process vessels	Horizontal	3.5565	0.3776	0.0905	Volume, m^3	0.1	628
	Vertical	3.4974	0.4485	0.1074	Volume, m^3	0.3	520
Pumps	Reciprocating	3.8696	0.3161	0.1220	Shaft power, kW	0.1	200
	Positive displacement	3.4771	0.1350	0.1438	Shaft power, kW	1	100
	Centrifugal	3.3892	0.0536	0.1538	Shaft power, kW	1	300

(continued)

Equipment Type	Equipment Description	K_1	K_2	K_3	Capacity, Units	Min Size	Max Size
Reactors	Autoclave	4.5587	−0.7014	0.0020	Volume, m^3	1	15
	Fermenter	4.1052	−0.4680	−0.0005	Volume, m^3	0.1	35
	Inoculum tank	3.7957	−0.5407	0.0160	Volume, m^3	0.07	1
	Jacketed agitated	4.1052	−0.4680	−0.0005	Volume, m^3	0.1	35
	Jacketed nonagitated	3.3496	−0.2765	0.0025	Volume, m^3	5	45
	Mixer/settler	4.7116	−0.5521	0.0004	Volume, m^3	0.04	60
Screens	DSM	3.8050	−0.4144	0.2120	Area, m^2	0.3	6
	Rotary	4.0485	−0.8882	0.3260	Area, m^2	0.3	15
	Stationary	3.8219	0.0368	−0.6050	Area, m^2	2	11
	Vibrating	4.0485	−0.8882	0.3260	Area, m^2	0.3	15
Towers	Tray and packed	3.4974	0.4485	0.1074	Volume, m^3	0.3	520
Tanks	API—fixed roof	4.8509	−0.3973	0.1445	Volume, m^3	90	30000
	API—floating roof	5.9567	−0.7585	0.1749	Volume, m^3	1000	40000
Trays	Sieve	2.9949	0.4465	0.3961	Area, m^2	0.07	12.30
	Valve	3.3322	0.4838	0.3434	Area, m^2	0.70	10.50
	Demisters	3.2353	0.4838	0.3434	Area, m^2	0.70	10.50
Turbines	Axial gas turbines	2.7051	1.4398	−0.1776	Fluid power, kW	100	4000
	Radial gas/liquid expanders	2.2476	1.4965	−0.1618	Fluid power, kW	100	1500
Vaporizers	Internal coils/jackets	4.0000	0.4321	0.1700	Volume, m^3	1	100
	Jacketed vessels	3.8751	0.3328	0.1901	Volume, m^3	1	100

Figure A.1 Purchased Costs for Compressors and Drives (Cost Data for Compressors and Drives Taken from R-Books Software by Richardson Engineering Services, Inc. [4])

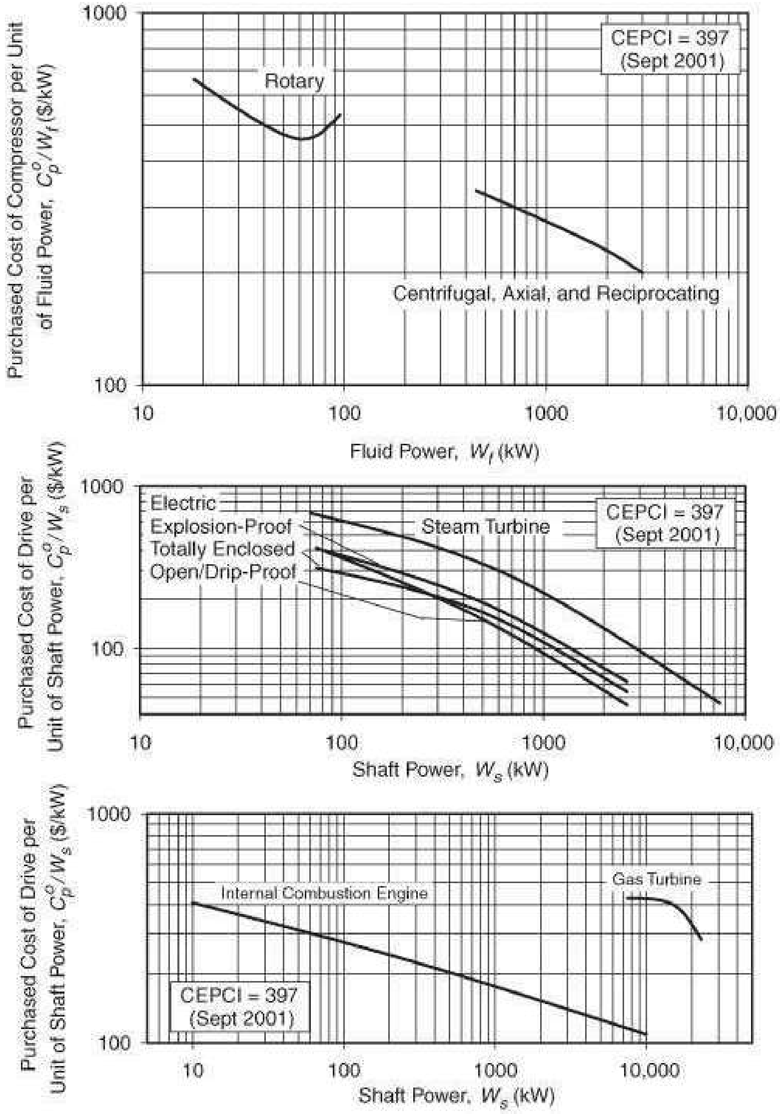


Figure A.2 Purchased Costs for Evaporators and Vaporizers

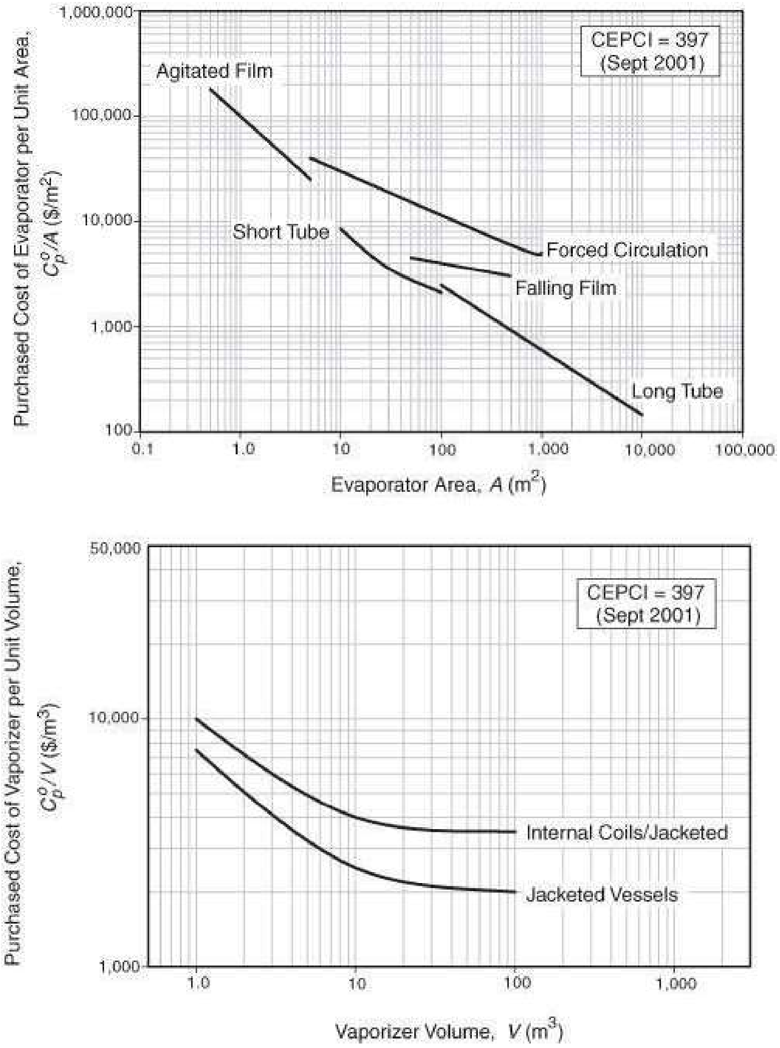


Figure A.3 Purchased Costs for Fans, Pumps, and Power Recovery Equipment (Cost Data for Fans Taken from R-Books Software by Richardson Engineering services [4])

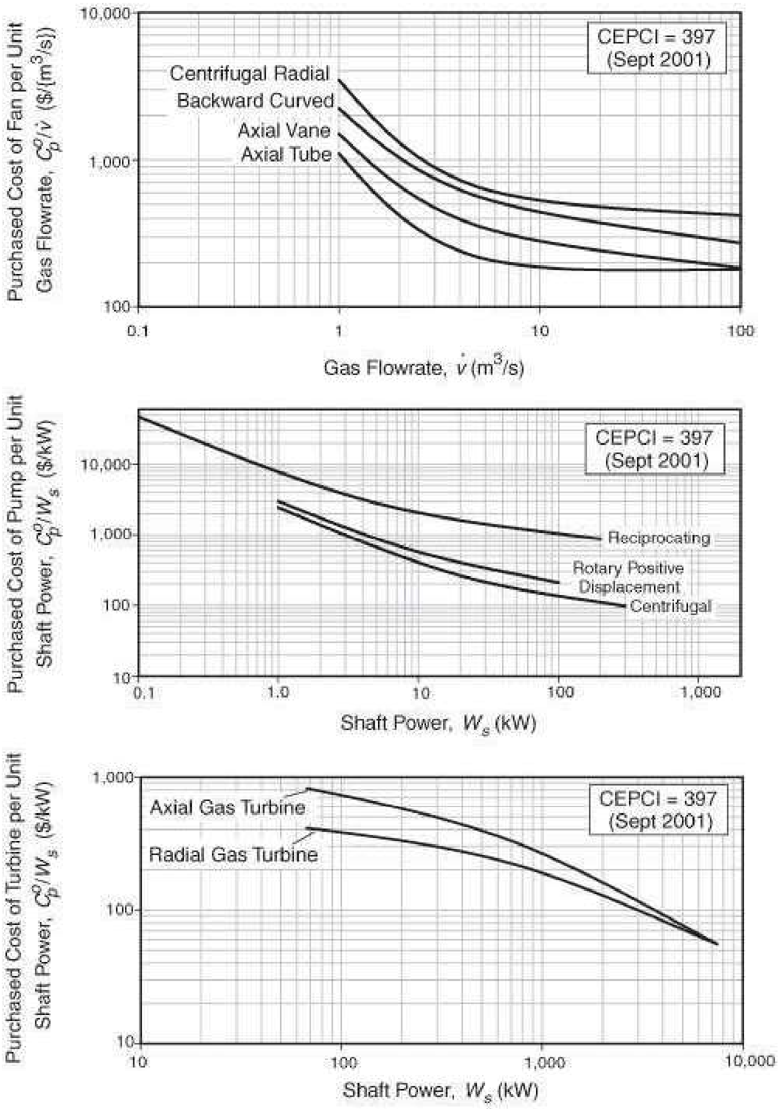


Figure A.4 Purchased Costs for Fired Heaters and Furnaces

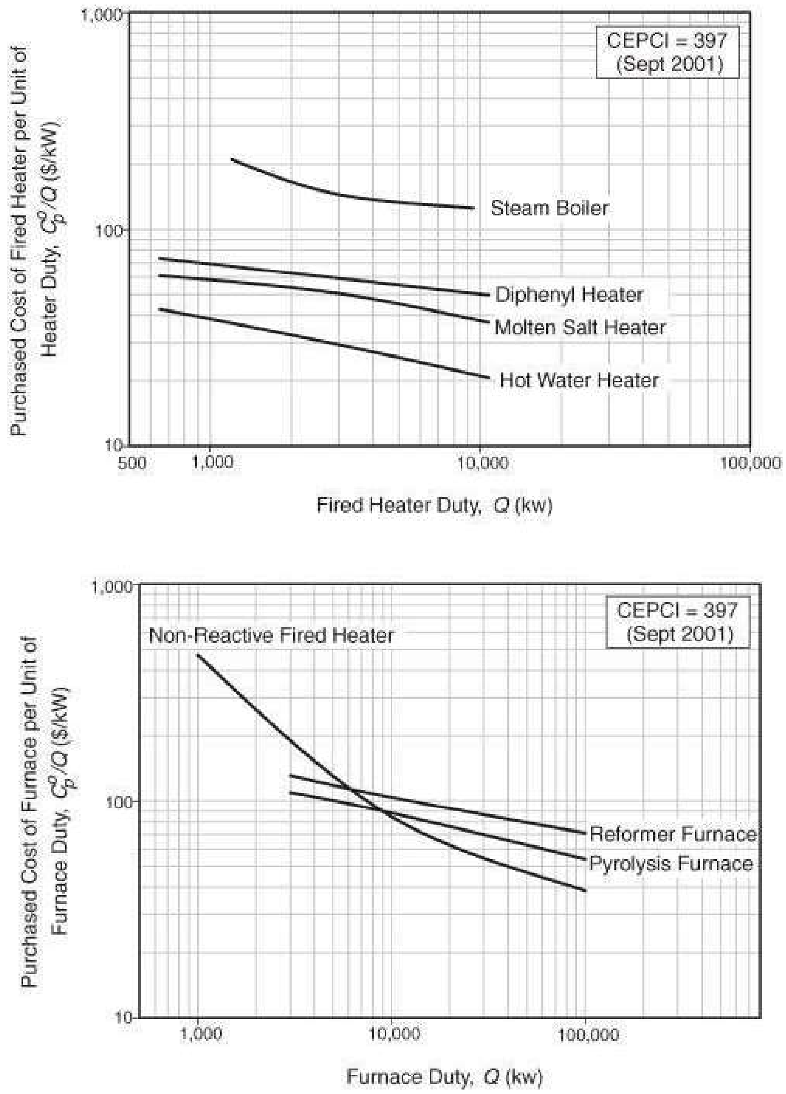


Figure A.5 Purchased Costs for Heat Exchangers

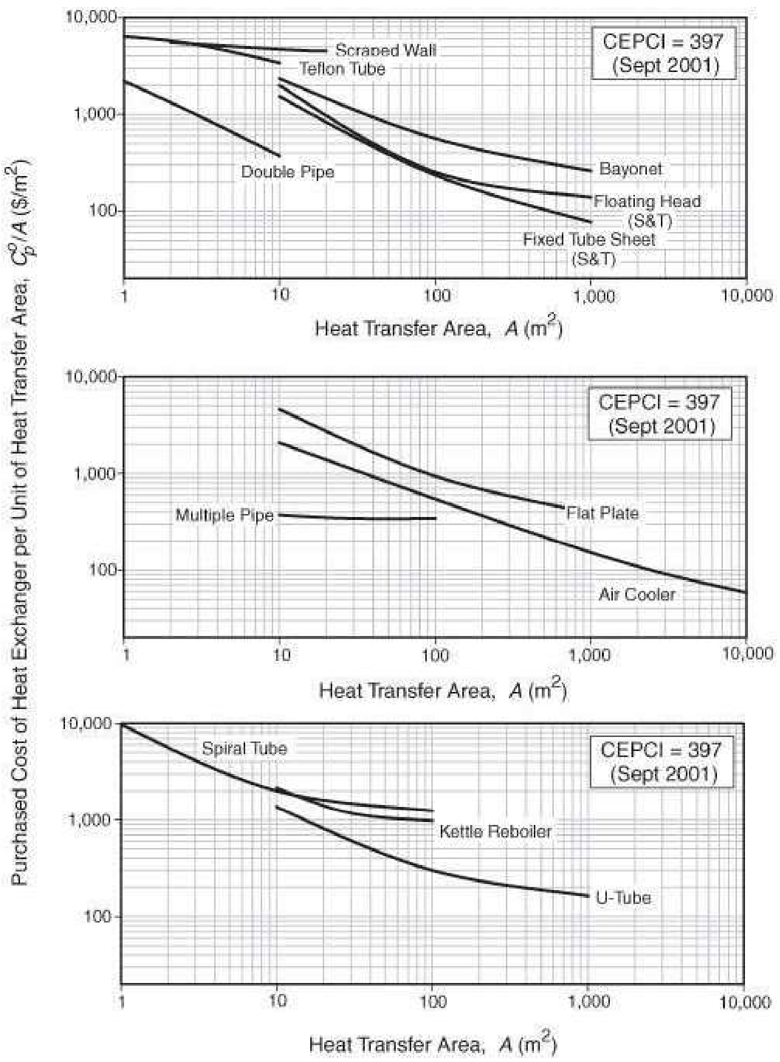


Figure A.6 Purchased Costs for Packing, Trays, and Demisters

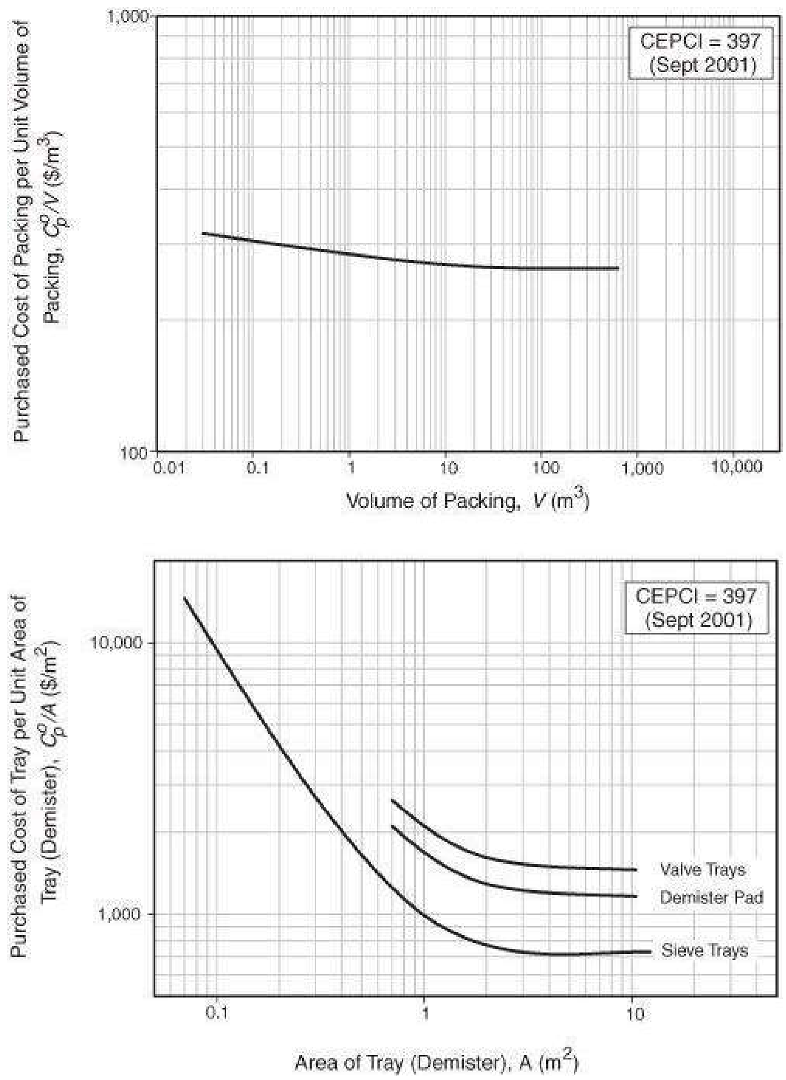
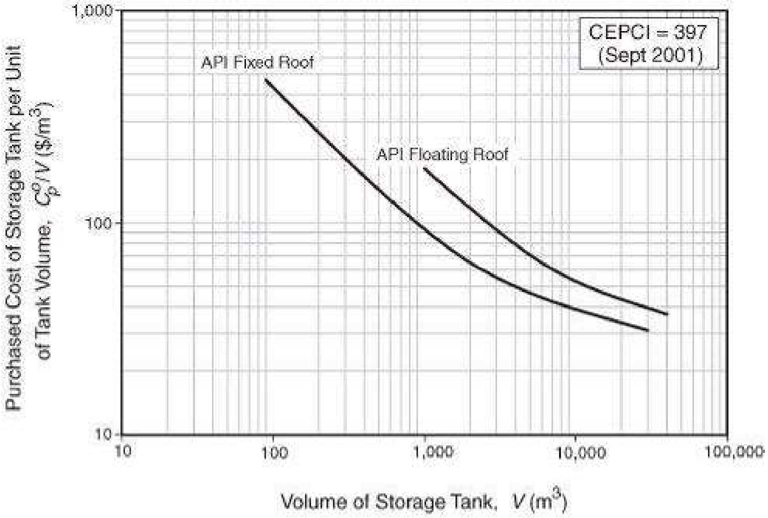


Figure A.7 Purchased Costs of Storage Tank and Process Vessels. (Data for Storage Tanks Taken from R-Books Software by Richardson Engineering Services [4])



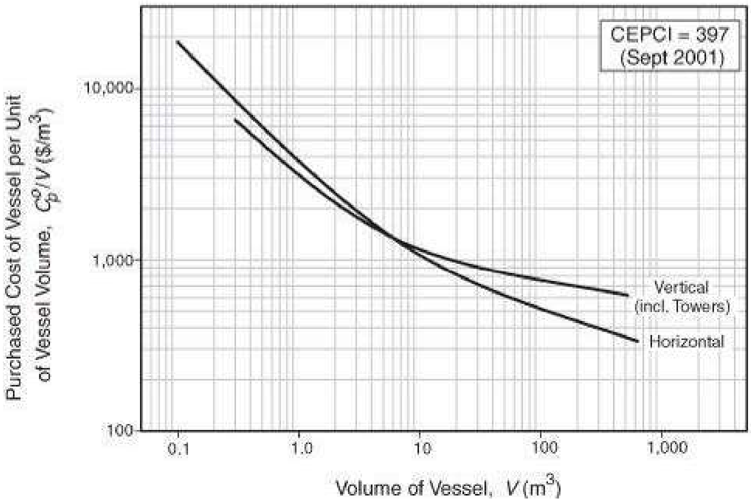
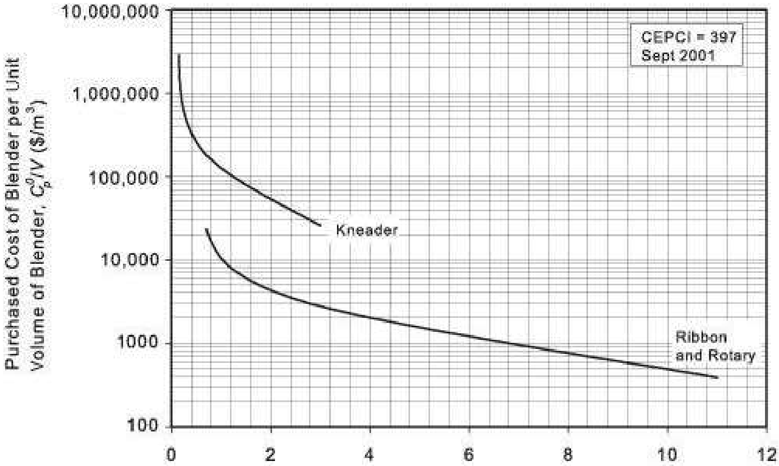


Figure A.8 Purchased Costs for Blenders



Volume of Blender, V (m^3)

Figure A.9 Purchased Costs of Centrifuges

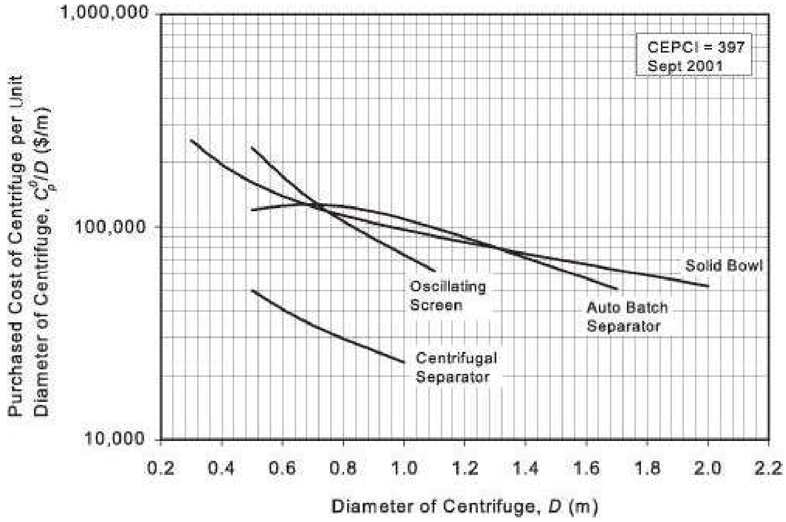


Figure A.10 Purchased Costs for Conveyors

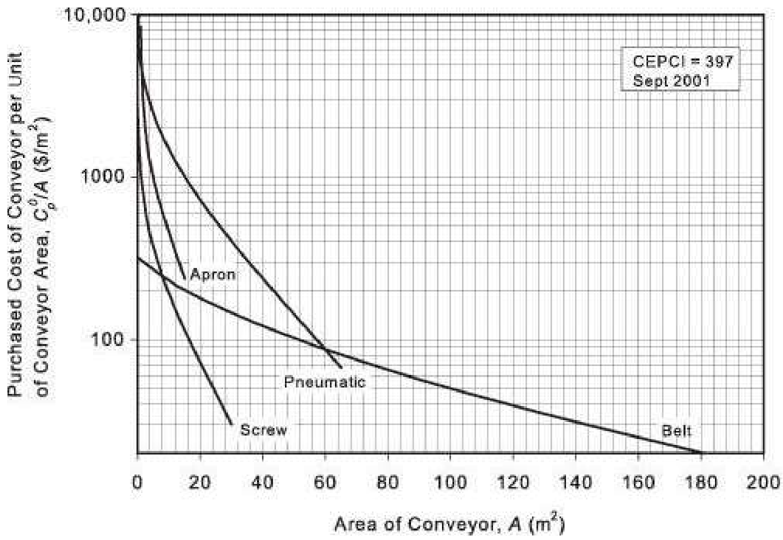


Figure A.11 Purchased Costs for Crystallizers

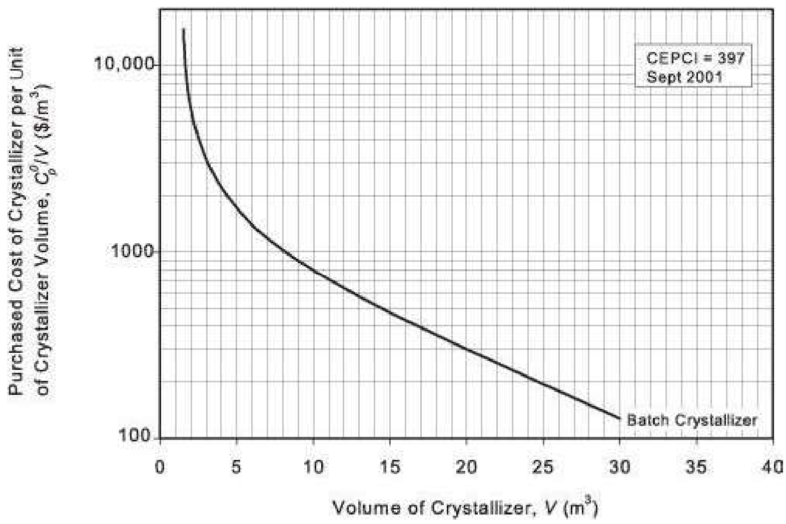


Figure A.12 Purchased Costs for Dryers

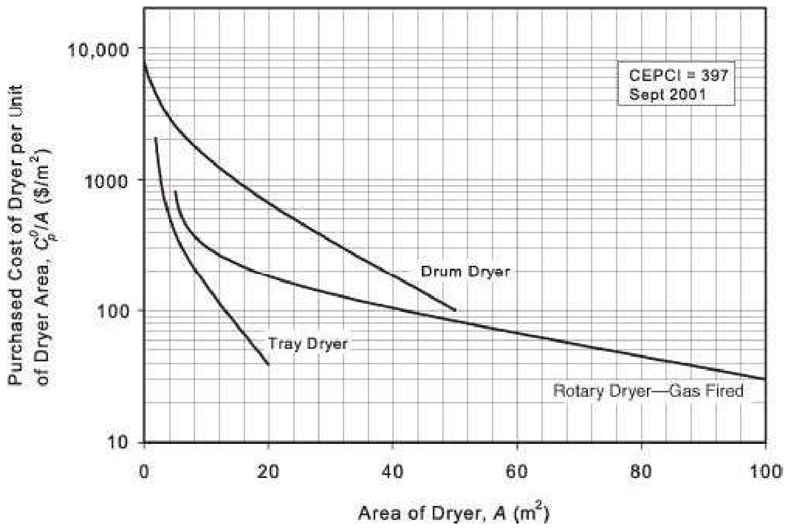


Figure A.13 Purchased Costs of Dust Collectors

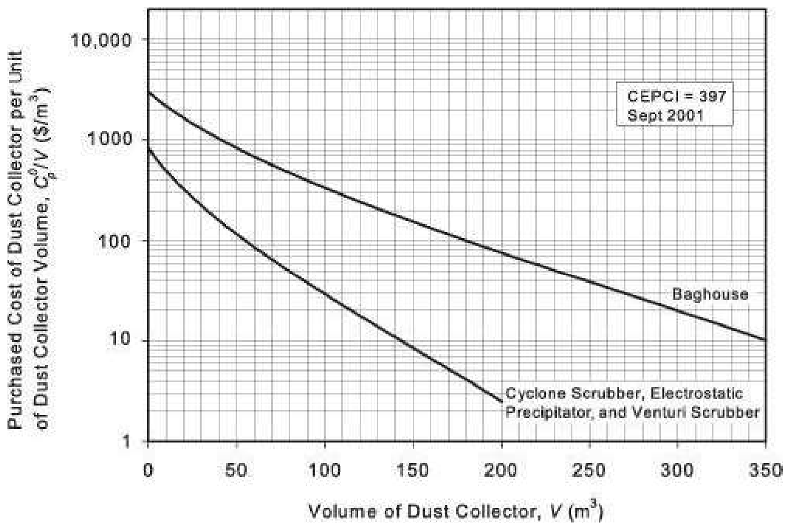


Figure A.14 Purchased Costs of Filters

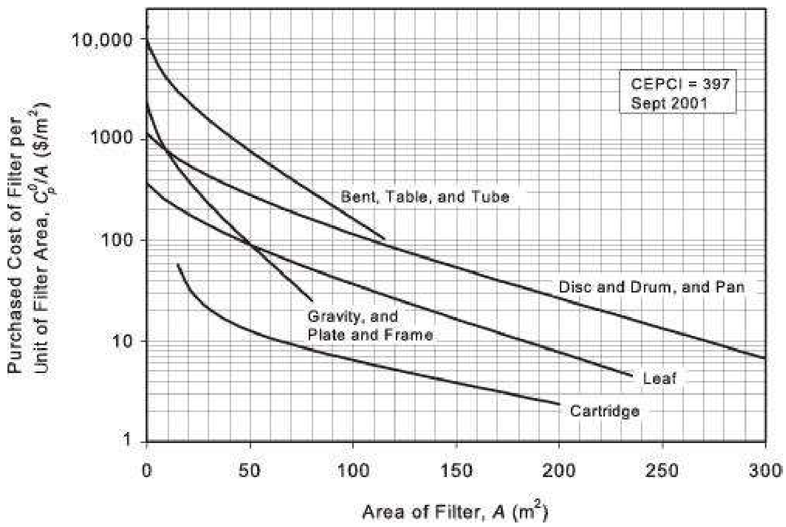


Figure A.15 Purchased Costs of Mixers

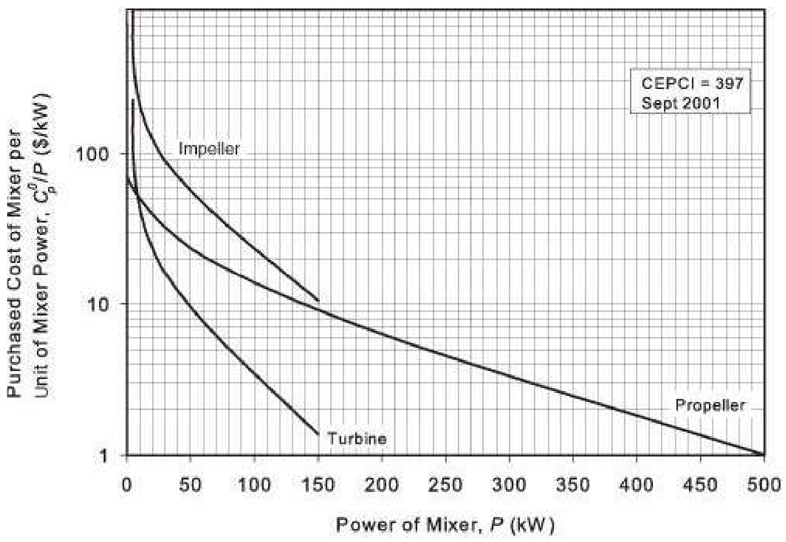


Figure A.16 Purchased Costs of Reactors

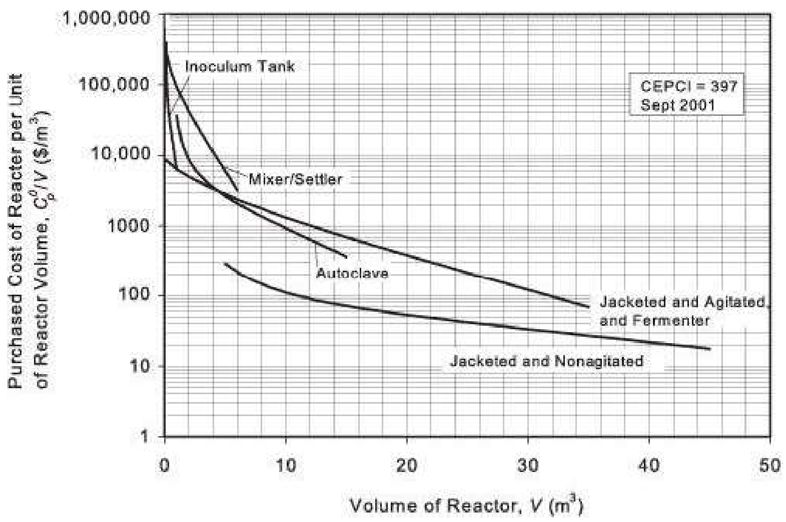
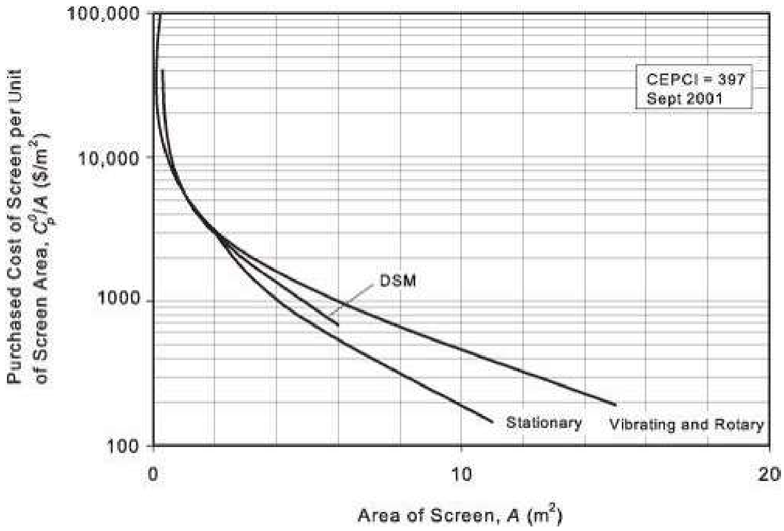


Figure A.17 Purchased Costs of Screens



Data from the R-Books software marketed by Richardson Engineering Services, Inc. [4], were used as a basis for several of the graphs and correlations; acknowledgment is given in the appropriate figures.

A.2 Pressure Factors

As was pointed out in Chapter 7, the costs of equipment increase with increasing operating pressure. In this section, the method of accounting for changes in operating pressure through the use of pressure factors is covered.

A.2.1 Pressure Factors for Process Vessels

The pressure factor for horizontal and vertical process (pressurized) vessels of diameter D meters and operating at a pressure of P barg is based on the ASME code for pressure vessel design [5]. At base material conditions using a maximum allowable stress for carbon steel, S, of 944 bar, a weld efficiency, E, of 0.9, a minimum allowable vessel thickness of 0.0063 m (1/4 inch), and a corrosion allowance, CA, of 0.00315 m (1/8 inch) gives the following expression:

(A.2)

$$F_{P,vessel} = \frac{\dfrac{(P+1)D}{2[850-0.6(P+1)]} + 0.00315}{0.0063} \quad \textit{for } t_{vessel} > 0.0063 \text{ m}$$

If $F_{P,\ vessel}$ is less than 1 (corresponding to $t_{vessel} < 0.0063$ m), then $F_{P,\ vessel} = 1$. For pressures less than –

0.5 barg, $F_{P,\ vessel}$ =1.25. It should be noted that Equation (A.2) is strictly true for the case when the thickness of the vessel wall is less than ¼ *D;* for vessels in the range *D* = 0.3 to 4.0 m, this occurs at pressures of approximately 320 barg.

A.2.2 Pressure Factors for Other Process Equipment

The pressure factors, F_P, for the remaining process equipment are given by the following general form:

(A.3)

$$\log_{10} F_P = C_1 + C_2 \log_{10} P + C_3 (\log_{10} P)^2$$

The units of pressure, *P*, are bar gauge or barg (1 bar = 0.0 barg) unless stated otherwise. The pressure factors are always greater than unity. The values of constants in Equation (A.3) for different equipment are given in Table A.2, and also shown are the ranges of pressures over which the correlations are valid. The values for the constants given in Table A.2 were regressed from data in Guthrie [1, 2] and Ulrich [3]. Extrapolation outside this range of pressures should be done with extreme caution. Some equipment does not have pressure ratings and therefore has values of C_1–C_3 equal to zero. If cost estimates are required for these units at high pressures and the equipment cost is affected by pressure, then the correlations should again be used with caution.

Table A.2 Pressure Factors for Process Equipment (Correlated from Data in Guthrie [1, 2], and Ulrich [3])

Equipment Type	Equipment Description	C_1	C_2	C_3	Pressure Range (barg)
Compressors	Centrifugal, axial, rotary, and reciprocating	0	0	0	–
Drives	Gas turbine	0	0	0	–
	Intern. comb. engine	0	0	0	–
	Steam turbine	0	0	0	–
	Electric—explosion-proof	0	0	0	–
	Electric—totally enclosed	0	0	0	–
	Electric—open/drip-proof	0	0	0	–
Evaporators	Forced circulation (pumped), falling film, agitated film (scraped wall), short tube, and long tube	0	0	0	P<10
		0.1578	−0.2992	0.1413	10<P<150
Fans*	Centrifugal radial, and centrifugal backward curve	0	0	0	ΔP<1kPa
		0	0.20899	−0.0328	1<ΔP<16kPa
	Axial vane and axial tube	0	0	0	ΔP<1kPa
		0	0.20899	−0.0328	1<ΔP<4kPa
Furnaces	Reformer furnace	0	0	0	P<10
		0.1405	−0.2698	0.1293	10<P<200
	Pyrolysis furnace	0	0	0	P<10
		0.1017	−0.1957	0.09403	10<P<200
	Nonreactive fired heater	0	0	0	P<10
		0.1347	−0.2368	0.1021	10<P<200
Heat exchangers	Scraped wall	0	0	0	P<40
		0.6072	−0.9120	0.3327	40<P<100
		13.1467	−12.6574	3.0705	100<P<300
	Teflon tube	0	0	0	P<15

(continued)

Equipment Type	Equipment Description	C_1	C_2	C_3	Pressure Range (barg)
	Bayonet, fixed tube sheet, floating head, kettle reboiler, and U-tube (both shell and tube)	0	0	0	P<5
		0.03881	−0.11272	0.08183	5<P<140
	Bayonet, fixed tube sheet, floating head, kettle reboiler, and U-tube (tube only)	0	0	0	P<5
		−0.00164	−0.00627	0.0123	5<P<140
	Double pipe and multiple pipe	0	0	0	P<40
		0.6072	−0.9120	0.3327	40<P<100
		13.1467	−12.6574	3.0705	100<P<300
	Flat plate and spiral plate	0	0	0	P<19
	Air cooler	0	0	0	P<10
		−0.1250	0.15361	−0.02861	10<P<100
	Spiral tube (both shell and tube)	0	0	0	P<150
		−0.4045	0.1859	0	150<P<400
	Spiral tube (tube only)	0	0	0	P<150
		−0.2115	0.09717	0	150<P<400
Heaters	Diphenyl heater, molten salt heater, and hot water heater	0	0	0	P<2
		−0.01633	0.056875	−0.00876	2<P<200
	Steam boiler	0	0	0	P<20
		2.594072	−4.23476	1.722404	20<P<40
Packing	Loose (for towers)	0	0	0	–
Process vessels	Horizontal and vertical			†	
Pumps	Reciprocating	0	0	0	P<10
		−0.245382	0.259016	−0.01363	10<P<100
	Positive displacement	0	0	0	P<10
		−0.245382	0.259016	−0.01363	10<P<100
	Centrifugal	0	0	0	P<10
		−0.3935	0.3957	−0.00226	10<P<100
Towers	Tray and packed			†	
Tanks	API—fixed roof	0	0	0	P<0.07
	API—floating roof	0	0	0	P<0.07
Trays	Sieve	0	0	0	–
	Valve	0	0	0	–
	Demisters	0	0	0	–
Turbines	Axial gas turbines	0	0	0	–
	Radial gas/liquid expanders	0	0	0	–
Vaporizers	Internal coils / jackets and jacket vessels	0	0	0	P<5
		−0.16742	0.13428	0.15058	5<P<320

*Pressure factors for fans are written in terms of the pressure rise across the fan, ΔP, where ΔP is measured in kPa.
†See Equation (A.2).

A.3 Material Factors and Bare Module Factors

As was pointed out in Chapter 7, the costs of equipment change with changes in the material of construction. In this section, the method of accounting for different materials of construction is covered.

A.3.1 Bare Module and Material Factors for Heat Exchangers, Process Vessels, and Pumps

The material factors, F_M, for heat exchangers, process vessels, and pumps are given in Figure A.18, with the appropriate identification number listed in Table A.3. The bare module factors for this equipment are given by the following equation:

(A.4)

$$C_{BM} = C_p^o F_{BM} = C_p^o (B_1 + B_2 F_M F_P)$$

Figure A.18 Material Factors for Equipment in Table A.3 (Averaged Data from References [1, 2, 3, 6, 7, and 8])

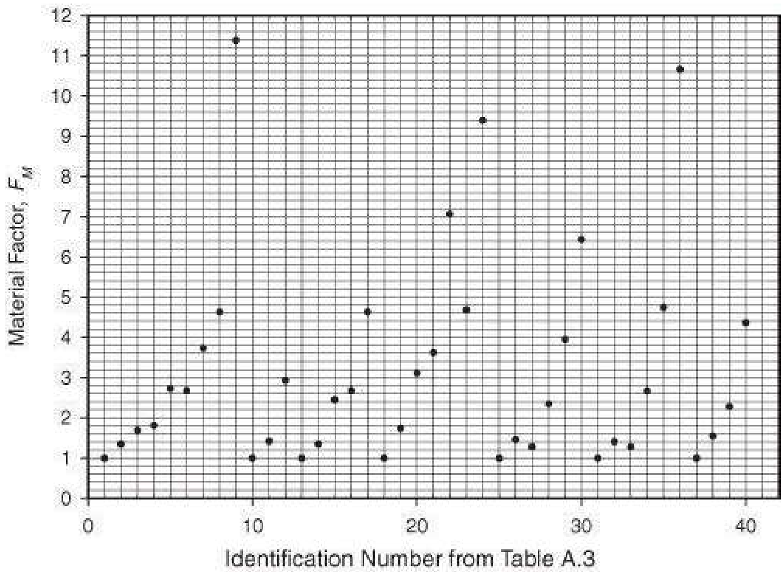


Table A.3 Identification Numbers for Material Factors for Heat Exchangers, Process Vessels, and Pumps to Be Used with Figure A.18

Identification Number	Equipment Type	Equipment Description	Material of Construction
1	Heat exchanger	Double pipe, multiple pipe,	CS-shell/CS-tube
2		fixed tube sheet, floating head,	CS-shell/Cu-tube
3		U-tube, bayonet, kettle reboiler, scraped	Cu-shell/Cu-tube
4		wall, and spiral tube	CS-shell/SS-tube
5			SS-shell/SS-tube
6			CS-shell/Ni alloy tube
7			Ni alloy, shell/Ni alloy-tube
8			CS-shell/Ti-tube
9			Ti-shell/Ti-tube
10		Air cooler	CS tube
11		Air cooler	Al tube
12		Air cooler	SS tube
13		Flat plate and spiral plate	CS (in contact with fluid)
14		Flat plate and spiral plate	Cu (in contact with fluid)
15		Flat plate and spiral plate	SS (in contact with fluid)
16		Flat plate and spiral plate	Ni alloy (in contact with fluid)
17		Flat plate and spiral plate	Ti (in contact with fluid)
18	Process vessels	Horizontal, vertical (including towers)	CS
19		Horizontal, vertical (including towers)	SS clad
20		Horizontal, vertical (including towers)	SS
21		Horizontal, vertical (including towers)	Ni alloy clad
22		Horizontal, vertical (including towers)	Ni alloy
23		Horizontal, vertical (including towers)	Ti clad
24		Horizontal, vertical (including towers)	Ti

Identification Number	Equipment Type	Equipment Description	Material of Construction
25	Pumps	Reciprocating	Cast iron
26		Reciprocating	Carbon steel
27		Reciprocating	Cu alloy
28		Reciprocating	SS
29		Reciprocating	Ni alloy
30		Reciprocating	Ti
31		Positive displacement	Cast iron
32		Positive displacement	Carbon steel
33		Positive displacement	Cu alloy
34		Positive displacement	SS
35		Positive displacement	Ni alloy
36		Positive displacement	Ti
37		Centrifugal	Cast iron
38		Centrifugal	Carbon steel
39		Centrifugal	SS
40		Centrifugal	Ni alloy

The values of the constants B_1 and B_2 are given in Table A.4. The bare module cost for ambient pressure and carbon steel construction, C°_{BM}, and the bare module factor for the equipment at these conditions, F°_{BM}, are found by setting F_M and F_P equal to unity. The data given in Tables A.3 and A.4 and Figure A.18 are average values from the following references: Guthrie [1, 2], Ulrich [3], Navarrete [6], Perry et al. [7], and Peters and Timmerhaus [8].

Table A.4 Constants for Bare Module Factor to Be Used in Equation A.4 (Correlated from Data in Guthrie [1, 2] and Ulrich [3])

Equipment Type	Equipment Description	B_1	B_2
Heat exchangers	Double pipe, multiple pipe, scraped wall, and spiral tube	1.74	1.55
	Fixed tube sheet, floating head, U-tube, bayonet, kettle reboiler, and Teflon tube	1.63	1.66
	Air cooler, spiral plate, and flat plate	0.96	1.21
Process vessels	Horizontal	1.49	1.52
	Vertical (including towers)	2.25	1.82
Pumps	Reciprocating	1.89	1.35
	Positive displacement	1.89	1.35
	Centrifugal	1.89	1.35

A.3.2 Bare Module and Material Factors for the Remaining Process Equipment

For the remaining equipment, the bare module costs are related to the material and pressure factors by equations different from Equation (A.4). The form of these equations is given in Table A.5. The bare module factors that correspond to the equations in Table A.5 are given in Figure A.19 using the identification numbers listed in Table A.6. Again, the data used to construct Figure A.19 are compiled from average values taken from Guthrie [1, 2], Ulrich [3], Navarrete [6], Perry et al. [7], and Peters and Timmerhaus [8]. In addition, bare module factors for the equipment added to the third edition of the book (conveyors, crystallizers, dryers, dust collectors, filters, mixers, reactors, and screens) are given separately in Table A.7.

Table A.5 Equations for Bare Module Cost for Equipment Not Covered by Tables A.3 and A.4

Equipment Type	Equation for Bare Module Cost
Compressors and blowers without drives	$C_{BM} = C_p^o F_{BM}$
Drives for compressors and blowers	$C_{BM} = C_p^o F_{BM}$
Evaporators and vaporizers	$C_{BM} = C_p^o F_{BM} F_P$
Fans with electric drives	$C_{BM} = C_p^o F_{BM} F_P$
Fired heaters and furnaces	$C_{BM} = C_p^o F_{BM} F_P F_T$ F_T is the superheat correction factor for steam boilers ($F_T = 1$ for other heaters and furnaces) and is given by $F_T = 1 + 0.00184\Delta T - 0.00000335(\Delta T)^2$ where ΔT is the amount of superheat in °C.
Power recovery equipment	$C_{BM} = C_p^o F_{BM}$
Sieve trays, valve trays, and demister pads	$C_{BM} = C_p^o N F_{BM} F_q$ Where N is the number of trays and F_q is a quantity factor for trays only given by $\log_{10} F_q = 0.4771 + 0.08516 \log_{10} N - 0.3473 (\log_{10} N)^2$ for $N<20$ $F_q = 1$ for $N \geq 20$
Tower packing	$C_{BM} = C_p^o F_{BM}$

Figure A.19 Bare Module Factors for Equipment in Table A.6 (Average Data from References [1, 2, 3, 6, 7, and 8])

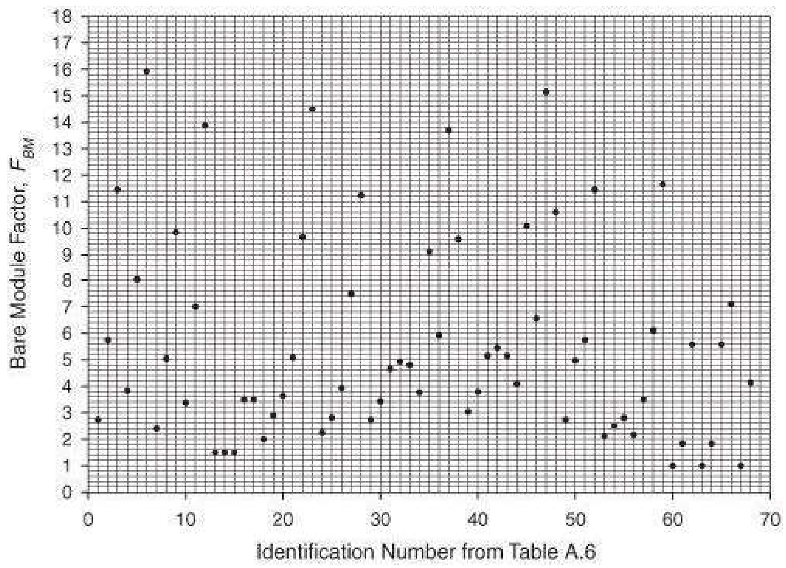


Table A.6 Identification of Material Factors for Equipment Listed in Table A.5 to Be Used with Figure A.19

Identification Number	Equipment Type	Equipment Description	Material of Construction
1	Compressors/blowers	Centrifugal compressor or blower	CS
2		Centrifugal compressor or blower	SS
3		Centrifugal compressor or blower	Ni alloy
4		Axial compressor or blower	CS
5		Axial compressor or blower	SS
6		Axial compressor or blower	Ni alloy
7		Rotary compressor or blower	CS
8		Rotary compressor or blower	SS
9		Rotary compressor or blower	Ni alloy
10		Reciprocating compressor or blower	CS
11		Reciprocating compressor or blower	SS
12		Reciprocating compressor or blower	Ni alloy
13	Drives for compressors and blowers	Electric—explosionproof	—
14		Electric—totally enclosed	—
15		Electric—open/dripproof	—
16		Gas turbine	—
17		Steam turbine	—
18		Internal combustion engine	—
19	Evaporators and vaporizers	Evaporator—forced circ, short or long tube	CS
20		Evaporator—forced circ, short or long tube	Cu alloy
21		Evaporator—forced circ, short or long tube	SS
22		Evaporator—forced circ, short or long tube	Ni alloy
23		Evaporator—forced circ, short or long tube	Ti
24		Evaporator—falling film, scraped-wall	CS
25		Evaporator—falling film, scraped-wall	Cu alloy
26		Evaporator—falling film, scraped-wall	SS
27		Evaporator—falling film, scraped-wall	Ni alloy
28		Evaporator—falling film, scraped-wall	Ti
29		Vaporizer—jacketed vessel	CS
30		Vaporizer—jacketed vessel	Cu
31		Vaporizer—jacketed vessel	Glass lined/SS coils
32		Vaporizer—jacketed vessel	Glass lined/Ni coils
33		Vaporizer—jacketed vessel	SS
34		Vaporizer—jacketed vessel	SS clad
35		Vaporizer—jacketed vessel	Ni alloy
36		Vaporizer—jacketed vessel	Ni alloy clad
37		Vaporizer—jacketed vessel	Ti
38		Vaporizer—jacketed vessel	Ti clad
39		Vaporizer—jacketed vessel + internal coil	CS
40		Vaporizer—jacketed vessel + internal coil	Cu
41		Vaporizer—jacketed vessel + internal coil	Glass lined/SS coils
42		Vaporizer—jacketed vessel + internal coil	Glass lined/Ni coils
43		Vaporizer—jacketed vessel + internal coil	SS
44		Vaporizer—jacketed vessel + internal coil	SS clad
45		Vaporizer—jacketed vessel + internal coil	Ni alloy
46		Vaporizer—jacketed vessel + internal coil	Ni alloy clad
47		Vaporizer—jacketed vessel + internal coil	Ti
48		Vaporizer—jacketed vessel + internal coil	Ti clad
49	Fans	Fan with electric drive	CS

Identification Number	Equipment Type	Equipment Description	Material of Construction
50		Fan with electric drive	Fiberglass
51		Fan with electric drive	SS
52		Fan with electric drive	Ni alloy
53	Fired heaters and furnaces	Tube for furnaces and nonreactive process heater	CS
54		Tube for furnaces and nonreactive process heater	Alloy steel
55		Tube for furnaces and nonreactive process heater	SS
56		Thermal fluid heater—hot water, molten salt, or diphenyl-based oil	—
57	Power recovery equipment	Turbines	CS
58		Turbines	SS
59		Turbines	Ni alloy
60	Trays and demister pads	Sieve and valve trays	CS
61		Sieve and valve trays	SS
62		Sieve and valve trays	Ni alloy
63		Demister pad	SS
64		Demister pad	Fluorocarbon
65		Demister pad	Ni alloy
66	Tower packing	Packing	Metal (304SS)
67		Packing	Polyethylene
68		Packing	Ceramic

Figure A.7 Bare Module Factors for Conveyors, Crystallizers, Dryers, Dust Collectors, Filters, Mixers, Reactors, and Screens

Equipment Type	Equipment Description	Bare Module Factor, FBM
Blenders	Kneader	1.12*
	Ribbon	1.12*
	Rotary	1.12
Centrifuges	Auto batch separator	1.57*
	Centrifugal separator	1.57
	Oscillating screen	1.57*
	Solid bowl w/o motor	1.27
Conveyors	Apron	1.20
	Belt	1.25
	Pneumatic	1.25*
	Screw	1.10
Crystallizers	Batch	1.60
Dryers	Drum	1.60
	Rotary, gas fired	1.25
	Tray	1.25
Dust Collectors	Baghouse	2.86*
	Cyclone scrubbers	2.86*
	Electrostatic precipitator	2.86*
	Venturi scrubber	2.86*
Filters	Bent	1.65*
	Cartridge	1.65*
	Disc and drum	1.65*
	Gravity	1.65*
	Leaf	1.65
	Pan	1.65*
	Plate and frame	1.80
	Table	1.65*
	Tube	1.65*
Mixers	Impeller	1.38*
	Propeller	1.38
	Turbine	1.38
Reactors	Autoclave	4.0*
	Fermenter	4.0*
	Inoculum tank	4.0*
	Jacketed agitated	4.0*
	Jacketed nonagitated	4.0*
	Mixer/settler	4.0*
Screens	DSM	1.34*
	Rotary	1.34*
	Stationary	1.34*
	Vibrating	1.34

When possible, bare module factors are taken to be equal to the Field Installation Factors from Guthrie [2]. Items marked * are estimates.

References

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