

**Equilibrium Constant (*T* is in units of K)**

$$\ln(K_p) = 13.51 + \frac{5037}{T} - 2.073\ln(T) + 3.499 \times 10^{-4}T + 4.173 \times 10^{-8}T^2 + \frac{3017}{T^2}$$

**Heat of Reaction**

$$\Delta H_{\text{reaction}} = -37,190 - 17.24T + 29.09 \times 10^{-4}T^2 + 0.6939 \times 10^{-6}T^3 + \frac{50,160}{T} \quad \frac{\text{kJ}}{\text{kmol}}$$

**At the Reaction Conditions of 600°C (873 K)**

Equilibrium Constant,  $K_p = 265$

$$\text{Heat of Reaction, } \Delta H_{\text{reaction}} = -49,500 \quad \frac{\text{kJ}}{\text{kmol}}$$

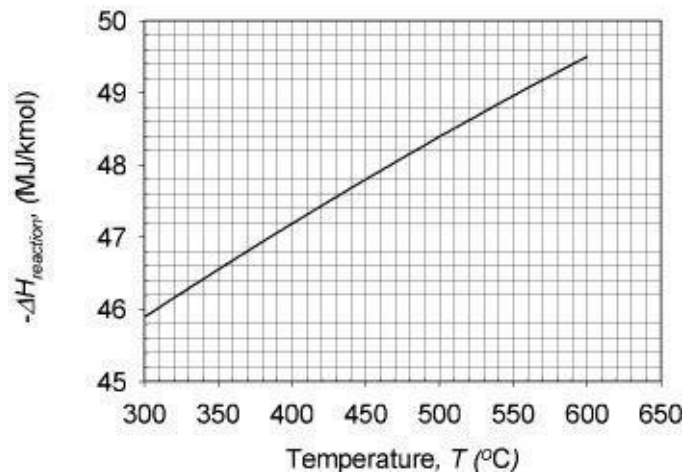
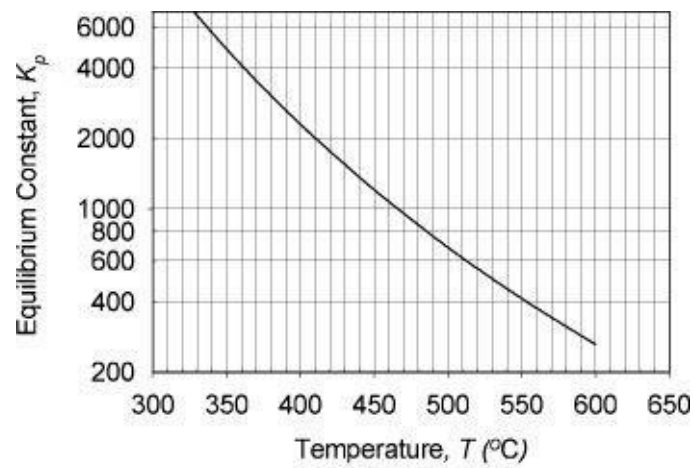
**Information on Reaction Kinetics**

No side reactions

Reaction is kinetically controlled

[Figure 6.1](#) is a plot of the heat of reaction and the equilibrium constant as a function of temperature, evaluated from the information provided in [Table 6.6](#). From these plots it is evident that the chemical reaction is slightly exothermic, causing the equilibrium constant to decrease with temperature.

**Figure 6.1** Equilibrium Constant and Heat of Reaction as a Function of Temperature for the Toluene Hydrodealkylation Reaction



**Reaction Kinetics Information:** This information is reaction specific and must be obtained experimentally. The overall kinetics may involve homogeneous and heterogeneous reactions both catalytic and noncatalytic. The expressions are often complex.

Before a process is commercialized, reaction kinetics information, such as space velocity and residence times, must be obtained for different temperatures and pressures from pilot plant studies. Such data are necessary to design the reactor. At this point, we are not interested in the reactor design, and hence specific kinetics expressions have not been included in [Table 6.6](#) and are not necessary for the following analysis.

The analysis of the reactor takes place in two parts.

- a. Evaluation of the special conditions from the thermodynamic point of view. This assumes that chemical equilibrium is reached and provides a limiting case.
- b. Evaluation of the special conditions from the kinetics point of view. This accounts for the limitations imposed by reaction kinetics, mass transfer, and heat transfer.

If a process is unattractive under equilibrium (thermodynamic) conditions, analysis of the kinetics is not necessary. For processes in which the equilibrium conditions give favorable results, further study is necessary. The reason for this is that conditions that favor high equilibrium conversion may be unfavorable from the standpoint of reaction kinetics.

**Thermodynamic Considerations.** We consider the use of high-temperature, high-pressure, and non-

stoichiometric feed conditions separately.

**High-Temperature Concern (see Table 6.1).** Figure 6.1 provided the important information that the reaction is exothermic. Table 6.1 notes that for an exothermic reaction, the result of increasing temperature is a reduction in equilibrium conversion. This is confirmed by the plot of the equilibrium constant versus temperature given in Figure 6.1. The decrease in the equilibrium conversion is undesirable. The actual conversion for the HDA process is compared with the equilibrium conversion in Example 6.3.

### Example 6.3

For the PFD presented in Figure 1.5,

- a. Calculate the actual conversion.
- b. Evaluate the equilibrium conversion at 600°C.

Assuming ideal gas behavior:  $K_p = (N_{benzene} N_{methane}) / (N_{toluene} N_{hydrogen})$

where  $N$  represents the moles of each species at equilibrium.

Information on the feed stream to the reactor from Table 1.5 (Stream 6 on Figure 1.5):

Hydrogen 735.4 kmol/h

Methane 317.3

Benzene 7.6

Toluene 144.0

Total 1204.3

- c. Actual Conversion: Toluene in exit stream (Stream 9) = 36 kmol/h  
Conversion =  $(144 - 36) / 144 = 0.75$  (75%)

- d. Equilibrium Conversion at 600°C. From Table 6.6 @600°C  $K_p = 265$

Let  $N$  = kmol/h of benzene formed

$$265 = [(N + 7.6)(N + 317.3)] / [(735.4 - N)(144 - N)]$$

$$N = 143.6$$

$$\text{Equilibrium Conversion} = 143.6 / 144 = 0.997 \text{ (99.7\%)}$$

The equilibrium conversion for the hydrodealkylation reaction remained high in spite of the high temperature. Although there is no real problem with using the elevated temperature in the reactor, it cannot be justified from a thermodynamic point of view.

**High-Pressure Concern (see Table 6.2).** From the reaction stoichiometry, we see that there are equal numbers of reactant and product moles in the hydrodealkylation reaction. For this case, there is no effect of pressure on equilibrium conversion. From a thermodynamic point of view there is no reason for the high pressure in the reactor.

**Non-stoichiometric Feed (see Table 6.3).** The component feed rates to the reactor (see Example 6.3) show that

1. Toluene is the limiting reactant.
2. Hydrogen is an excess reactant (more than 400% excess).
3. Methane, a reaction product, is present in significant amounts.

**Reaction Products (Methane) in Feed.** The presence of reaction product in the feed results in a reduction in the equilibrium conversion (see [Table 6.3](#)). However, [Example 6.3](#) shows that at the conditions selected for the reactor, the equilibrium conversion remained high despite the presence of the methane in the feed.

**Excess Reactant (Hydrogen) in Feed.** The presence of excess reactants in the feed results in an increase in equilibrium conversion (see [Table 6.3](#)). [Example 6.4](#) explores the effect of this excess hydrogen on conversion.

#### Example 6.4

(Reference [Example 6.3](#)). Reduce the amount of hydrogen in the feed to the reactor to the stoichiometric amount—that is, 144 kmol/h—and determine the effect on the equilibrium conversion at 600°C.

The calculations are not shown. They are similar to those in [Example 6.3\(b\)](#). The total moles of hydrogen in the feed were changed from 735.4 kmol/h to the stoichiometric value of 144 kmol/h.

The results obtained were  $N = 128.8$  kmol/h, equilibrium conversion = 0.895 (89.5%).

[Example 6.4](#) reveals that the presence of the large excess of hydrogen had a noticeable effect on the equilibrium conversion.

We conclude that thermodynamic considerations do not explain the selection of the high temperature, the high pressure, and the presence of reaction products in the feed. The presence of a large excess of hydrogen is the only positive effect predicted by thermodynamics.

**Consideration of Reaction Kinetics.** The information on reaction kinetics is limited in this chapter. We will present a more detailed description of the kinetics rate expression in a case study in [Chapter 20](#), and we investigate the predictions made in this chapter with this limited information. However, you will find that a great deal of understanding can be extracted from the limited information presented here.

From the information provided in [Table 6.6](#) and [Chapter 1](#) we know that

1. The reaction takes place in the gas phase.
2. The reaction is kinetically controlled.
3. There are no significant side reactions.

**High-Temperature Concern (see [Table 6.1](#)).** In a region where the reaction kinetics control, the reaction rate increases rapidly with temperature, as [Example 6.5](#) illustrates.

#### Example 6.5

The activation energy for the rate of reaction for the hydrodealkylation of toluene is equal to 148.1 kJ/mol (Tarhan [2]). What is the reaction rate at 600°C relative to that at 400°C?

$$\begin{aligned}\text{Ratio of Reaction Rates} &= \exp[-E/R\{1/T_2 - 1/T_1\}] \\ &= \exp[148100/8.314\{1/673 - 1/873\}] = 430\end{aligned}$$

The size of a reactor would increase by nearly three orders of magnitude if the reaction were carried out at 400°C (the critical temperature for materials selection, [Table 6.1](#)) rather than 600°C. Clearly the effect of temperature is significant.

Most reactions are not kinetically controlled as is the case here. In most cases the rate is controlled by heat or mass transfer considerations. These are not as sensitive to temperature changes as chemical reaction rates. For more detail, see [Chapter 20](#).

**High-Pressure Concern (see [Table 6.2](#)).** For gas phase reactions, the concentration of reactants is proportional to the pressure. For a situation where the reaction rate is directly proportional to the concentration, operation at 25 bar rather than at 1 bar would increase the reaction rate by a factor of 25 (assuming ideal gas behavior). Although we do not know that the rate is directly proportional to the concentration, we can predict that the effect of pressure is likely to be substantial, and the reactor size will be substantially reduced.

**Non-stoichiometric Feed (see [Table 6.3](#)).** The reactor feed contains both excess hydrogen and the reaction product methane.

**Methane in the Feed.** The effect of methane is to reduce the reactant concentrations. This decreases the reaction rate and represents a negative impact. The methane could possibly reduce the formation of side products, but we have no information to suggest that this is the case.

**Excess Hydrogen in the Feed.** The large amount of excess hydrogen in the feed ensures that the concentration of hydrogen will remain large throughout the reactor. This increases the reaction rate. Although there is no information provided regarding the decision to maintain the high hydrogen levels, it may be linked to reducing the formation of side products.

With the exception of the presence of methane product in the feed, the high-temperature operation, the excess hydrogen, and the elevated pressure all support an increase in reaction rate and a reduction in reactor volume. This suggests that the catalyst is not “hot”—that is, the catalyst is still operating in the reaction-controlled regime and mass transfer effects have not started to intrude. For these conditions, the manipulation of temperatures and pressures is essential to limit the reactor size.

There is a significant economic penalty for using more than 400% excess hydrogen in the reactor feed. The raw material cost of hydrogen would be reduced significantly if excess hydrogen were not used. The fact that this large excess is used in spite of the economic penalty involved suggests that the hydrogen plays an important role in the prevention of side products. The concept of selectivity is discussed further in [Chapter 20](#).

The presence of methane in the feed has not yet been resolved. At best it behaves as an inert and occupies volume that must be handled downstream of the reactor, thus making all the equipment larger and more expensive. This question is considered in more detail in [Example 6.6](#).

### Example 6.6

It has been proposed that we handle the hydrogen/methane stream in the same manner that we handled the toluene/benzene stream. We recall that the unreacted toluene was separated from the benzene product and then recycled. It is proposed that the methane be separated from the hydrogen. The methane would then become a process by-product and the hydrogen would be recycled. Discuss this proposal using the arguments provided in [Tables 6.1](#) and [6.2](#).

To use distillation for the separation of methane from hydrogen, as was used with the toluene/benzene, requires a liquid phase. For methane/hydrogen systems, this requires extremely high pressures together with cryogenic temperatures.

If the hydrogen could be separated from the methane and recycled, then the reactor feed would not contain significant quantities of methane, and the large excess of hydrogen could be maintained without the steep cost of excess hydrogen feed. Note that the overall conversion of hydrogen in the process is only 37%, whereas for toluene it is 99%.

Alternative separation schemes that do not require a liquid phase (e.g., a membrane separator) should be considered. The use of alternative separation technologies is addressed further in [Chapter 12](#).

#### **6.4.2 Evaluation of High-Pressure Phase Separator V-102**

This vessel separates toluene and benzene as a liquid from the noncondensable gases hydrogen and methane. The reactor product is cooled and forms a vapor and a liquid stream that are in equilibrium. The vapor-liquid equilibrium is that at the temperature and pressure of the stream entering V-102. From [Tables 6.1](#) and [6.2](#), we conclude that the lower temperature (38°C) was provided to obtain a liquid phase for the vapor-liquid equilibrium. The pressure was maintained to support the formation of the liquid phase. Because the separation can be affected relatively easily at high pressure, it is worthwhile maintaining V-102 at this high pressure.

#### **6.4.3 Evaluation of Large Temperature Driving Force in Exchanger E-101**

There is a large temperature driving force in this exchanger, because the heating medium is at a temperature of approximately 250°C, and the inlet to the exchanger is only 30°C. This is greater than the 100°C suggested in [Table 6.4](#). This is an example of poor heat integration, and we will take a closer look at improving this in [Chapter 15](#) (also see the case study presented in [Chapter 28](#)).

#### **6.4.4 Evaluation of Exchanger E-102**

Stream 9 is cooled from 654°C to 40°C using cooling water at approximately 35°C. Again this is greater than the 100°C suggested in [Table 6.4](#), and the process stream has a lot of valuable energy that is being wasted. Again, we can save a lot of money by using heat integration (see [Chapter 15](#)).

#### **6.4.5 Pressure Control Valve on Stream 8**

The purpose of this control valve is to reduce the pressure of the stream entering the fuel gas line from 23.9 bar to 2.5 bar. This reduction in pressure represents a potential loss of useful work due to the throttling action of the valve. Referring to [Table 6.4](#), we can see that when we throttle a gas, we can recover work by using a turbine, although this may not be economically attractive. The operation of this valve is justified because of its control function.

#### **6.4.6 Pressure Control Valve on Stream from V-102 to V-103**

The purpose of this valve is to reduce the pressure of the liquid leaving V-102. This reduction in pressure causes some additional flashing and recovery of dissolved methane and hydrogen from the toluene/benzene mixture. The flashed gas is separated in V-103 and sent to the fuel gas line. The purpose of this valve is to control the pressure of the material fed to the distillation column T-101. Because the stream passing through the valve is essentially all liquid, little useful work could be recovered from this stream.

This completes our review of the conditions of special concern for the toluene hydrodealkylation process.

### **6.5 Summary**

In this chapter, you learned to identify process conditions that are of special interest or concern in the analysis of the PFD. A series of tables was presented in which justifications for using process conditions of special concern were given. We introduced the process conditions matrix (PCM) for the toluene hydrodealkylation process and identified all the equipment in which process conditions of special concern existed. Finally, by comparing the process conditions from the PFD to those given in the tables, we learned to analyze why these conditions were selected for the process and where improvements may be made.

### **References**

1. Walas, S. M., *Chemical Process Equipment: Selection and Design* (Stoneham, MA: Butterworth, 1988).
2. Tarhan, M. O., *Catalytic Reactor Design* (New York: McGraw-Hill, 1983).

### **Short Answer Questions**

1. State two common criteria for setting the pressure of a distillation column.

2. Suggest two reasons each why distillation columns are run *above* or *below* ambient pressure. Be sure to state clearly which explanation is for above and which is for below ambient pressure.
3. Suggest two reasons why reactors are run at elevated pressures and/or temperatures. Be sure to state clearly which explanation is for elevated pressure and which is for elevated temperature.
4. Give two reasons why operation of a process greater than 250°C is undesirable. Give one reason each why one would operate a distillation column and a reactor at temperature greater than 250°C.
5. Define a “condition of special concern.” Define two such conditions, and state one possible justification for each.

- In the food and drug industries, many processes used to produce new active ingredients (drugs) or to separate and purify drugs and foods occur at vacuum conditions and often at low temperatures (less than room temperature). What is it about these types of products that requires that these conditions of special concern be used?
- 6.

## Problems

For the separation of a binary mixture in a distillation column, what will be the effect of an increase in column pressure on the following variables?

7.
  - a. Tendency to flood at a fixed reflux ratio
  - b. Reflux ratio for a given top and bottom purity at a constant number of stages
  - c. Number of stages required for a given top and bottom purity at constant reflux ratio
  - d. Overhead condenser temperature

- In a new chemical process, a reboiler for a tower requires a heating medium at 290°C. Two possible solutions have been suggested: (a) use high-pressure steam superheated to 320°C, and (b) use saturated steam at 320°C. Suggest one disadvantage for each suggestion.
- 8.

- As the ambient temperature and humidity increase, the temperature at which cooling water (cw) can be supplied to any piece of equipment increases. For example, in the winter, cw may be available at 27°C whereas in midsummer it may rise to 34°C. How, if at all, does this affect the pressure at which a distillation column operates (assuming that the overhead condenser uses cooling water as the heat exchange utility)?
- 9.
  - 10.

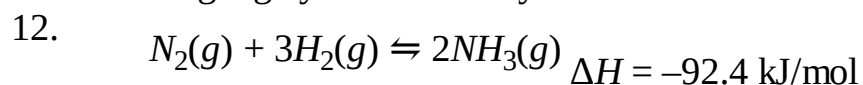
- It is desired to produce a hot vapor stream of benzene to feed a reactor for a certain petrochemical process. The benzene is available from an off-site storage facility at 1 atm pressure and ambient temperature (assume 25°C), and the reactor requires the benzene to be at 250°C at 10 atm. Two possible process schemes are being considered to heat and pressurize the feed: (1) pump the liquid benzene to pressure and then vaporize it in a heat exchanger, and (2) vaporize the benzene first and then compress it to the desired pressure. Answer the following.
10.
    - a. Discuss qualitatively which scheme (if either) is better.
    - b. Confirm your answer to part (a) by comparing the costs using both schemes to feed 1000 kg/h of benzene to the reactor. (Assume that the cost of heating is \$15/GJ and that

electricity costs \$0.06/kWh.)

One way to produce very pure oxygen and nitrogen is to separate air using a distillation process. For such a separation determine the following.

- a. Find the normal boiling point (at 1 atm pressure) of nitrogen and oxygen.
- b. For a distillation column operating at 1 atm pressure, what would be the top and bottom temperatures and top and bottom compositions of a distillation column that separates air into nitrogen and oxygen? (For this problem, you may assume that air contains only nitrogen and oxygen and that pure components leave at the top and bottom of the column.)
11. c. At what pressure can oxygen and nitrogen be liquefied at ambient temperature (say 40°C)?
- d. What does the answer to part (c) tell you about the potential to distill air at ambient conditions?

The production of ammonia (a key ingredient for fertilizer) using the Haber process takes place at temperatures of around 500°C and pressures of 250 atm using a porous iron catalyst according the following highly exothermic synthesis reaction:



Give possible reasons for the high temperature and pressure used for this reaction.

- Consider the ammonia process in Problem 12. For the given conditions, the maximum single-pass conversion obtained in the reactor is about 15–20%. Explain how the temperature and pressure should be adjusted to increase this conversion and the penalties for making these changes.
- 13.

For the production of drying oil shown as Project B.4 in [Appendix B](#), do the following.

- a. Construct a process conditions matrix (PCM) for the process, and determine all conditions of special concern.
14. b. For each condition of special concern identified in part (a), suggest at least one reason why such a condition was used.
- c. For each condition of special concern identified in part (a), suggest at least one process alternative to eliminate the condition.

For the styrene production process given in Project B.3 in [Appendix B](#), do the following.

- a. Construct a process conditions matrix (PCM) for the process, and determine all conditions of special concern.
15. b. Explain the reasons for using the conditions of special concern in the reactor.
- c. Suggest any process alternatives for part (b).

## Section 2

# Engineering Economic Analysis of Chemical Processes

In this section, we concentrate on the evaluation of the economics of a chemical process. In order for a chemical engineer or cost engineer to evaluate the economic impact of a new (or existing) chemical process, certain technical information must be available. Although this information is gleaned from a variety of sources, it is generally presented in the form of the technical diagrams discussed in [Chapter 1](#).

In the chapters of this section, methods to evaluate the economics of a chemical process are covered. The term *economics* refers to the evaluation of capital costs and operating costs associated with the construction and operation of a chemical process. The methods by which the one-time costs associated with the construction of the plant and the continuing costs associated with the daily operation of the process are combined into meaningful economic criteria are provided.

This material is treated in the following chapters.

### [Chapter 7](#): Estimation of Capital Costs

The common types of estimates are presented along with the basic relationships for scaling costs with equipment size. The concept of cost inflation is presented, and some common cost indexes are presented. The concept of total fixed capital investment to construct a new process is discussed, and the cost module approach to estimating is given. Finally, the software program (CAPCOST) to evaluate fixed capital costs (and other financial calculations) is described.

### [Chapter 8](#): Estimation of Manufacturing Costs

The basic components of the manufacturing costs of a process are presented. A method to relate the total cost of manufacturing (COM) to five elements—fixed capital investment, cost of operating labor, cost of raw materials, cost of utilities, and cost of waste treatment—is given. Examples of how utility costs can be calculated from the basic costs of fuel, power, and water are discussed. The estimation of labor costs based on the size and complexity of the process are also given.

### [Chapter 9](#): Engineering Economic Analysis

The concept of the time value of money is discussed. The following topics are presented: simple and compound interest, effective and nominal interest rates, annuities, cash flow diagrams, and discount factors. In addition, the concepts of depreciation, inflation, and taxation are covered.

### [Chapter 10](#): Profitability Analysis

The ideas discussed in [Chapter 9](#) are extended to evaluate the profitability of chemical processes. Profitability criteria using nondiscounted and discounted bases are presented and include net present value (NPV), discounted cash flow rate of return (DCFROR), and payback period (PBP). A discussion of evaluating equipment alternatives using equivalent annual operating costs (EAOC) and other methods is presented. Finally, the concept of evaluating risk is covered and an introduction to the Monte Carlo method is presented.

# 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.

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### 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.

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### **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 <sup>2</sup>	0.59
Vertical tank carbon steel	0.4 to 76	m <sup>3</sup>	0.30
Centrifugal blower	0.24 – 71	std m <sup>3</sup> /s	0.60
Jacketed kettle glass lined	0.2 to 3.8	m <sup>3</sup>	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)

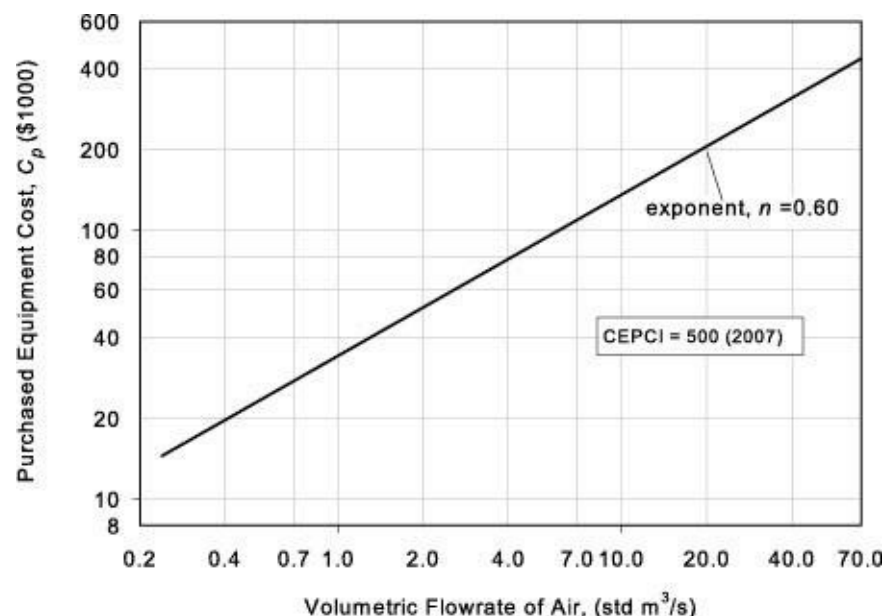
$$C_a = K A_a^n$$

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 =  $((2.63 - 3.86)/3.86)(100) = -32\%$

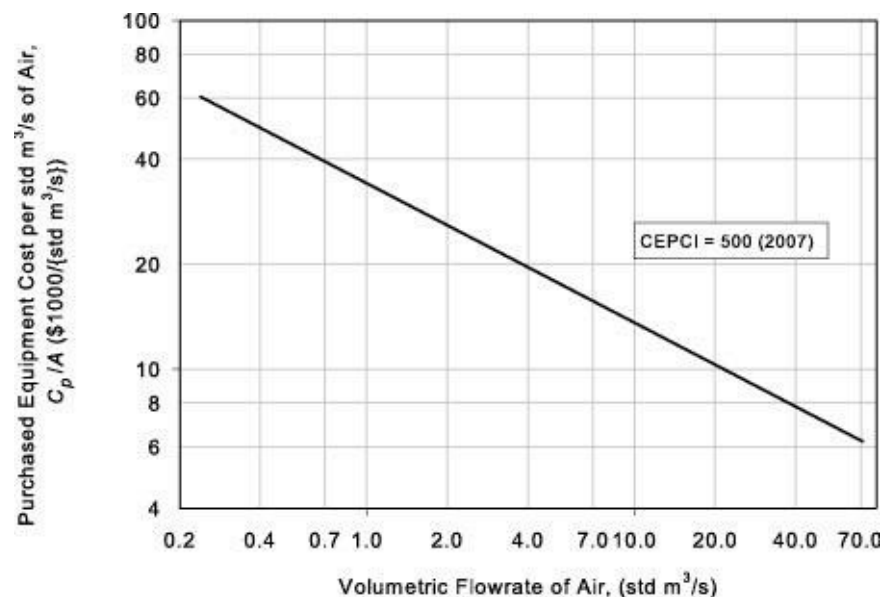
Another way to think of the economy of scale is to consider the purchased cost of equipment per unit capacity. [Equation 7.2](#) can be rearranged to give the following relationship:

(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  $100 \text{ m}^2$  was \$10,000.

Determine

- a. The constant  $K$  in [Equation 7.2](#)
- b. The cost of a new heat exchanger with area equal to  $180 \text{ m}^2$

From [Table 7.3](#):  $n = 0.59$ : for [Equation 7.2](#):

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

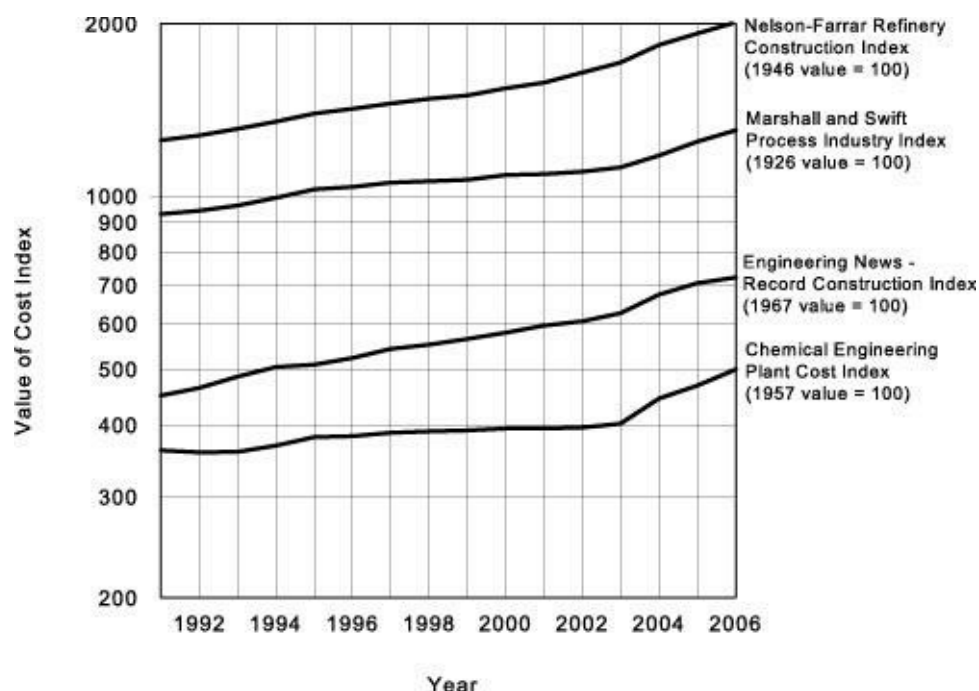
$I$  = Cost index

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](#).

**Table 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	<u>10</u>	
	100	61% of total
Erection and installation labor		22
Buildings, materials, and labor		7
Engineering and supervision		<u>10</u>
Total		100

### Example 7.6

The purchased cost of a heat exchanger of 500 m<sup>2</sup> area in 1992 was \$25,000.

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

From <a href="#">Table 7.4</a>	1992	2006
Marshall and Swift Index	943	1302

Chemical Engineering Plant Cost Index      358                      500

a. Marshal and Swift:                       $\text{Cost} = (\$25,000)(1302/943) = \$34,518$   
Chemical Engineering:

$$\text{Cost} = (\$25,000)(500/358) = \$34,916$$

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.

**Table 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
<b>1. Direct Project Expenses</b>		
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).
<b>2. Indirect Project Expenses</b>		
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.

Factor Associated with the Installation of Equipment	Symbol	Comments
<b>3. Contingency and Fee</b>		
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.
<b>4. Auxiliary Facilities</b>		
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).

$$\begin{aligned}\text{Cost in 2007} &= (\text{Cost in 1992})(\text{Capacity Correction})(\text{Inflation Correction}) \\ &= (\$23,000,000)(50,000/30,000)^{0.6}(500/358)\end{aligned}$$

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

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.

**Table 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.

$$\text{Capital Costs} = (\$6,800,000)(4.74) = \$32,232,000$$

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  $C_{BM}^0$  and  $F_{BM}^0$  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](#).

**Table 7.8 Equations for Evaluating Direct, Indirect, Contingency, and Fee Costs**

Factor	Basic Equation	Multiplying Factor to Be Used with Purchased Cost, $C_p^0$
<b>1. Direct</b>		
a. Equipment	$C_p^0 = C_p^0$	1.0
b. Materials	$C_M = \alpha_M C_p^0$	$\alpha_M$
c. Labor	$C_L = \alpha_L (C_p^0 + C_M)$	$(1.0 + \alpha_M)\alpha_L$
<b>Total Direct</b>	$C_{DE} = C_p^0 + C_M + C_L$	$(1.0 + \alpha_M)(1.0 + \alpha_L)$
<b>2. Indirect</b>		
a. Freight, etc.	$C_{FIT} = \alpha_{FIT} (C_p^0 + C_M)$	$(1.0 + \alpha_M)\alpha_{FIT}$
b. Overhead	$C_O = \alpha_O C_L$	$(1.0 + \alpha_M)\alpha_L\alpha_O$
c. Engineering	$C_E = \alpha_E (C_p^0 + C_M)$	$(1.0 + \alpha_M)\alpha_E$
<b>Total Indirect</b>	$C_{IDE} = C_{FIT} + C_O + C_E$	$(1.0 + \alpha_M)(\alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)$
<b>Bare Module</b>	$C_{BM}^0 = C_{IDE} + C_{DE}$	$(1.0 + \alpha_M)(1.0 + \alpha_L + \alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)$
<b>3. Contingency and Fee</b>		
a. Contingency	$C_{Cont} = \alpha_{Cont} C_{BM}^0$	$(1.0 + \alpha_M)(1.0 + \alpha_L + \alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)\alpha_{Cont}$
b. Fee	$C_{Fee} = \alpha_{Fee} C_{BM}^0$	$(1.0 + \alpha_M)(1.0 + \alpha_L + \alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)\alpha_{Fee}$
<b>Total Module</b>	$C_{TM} = C_{BM}^0 + C_{Cont} + C_{Fee}$	$(1.0 + \alpha_M)(1.0 + \alpha_L + \alpha_{FIT} + \alpha_L\alpha_O + \alpha_E)(1.0 + \alpha_{Cont} + \alpha_{Fee})$

**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,  $\alpha_i$ . 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  $C_p^0$  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:

- Bare module cost factor,  $F_{BM}^o$
- 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	$\alpha_M$	0.714
Labor	63.0	$\alpha_L$	$0.63/(1 + 0.714) = 0.368$
Freight	8.0	$\alpha_{FIT}$	$0.08/(1 + 0.714) = 0.047$
Overhead	63.4	$\alpha_O$	$0.634/0.368/(1 + 0.714) = 1.005$
Engineering	23.3	$\alpha_E$	$0.233/(1 + 0.714) = 0.136$
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](#),

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

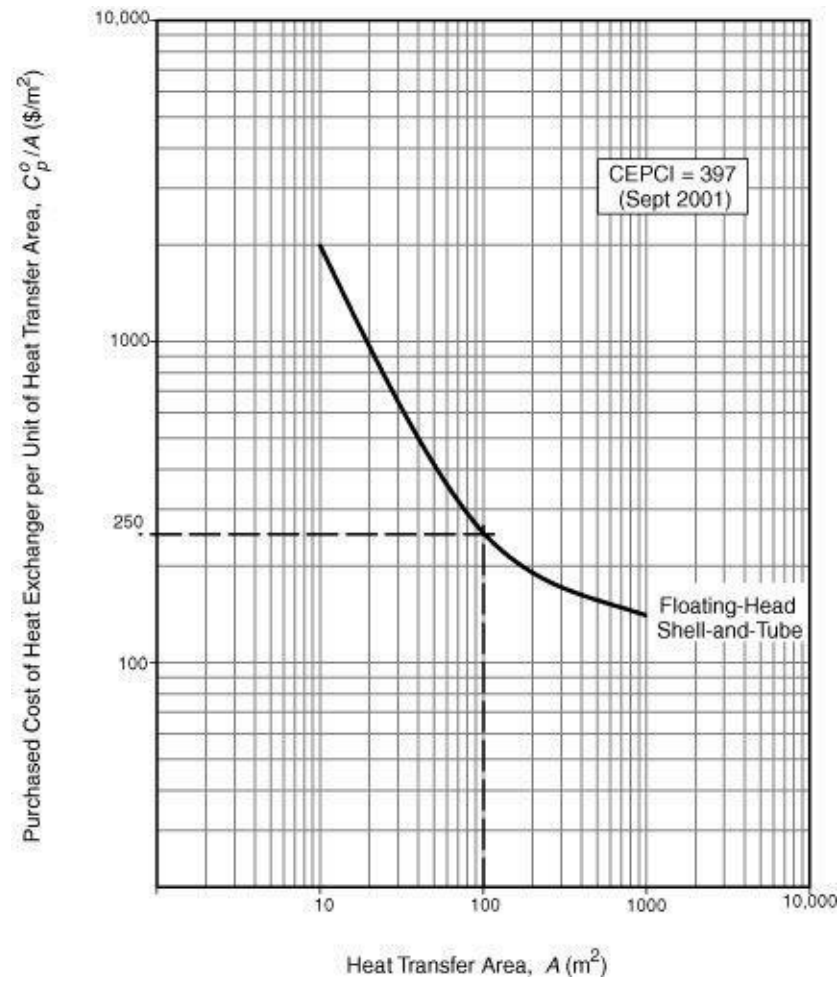
Fortunately, we do not have to repeat the procedure illustrated in [Example 7.9](#) in order to estimate  $F_{BM}^o$  for every piece of equipment. This has already been done for a large number of equipment modules, and the results are given in [Appendix A](#).

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<sup>2</sup> 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

$$C_{BM}^o(2001) = C_p^o(2001)[1.63 + 1.66(F_p = 1)(F_M = 1)] = 3.29C_p^o = (3.29)(\$25,000) = \$82,300$$

$$C_{BM}^o(2006) = C_{BM}^o(2001) (500/394) = \$82,300 (500/397) = \$103,590$$

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  $\alpha_i$ , 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,  $F_{BM}^o$ , 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

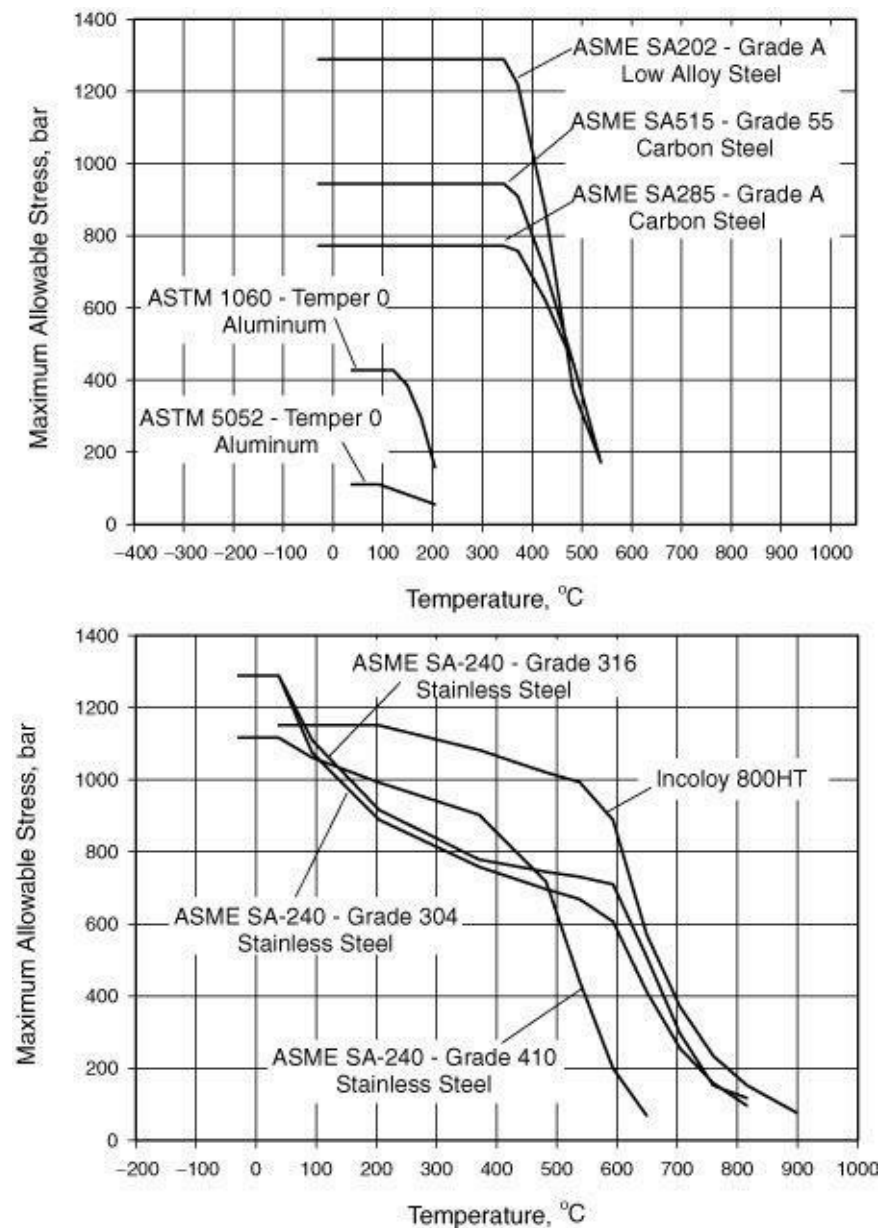
(7.9)

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

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](#).

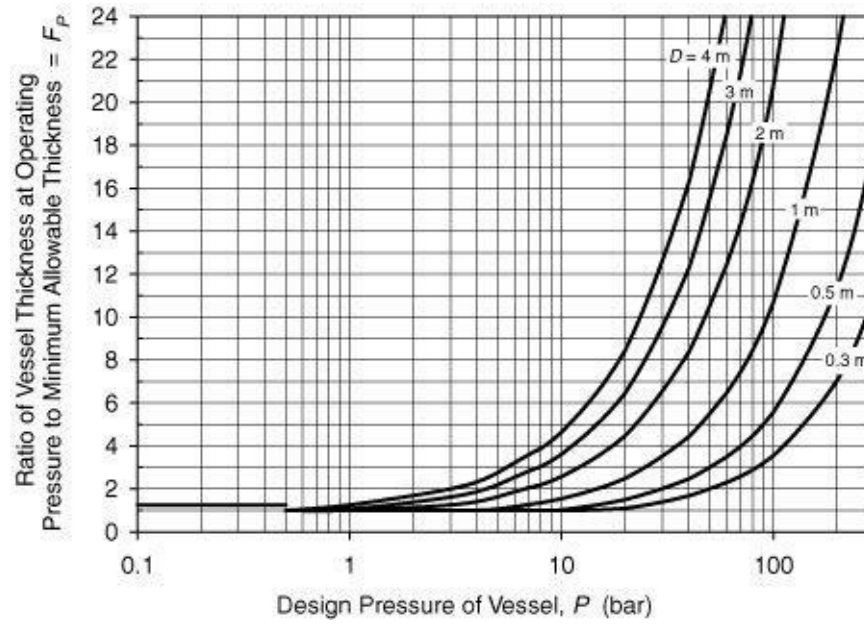
**Figure 7.5** Maximum Allowable Stresses for Materials of Construction as a Function of Operating Temperature (Data from Perry et al. [3], [Chapter 10](#), and Ref [15])



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  $1.67SE$ , 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)

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

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

- Repeat [Example 7.10](#) except consider the case when the operating pressures in both the shell- and the tube-side are 100 barg.
- 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

- 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](#):

$$C_{BM}(2001) = C_p^o(2001)[B_1 + B_2 F_p F_M] = \$25,000[1.63 + 1.66(1.383)(1.0)] = \$98,100$$
$$C_{BM}(2006) = \$98,100 (500/397) = \$123,590$$

- 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 <sub>2</sub> O <sub>2</sub> , 30%	C	C	A	C	C	N	C	A	A	A	A
H <sub>2</sub> O <sub>2</sub> , 90%	C	C	A	C	C	N		A	A	A	A
H <sub>2</sub> S, 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

Chemical Component	Carbon Steel	304 Stainless Steel	316 Stainless Steel	Aluminum	Copper	Brass	Monel	Hastelloy C	Titanium	TFE	Graphite
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
Tartaric 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 ( $\text{NH}_4^+$ ) 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, $W_{vessel}$
<b>Direct Expenses</b>	
Cost of raw materials	$\beta_{RM} W_{vessel}$
Machining costs	$\beta_{MC} W_{vessel}$
Labor costs	$\beta_L W_{vessel}$
<b>Indirect Costs</b>	
Overhead	$\beta_{OH} \beta_L W_{vessel}$
Engineering expenses	$\beta_E (\beta_{RM} + \beta_{MC}) W_{vessel}$
<b>Contingencies</b>	$\beta_{Cont} W_{vessel}$
Total manufacturing cost	$[\beta_{RM} + \beta_{MC} + \beta_L + \beta_{OH} \beta_L + \beta_E (\beta_{RM} + \beta_{MC}) + \beta_{Cont}] W_{vessel}$

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  $\beta_{RM}$ . Clearly, as the raw material costs increase, the total manufacturing costs will not increase proportionally to  $\beta_{RM}$ . 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<sup>2</sup> for the following cases.

- 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.
- 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  $C_p^o(2006) = \$25,000 (500/397) = \$31,490$ .

- 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](#),

$$C_{BM}(2006) = C_p^o[B_1 + B_2F_pF_M] = \$31,490[1.63 + 1.66(1.0)(2.73)] = \$194,000$$

- 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)

$$\begin{aligned} \text{Bare module cost, } C_{BM} &= C_p^o F_P F_M + C_p^o (F_{BM}^o - 1) + C_p^o (F_P F_M - 1) f_{P\&I} \\ &= C_p^o [F_P F_M (1 + f_{P\&I}) + F_{BM}^o - 1 - f_{P\&I}] = C_p^o [B_1 + B_2 F_P F_M] \end{aligned}$$

[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,

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

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](#),

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

$$N = 40$$

$$f_q = 1.0 \text{ (since number of trays} > 20, \text{ [Table A.5](#))}$$

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

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

**For the tower plus trays,**

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

### 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)

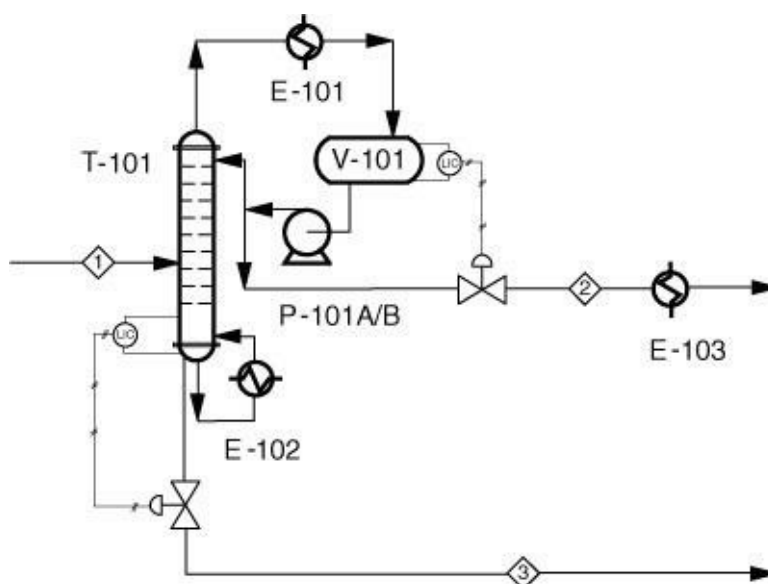
$$C_{GR} = C_{TM} + 0.50 \sum_{i=1}^n C_{BM,i}^o$$

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) <sup>†</sup>
E-101 Overhead condenser	Area = 170 m <sup>2</sup> Shell and tube (floating-head)	Tube - CS Shell - CS	Tube = 5.0 Shell = 5.0
E-102 Reboiler	Area = 205 m <sup>2</sup> Shell and tube (floating-head)	Tube - SS Shell - CS	Tube = 18.0 Shell = 6.0
E-103 Product cooler	Area = 10 m <sup>2</sup> (double pipe)	All CS construction	Inner = 5.0 Outer = 5.0
P-101A/B Reflux pumps	Power <sub>shaft</sub> = 5 kW Centrifugal	CS	Discharge = 5.0
T-101 Aromatics column	Diameter = 2.1 m Height = 23 m	Vessel - CS	Column = 5.0
V-101 Reflux drum	32 sieve trays Diameter = 1.8 m Length = 6 m Horizontal	Trays - SS Vessel - CS	Vessel = 5.0

\*CS = Carbon steel; SS = Stainless steel

<sup>†</sup>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.

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

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

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.

## References

1. Pikulik, A., and H. E. Diaz, "Cost Estimating for Major Process Equipment," *Chem. Eng.* 84, no. 21 (1977): 106.
2. Peters, M. S., and K. D. Timmerhaus, *Plant Design and Economics for Chemical Engineers*, 4th ed. (New York: McGraw-Hill, 1991).
3. Perry, R. H., D. W. Green, and J. O. Maloney, eds., *Chemical Engineers Handbook*, 7th ed. (New York: McGraw-Hill, 1997).
4. Cost Estimate Classification System, AACE International Recommended Practice No. 17R-97, 1997.
5. Ulrich, G. D., *A Guide to Chemical Engineering Process Design and Economics* (New York: John Wiley and Sons, 1984).
6. Lang, H. J., "Engineering Approach to Preliminary Cost Estimates," *Chem. Eng.* 54, no. 9 (1947): 130.

7. Lang, H. J., "Cost Relationships in Preliminary Cost Estimates," *Chem. Eng.* 54, no. 10 (1947): 117.
8. Lang, H. J., "Simplified Approach to Preliminary Cost Estimates," *Chem. Eng.* 55, no. 6 (1948): 112.
9. Guthrie, K. M., "Capital Cost Estimating," *Chem. Eng.* 76, no. 3 (1969): 114.
10. Guthrie, K. M., *Process Plant Estimating, Evaluation and Control* (Solana Beach, CA: Solana, 1974).
11. Navarrete, P. F., *Planning, Estimating, and Control of Chemical Construction Projects* (New York: Marcel Dekker, Inc., 1995).
12. R-Books Software, Richardson Engineering Services, Inc., 2001.
13. Section VIII, *ASME Boiler and Pressure Vessel Code*, ASME Boiler and Pressure Vessel Committee (New York: ASME, 2000).
14. Sandler, H. J., and E. T. Luckiewicz, *Practical Process Engineering, a Working Approach to Plant Design* (Philadelphia: XIMIX, Inc., 1987).
15. Incoloy Alloys 800 and 800HT, Table 22, Inco Alloys International Publication, IAI-20 4M US/1M UK (1986).
16. Construction Economics Section, *Engineering News Record*, December 24, 2001, 26.
17. Miller, C. A., "Factor Estimating Refined for Appropriation of Funds," *Amer. Assoc. Cost Engin. Bull.*, September 1965, 92.

## 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.

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<sup>3</sup> capacity purchased today (assume the current CEPCI = 500).

10.

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

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.

11.

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

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

12. 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 <sup>3</sup> \$)
???	1000	645.93
2000	500	500.00
2006	775	811.68

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

- 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.
13. 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.

- 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.
14. 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).***

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

15. 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<sup>2</sup>

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?

- 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<sup>2</sup>. Do the following.
17. 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<sup>2</sup>.  
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)?

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  $CEPCI = 500$ .

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
19. 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/>