

# Capital cost estimating

## KEY LEARNING OBJECTIVES

- Techniques for estimating the capital cost of a process
- How to estimate the costs of major process equipment
- How to use commercial cost estimating software
- How to update cost data
- How to estimate cost data for international projects
- How to estimate costs for proprietary equipment when data are scarce

## 7.1 Introduction

Most chemical engineering design projects are carried out to provide information from which estimates of capital and operating costs can be made. Chemical plants are built to make a profit, and an estimate of the investment required is needed before the profitability of a project can be assessed. Cost estimation is a specialized subject and a profession in its own right, but the design engineer must be able to make rough cost estimates to decide between project alternatives and optimize the design.

This chapter introduces the components of capital costs and the techniques used for capital cost estimating. Simple costing methods and some cost data are given, which can be used to make preliminary estimates of capital costs in the early stages of design. Sources of cost data and methods for updating cost estimates are described. The more sophisticated software that is used in industry for preliminary estimating is discussed in Section 7.10.

For a more detailed treatment of the subject, the reader should refer to the numerous specialized texts that have been published on cost estimation. The following books are particularly recommended: [Happle and Jordan \(1975\)](#), [Guthrie \(1974\)](#), [Page \(1996\)](#), [Garrett \(1989\)](#), [Humphreys \(1991, 2005\)](#).

## 7.2 Components of capital cost

This section introduces the components of project costs.

### 7.2.1 Fixed capital investment

The fixed capital investment is the total cost of designing, constructing, and installing a plant and the associated modifications needed to prepare the plant site. The fixed capital investment is made up of:

1. The inside battery limits (ISBL) investment – the cost of the plant itself
2. The modifications and improvements that must be made to the site infrastructure, known as *offsite battery limits* (OSBL) investment
3. Engineering and construction costs
4. Contingency charges

### **ISBL plant costs**

The ISBL plant cost includes the cost of procuring and installing all the process equipment that makes up the new plant.

The direct field costs include:

1. All the major process equipment, such as vessels, reactors, columns, furnaces, heat exchangers, coolers, pumps, compressors, motors, fans, turbines, filters, centrifuges, driers, etc., including field fabrication and testing if necessary.
2. Bulk items, such as piping, valves, wiring, instruments, structures, insulation, paint, lube oils, solvents, catalysts, etc.
3. Civil works such as roads, foundations, piling, buildings, sewers, ditches, bunds, etc.
4. Installation labor and supervision.

In addition to the direct field costs there will be indirect field costs, including:

1. Construction costs such as construction equipment rental, temporary construction (rigging, trailers, etc.), temporary water and power, construction workshops, etc.
2. Field expenses and services such as field canteens, specialists' costs, overtime pay, and adverse weather costs.
3. Construction insurance.
4. Labor benefits and burdens (social security, workers compensation, etc.).
5. Miscellaneous overhead items such as agent's fees, legal costs, import duties, special freight costs, local taxes, patent fees or royalties, corporate overheads, etc.

In the early stages of a project it is important to define the ISBL scope carefully, as other project costs are often estimated from ISBL cost. The overall project economics can be badly miscalculated if the ISBL scope is poorly defined. Several methods for estimating ISBL costs are given in the following sections of this chapter.

### **Offsite costs**

Offsite cost or OSBL investment includes the costs of the additions that must be made to the site infrastructure to accommodate adding a new plant or increasing the capacity of an existing plant. Offsite investments may include:

- Electric main substations, transformers, switchgear, and power lines
- Power generation plants, turbine engines, and standby generators
- Boilers, steam mains, condensate lines, boiler feed water treatment plant, and supply pumps
- Cooling towers, circulation pumps, cooling water mains, and cooling water treatment
- Water pipes, water demineralization, waste water treatment plant, site drainage, and sewers
- Air separation plants to provide site nitrogen for inert gas, as well as nitrogen lines
- Driers and blowers for instrument air, as well as instrument air lines
- Pipe bridges, feed, and product pipelines
- Tanker farms, loading facilities, silos, conveyors, docks, warehouses, railroads, and lift trucks
- Laboratories and analytical equipment
- Offices, canteens, changing rooms, and central control rooms
- Workshops and maintenance facilities
- Emergency services, firefighting equipment, fire hydrants, medical facilities, etc.
- Site security, fencing, gatehouses, and landscaping

Offsite investments often involve interactions with utility companies such as electricity or water suppliers. They may be subject to equal or greater scrutiny than ISBL investments because of their impact on the local community through water consumption and discharge, traffic, and other similar factors.

Offsite costs are typically estimated as a proportion of ISBL costs in the early stages of design. Offsite costs are usually in the range from 10% to 100% of ISBL costs, depending on the project scope and its impact on site infrastructure. For typical petrochemical projects, offsite costs are usually between 20% and 50% of ISBL cost, and 40% is usually used as an initial estimate if no details of the site are known. For an established site with well-developed

infrastructure, offsite costs will generally be lower. This is particularly true of sites that have undergone contraction, where some plants have closed, leaving underutilized infrastructure (“brownfield” sites). On the other hand, if the site infrastructure is in need of repair or upgrading to meet new regulations, or if the plant is built on a completely new site (a “greenfield” site), then offsite costs will be higher. Offsite costs are discussed in more detail in Section 7.9.

Once a site has been chosen for the project, the modifications to the site infrastructure that are needed can be designed in detail in the same manner as the ISBL investments. Infrastructure upgrades are usually the first part of a project to be implemented, as they usually need to be commissioned before the plant can begin operation.

### **Engineering costs**

The engineering costs, sometimes referred to as *home office costs* or *contractor charges*, include the costs of detailed design and other engineering services required to carry out the project:

1. Detailed design engineering of process equipment, piping systems, control systems and offsites, plant layout, drafting, cost engineering, scale models, and civil engineering.
2. Procurement of main plant items and bulks.
3. Construction supervision and services.
4. Administrative charges, including engineering supervision, project management, expediting, inspection, travel and living expenses, and home office overheads.
5. Bonding.
6. Contractor’s profit.

Very few operating companies retain a large enough engineering staff to carry out all of these activities internally, except for very small projects. In most cases, one or more of the major engineering contracting firms will be brought in.

Engineering costs are best estimated individually based on project scope, as they are not directly proportional to project size. A rule of thumb for engineering costs is 30% of ISBL plus OSBL cost for smaller projects and 10% of ISBL plus OSBL cost for larger projects. The actual charges paid for real industrial projects vary considerably from customer to customer and are strongly influenced by long-term client–contractor relationships and overall market demand for engineering services. Customers usually have to pay premiums or surcharges if they want to complete a project on an accelerated timeline or if they make a lot of changes once a project is underway.

### **Contingency charges**

Contingency charges are extra costs added into the project budget to allow for variation from the cost estimate. All cost estimates are uncertain (see Section 7.3.1), and the final installed cost of many items is not known until installation has been successfully completed. Apart from errors in the cost estimate, contingency costs also help cover:

- Minor changes in project scope
- Changes in prices (e.g., prices of steel, copper, catalyst, etc.)
- Currency fluctuations
- Labor disputes
- Subcontractor problems
- Other unexpected problems

A contingency charge can be thought of as an additional fee charged by the Engineering, Procurement and Construction (EPC) company to address the risk that the project will run over budget. Contingency charges therefore reduce the likelihood that the contractor will lose money on a fixed-price bid. A minimum contingency charge of 10% of ISBL plus OSBL cost should be used on all projects. If the technology is uncertain, higher contingency charges (up to 50%) are used. Contingency charges are discussed in more detail in Section 9.8.4.

## **7.2.2 Working capital**

In addition to the fixed capital investment that was used to design and construct the plant, the owner needs to invest some capital in maintaining plant operations. The capital that is tied up in maintaining inventories of feeds, products, and spare parts, together with cash on hand and the difference between money owed by costumers (accounts receivable) and money owed to suppliers (accounts payable), is termed the *working capital* of the plant. Working capital is required as long as the plant is in operation, but is recovered if the plant is shut down. Working capital is discussed in more detail in the chapter on project finance and economics; see Section 9.2.3.

### 7.3 Accuracy and purpose of capital cost estimates

The accuracy of an estimate depends on the amount of design detail available, the accuracy of the cost data available, and the time spent on preparing the estimate. In the early stages of a project only an approximate estimate will be justified based on the amount of information available.

#### 7.3.1 AACE International cost estimate classes

The Association for the Advancement of Cost Estimating International (AACE International) is the professional association representing the cost engineering profession in the United States. AACE International classifies capital cost estimates into five types according to their accuracy and purpose:

1. Order of magnitude estimates (“ballpark estimate,” “guesstimate” “Class 5 estimate”), accuracy typically  $\pm 30\%$  to  $50\%$ , usually based on the costs of similar processes and requiring essentially no design information. These are used in initial feasibility studies and for screening purposes.
2. Preliminary (“approximate,” “study,” “feasibility,” “Class 4”) estimates, accuracy typically  $\pm 30\%$ , which are used to make coarse choices between design alternatives. They are based on limited cost data and design detail.
3. Definitive (“authorization,” “budgeting,” “control,” “Class 3”) estimates, accuracy typically  $\pm 10\%$  to  $15\%$ . These are used for the authorization of funds to proceed with the design to the point where an accurate and more detailed estimate can be made. Authorization may also include funds to cover cancellation charges on any long delivery equipment ordered at this stage of the design to avoid delay in the project. In a contracting organization this type of estimate could be used with a large contingency factor to obtain a price for tendering a bid. Normally, however, an accuracy of about  $\pm 5\%$  would be needed and a more detailed estimate would be made if time permitted. With experience, and where a company has cost data available from similar projects, estimates of acceptable accuracy can be made at the flowsheet stage of the project. A rough P&I diagram and the approximate sizes of the major items of equipment would also be needed.
4. Detailed estimates (“quotation,” “tender,” “firm estimate,” “contractor’s estimate,” “Class 2 estimate”), accuracy  $\pm 5\%$  to  $10\%$ , which are used for project cost control and estimates for fixed-price contracts. Estimates of this quality can be made once the front-end engineering design (FEED) is completed, including a complete (or near-complete) process design, firm quotes for equipment, and a detailed breakdown and estimation of the construction cost. By this stage, the contractor can usually present a list of all the items that must be purchased and can make a firm commitment to the client.
5. Check estimates (“tender,” “as-bid,” “Class 1 estimate”), accuracy  $\pm 5\%$  to  $10\%$ . This is based on a completed design and concluded negotiations on procurement of specialized items and long lead-time items.

It is important for design engineers to understand how much information is necessary to arrive at a given level of confidence in an estimate, particularly when working with professional cost engineers. It is not possible to generate a Class 1 or Class 2 estimate without completing the design to a sufficient level of detail. A Class 3 estimate may also require a substantially complete design package unless the company has recently built projects with similar scope. Class 4 and 5 estimates are relatively easy to generate but have high uncertainty, which must be taken into consideration in economic analysis and sensitivity analysis; see Section 9.8.

#### 7.3.2 Development of cost estimates

As a project proceeds from initial concept through detailed design to start-up, costs begin to be accumulated, particularly once procurement and construction get underway (Fig. 7.1a). At the same time, the ability of the design engineer to influence project cost decreases and is minimal by the time construction begins (Fig. 7.1b). There is therefore a strong incentive to try to estimate project costs at as early a stage as possible, even if the design information is incomplete, so that the project can be optimized, evaluated, and abandoned if it is not attractive.

It is difficult to go beyond the accuracy of a Class 4 estimate without making a fairly detailed design of the plant. The cost of preparing the estimate thus becomes the cost of designing the process and sizing and optimizing the main equipment items. To obtain even greater accuracy, the contractor needs to establish the plot plan and plant layout so as to make more accurate estimates of the amount of piping, wiring, and structural steel that will be needed. The cost of preparing an estimate thus increases from about  $0.1\%$  of the total project cost for  $\pm 30\%$  accuracy to about  $3\%$  for a detailed estimate with an accuracy of  $\pm 5\%$ .

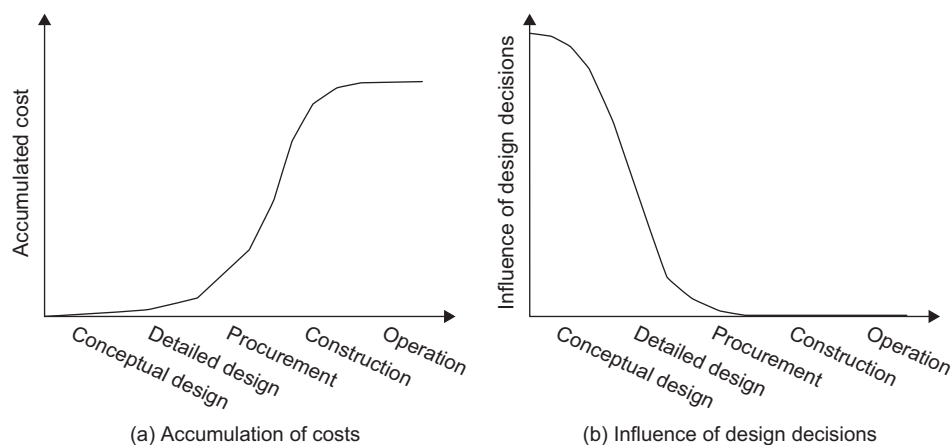


FIG. 7.1 Influence of design decisions on project cost.

## 7.4 Order-of-magnitude estimates

In the early stages of design, or when carrying out preliminary marketing studies, the design engineer may want to make a quick (Class 5) capital cost estimate without completing a plant design. Several shortcut methods have been developed that allow estimates of total plant cost to be made within  $\pm 50\%$  accuracy for preliminary studies. These methods can also be used to provide a rough check on more detailed estimates developed from process equipment costs later in the design of the process.

### 7.4.1 Cost curve methods

The quickest way to make an order-of-magnitude estimate of plant cost is to scale it from the known cost of an earlier plant that used the same technology or from published data. This requires no design information other than the production rate.

The capital cost of a plant can be related to capacity by the equation

$$C_2 = C_1 \left( \frac{S_2}{S_1} \right)^n \quad (7.1)$$

where  $C_2$  = ISBL capital cost of the plant with capacity  $S_2$

$C_1$  = ISBL capital cost of the plant with capacity  $S_1$

The exponent  $n$  is typically 0.8 to 0.9 for processes that use a lot of mechanical work or gas compression (e.g., methanol, paper pulping, solids-handling plants). For typical petrochemical processes  $n$  is usually about 0.7. For small-scale, highly instrumented processes, such as specialty chemical or pharmaceuticals manufacture,  $n$  is in the range of 0.4 to 0.5. Averaged across the whole chemical industry,  $n$  is about 0.6, and hence Equation 7.1 is commonly referred to as the *six-tenths rule*. The equation can be used with  $n = 0.6$  to get a rough estimate of the capital cost if there are not sufficient data available to calculate the index for the particular process. [Estrup \(1972\)](#) gives a critical review of the six-tenths rule. Equation 7.1 is only an approximation, and if sufficient data are available, the relationship is best represented on a log-log plot. [Garrett \(1989\)](#) has published capital cost-plant capacity curves for over 250 processes.

The journal *Hydrocarbon Processing* publishes supplements on refining, petrochemical, and gas processing processes every other year. These supplements are available in print or CD format to subscribers and give approximate capital cost data for various licensed processes, which can be fitted using a rearranged form of Equation 7.1:

$$C_2 = \frac{C_1}{S_1^n} \times S_2^n = a S_2^n \quad (7.2)$$

Values of the parameters  $a$  and  $n$  for some fuels and commodity chemical processes are given in [Table 7.1](#). Note that these correlations are only valid over a range of plant sizes between  $S_{lower}$  and  $S_{upper}$ , corresponding to the range over which the data were fitted. The costs in the *Hydrocarbon Processing* supplements are supplied by the technology vendors and are suitable for ballpark estimates only.

TABLE 7.1 Process cost correlations

Process	Licensor	Units	Capacity		<i>a</i>	<i>n</i>
			<i>S<sub>lower</sub></i>	<i>S<sub>upper</sub></i>		
ABS resin (15% Rubber) by emulsion polymerization	Generic	MMlb/y	50	300	12.146	0.6
Acetic acid by Cativa process	BP	MMlb/y	500	2000	3.474	0.6
Acetic acid by low water methanol carbonylation	Celanese	MMlb/y	500	2000	2.772	0.6
Acrolein by propylene oxidation with Bi/Mo catalyst	Generic	MMlb/y	30	150	6.809	0.6
Adipic acid from phenol	Generic	MMlb/y	300	1000	3.533	0.6
Alkylation (sulfuric acid effluent refrigeration process)	Stratco/DuPont	bpd	4000	20,000	0.160	0.6
Alkylation (HF process)	UOP	bpd	5000	12,000	0.153	0.6
Allyl chloride by propylene chlorination	Generic	MMlb/y	80	250	7.581	0.6
Alpha olefins (full range process)	Chevron Phillips	MMlb/y	400	1200	5.240	0.6
Alpha olefins (full range process)	Shell	MMlb/y	400	1000	8.146	0.6
Benzene by sulfolane extraction	UOP/Shell	MMgal/y	50	200	7.793	0.6
Benzene by toluene hydrodealkylation	Generic	MMgal/y	50	200	7.002	0.6
Benzene reduction by Bensat process	UOP	bpd	8000	15,000	0.0275	0.6
Biodiesel (FAME) from vegetable oil	Generic	MMlb/y	100	500	2.747	0.6
bis-HET by Eastman Glycolysis	Eastman	MMlb/y	50	200	0.500	0.6
BTX Aromatics by Cyclar process	BP/UOP	tpy	200,000	800,000	0.044	0.6
BTX Aromatics by CCR Platforming process	UOP	tpy	200,000	800,000	0.015	0.6
Butadiene by extractive distillation	UOP/BASF	MMlb/y	100	500	5.514	0.6
Butadiene by Oxo-D plus extractive distillation	Texas Petrochem.	MMlb/y	100	500	11.314	0.6
Butene-1 by alhabutol ethylene dimerization	Axens	tpy	5000	30,000	0.0251	0.6
Butene-1 by BP Process	BP	tpy	20,000	80,000	0.169	0.6

Caprolactam from nitration-grade toluene	SNIA BPD S.p.A.	tpy	40,000	120,000	0.321	0.6
Carbon monoxide by steam methane reforming	Generic	MMscf/y	2000	6000	0.363	0.6
Catalytic condensation for gasoline production	UOP	bpd	10,000	30,000	0.222	0.6
Catalytic reforming by CCR platforming process	UOP	bpd	15,000	60,000	0.179	0.6
Coking by Flexicoking, including fluid coking	ExxonMobil	bpd	15,000	40,000	0.343	0.6
Coking by selective yield delayed coking	Foster Wheeler/UOP	bpd	15,000	60,000	0.109	0.68
Copolymer polypropylene by INNOVENE	BP	MMlb/y	300	900	3.430	0.6
Copolymer polypropylene by Unipol	Dow	MMlb/y	300	900	3.641	0.6
Copolymer polypropylene by SPHERIPOL Bulk	Basell	MMlb/y	300	900	3.649	0.6
Copolymer polypropylene by BORSTAR	Borealis	MMlb/y	300	900	4.015	0.6
Crude distillation by D2000	TOTAL/Technip	bpd	150,000	300,000	0.151	0.6
Cumene by Q-Max process	UOP	tpy	150,000	450,000	0.0120	0.6
Cyclic olefin copolymer by Mitsui Process	Mitsui	MMlb/y	60	120	12.243	0.6
Cyclohexane by liq-phase hydrogenation of benzene	Axens	tpy	100,000	300,000	0.0061	0.6
Dewaxing by ISODEWAXING	Chevron Lummus	bpd	6,000	15,000	0.256	0.6
2,6-Dimethylnaphthalene by MeOH alkylation	Exxon Mobil/Kobe	MMlb/y	50	100	7.712	0.6
Dimethyl terephthalate by methanolysis	Generic	MMlb/y	30	80	5.173	0.6
Dimethyl terephthalate by Huels Oxidation	Huels	MMlb/y	300	800	7.511	0.6
Ethanol by ethylene hydration	Generic	Mgal/y	30	90	9.643	0.6
Ethanol (fuel grade) by corn dry milling	Generic	tpy	100,000	300,000	0.0865	0.6
Ethylbenzene by EBOne process	ABB Lummus/UOP	tpy	300,000	700,000	0.0085	0.6

Continued

TABLE 7.1 Process cost correlations—cont'd

Process	Licensor	Units	Capacity		<i>a</i>	<i>n</i>
			<i>S</i> <sub>lower</sub>	<i>S</i> <sub>upper</sub>		
Ethylene by ethane cracking	Generic	MMlb/y	500	2000	9.574	0.6
Ethylene by UOP/Hydro MTO process	UOP/INEOS	MMlb/y	500	2000	8.632	0.6
Ethylene: light naphtha cracker (max ethylene)	Generic	MMlb/y	1000	2000	16.411	0.6
Ethylene by ethane/propane cracker	Generic	MMlb/y	1000	2000	7.878	0.6
Ethylene by gas oil cracker	Generic	MMlb/y	1000	2000	17.117	0.6
Ethylene glycol via ethylene oxide hydrolysis	Shell	MMlb/y	500	1000	5.792	0.6
Expandable polystyrene by suspension process	Generic	MMlb/y	50	100	3.466	0.6
Fischer Tropsch process	ExxonMobil	tpy	200,000	700,000	0.476	0.6
Fluid catalytic cracking	KBR	bpd	20,000	60,000	0.210	0.6
Fluid catalytic cracking with power recovery	UOP	bpd	20,000	60,000	0.302	0.6
Gas to liquids by Syntroleum process	Syntroleum	bpd	30,000	100,000	2.279	0.6
Gas sweetening by Amine Guard FS process	UOP	MMscf/d	300	800	0.386	0.6
Gasification by GE gasification process Maya crude	GE Energy	bpd	7000	15,000	0.681	0.6
Gasoline desulfurization, ultra-deep by Prime-G+	Axens	bpd	7000	15,000	0.0420	0.58
Glucose (40% solution) by basic wet corn milling	Generic	MMlb/y	300	800	3.317	0.6
HDPE pellets by BP gas phase process	BP Amoco	MMlb/y	300	700	3.624	0.6
HDPE pellets by Phillips slurry process	Phillips	MMlb/y	300	700	3.370	0.6
HDPE pellets by Zeigler slurry process	Zeigler	MMlb/y	300	700	4.488	0.6
High-impact polystyrene by bulk polymerization	Dow	MMlb/y	70	160	2.970	0.6
Hydrocracking by isocracking	Chevron Lummus	bpd	20,000	45,000	0.221	0.6



Hydrocracking by unicracking (distillate) process	UOP	bpd	20,000	45,000	0.136	0.66
Hydrocracking	Axens	bpd	20,000	45,000	0.198	0.6
Hydrogen by steam methane reforming	Foster Wheeler	MMscf/d	10	50	1.759	0.79
Hydrotreating by Unionfining process	UOP	bpd	10,000	40,000	0.0532	0.68
Isomerization by once-through Penex process	UOP	bpd	8,000	15,000	0.0454	0.6
Isomerization by Penex-Molex process	UOP	bpd	8000	15,000	0.120	0.6
Isophthalic acid by m-xylene oxidation	Generic	MMlb/y	160	300	9.914	0.6
Isoprene via isobutylene carbonylation	IFP	MMlb/y	60	200	10.024	0.6
Isoprene by propylene dimerization and pyrolysis	Generic	MMlb/y	60	200	6.519	0.6
Linear alkylbenzene by PACOL/DeFine/PEP/Detal <sup>6</sup>	UOP	MMlb/y	100	250	4.896	0.6
Linear alpha olefins	Chevron	MMlb/y	300	700	5.198	0.6
Linear alpha olefins by Linear-1 process	UOP	tpy	200,000	300,000	0.122	0.6
Maleic anhydride by fluid bed process	Generic	MMlb/y	70	150	7.957	0.6
Methacrylic acid by isobutylene oxidation	Generic	MMlb/y	70	150	7.691	0.6
Methanol via steam reforming and synthesis	Davy Process Tech.	tpd	3000	7000	2.775	0.6
m-Xylene by MX Sorbex process	UOP	MMlb/y	150	300	4.326	0.6
Naphthalene by three-stage fractional crystallizer	Generic	MMlb/y	20	50	2.375	0.6
N-Butanol from crude C4s	BASF	MMlb/y	150	300	8.236	0.6
Norbornene by Diels–Alder reaction	Generic	MMlb/y	40	90	7.482	0.6
Pentaerythritol by condensation	Generic	MMlb/y	40	90	6.220	0.6
PET resin chip with comonomer by NG3	DuPont	MMlb/y	150	300	4.755	0.6

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TABLE 7.1 Process cost correlations—cont'd

Process	Licensor	Units	Capacity		<i>a</i>	<i>n</i>
			<i>S</i> <sub>lower</sub>	<i>S</i> <sub>upper</sub>		
Phenol from cumene (zeolite catalyst)	UOP/Sunoco	MMlb/y	200	600	6.192	0.6
Phthalic anhydride by catalytic oxidation	Generic	MMlb/y	100	200	7.203	0.6
Polycarbonate by interfacial polymerization	Generic	MMlb/y	70	150	20.680	0.6
Polyethylene terephthalate (melt phase)	Generic	MMlb/y	70	200	5.389	0.6
Polystyrene by bulk polymerization, plug flow	Generic	MMlb/y	70	200	2.551	0.6
Propylene by Oleflex process	UOP	tpy	150,000	350,000	0.0943	0.6
Propylene by metathesis	Generic	MMlb/y	500	1,000	1.899	0.6
Purified terephthalic acid	EniChem/Technimont	MMlb/y	350	700	10.599	0.6
p-Xylene by Isomar and Parex processes	UOP	tpy	300,000	700,000	0.0230	0.6
p-Xylene by Tatoray process	UOP	bpd	12,000	20,000	0.0690	0.6
Refined glycerin by distillation/adsorption	Generic	MMlb/y	30	60	2.878	0.6
Sebacic acid by cyclododecanone route	Sumitomo	MMlb/y	8	16	13.445	0.6
Sorbitol (70%) by continuous hydrogenation	Generic	MMlb/y	50	120	4.444	0.6
Styrene by SMART process	ABB Lummus/UOP	tpy	300,000	700,000	0.0355	0.6
Vinyl acetate by Cativa integrated process	BP	MMlb/y	300	800	7.597	0.6
Vinyl acetate by Celanese Vantage Process	Celanese	MMlb/y	300	800	6.647	0.6
Visbreaking by coil-type visbreaker	Foster Wheeler/UOP	bpd	6,000	15,000	0.278	0.48

## Notes:

1. Values of *a* are in January 2006 MM\$ on a USGC basis (Nelson Farrer index = 1961.6, CEPCI index = 478.6).
2. *S*<sub>lower</sub> and *S*<sub>upper</sub> indicate the bounds of the region over which the correlation can be applied.
3. *S* is based on product rate for chemicals, feed rate for fuels.
4. If the index *n* is 0.6, then the correlation is an extrapolation around a single cost point.
5. Correlations are based on data taken from [Hydrocarbon Processing \(2003, 2004a,b\)](#), except where the licensor is stated as "Generic," in which case the correlations are based on data from Nexant PERP reports (see [www.Nexant.com/products](http://www.Nexant.com/products) for a full list of reports available).
6. PACOL/DeFine/PEP/Detal processes.

### Economy of scale

It can be seen from the discussion earlier and the correlations in Table 7.1 that the exponent  $n$  is always less than 1.0. One result of this is that larger plants tend to cost less to construct per unit of product produced, an advantage that is known as an *economy of scale*. From Equation 7.2, the capital cost per unit of production of product will be:

$$\frac{C_2}{S_2} = a S_2^{n-1} \quad (7.3)$$

As  $n - 1$  is less than zero, the cost per unit of production decreases as  $S_2$  increases. A smaller capital cost per unit of product allows the owners of large plants to price their products more competitively and still recover their capital investment. This creates an incentive for chemical companies to build plants at larger scale.

### 7.4.2 Step count method

If cost data for a similar process are not available then an order-of-magnitude estimate can sometimes be made by adding contributions for different plant sections or functional units.

Experienced design engineers can often figure out costs of plant sections from historic total plant costs. For example, in many petrochemical processes roughly 20% of ISBL capital cost is in the reactor section and 80% is in the distillation and product purification sections. If the separation and recovery system has a similar size and complexity to that of a known process, then the capital cost can be estimated based on the cost of the known process and the reactor section cost taken as one-quarter of the separation section cost.

An alternative approach is Bridgewater's method, which correlates plant cost to the number of processing steps (Bridgewater & Mumford, 1979). For plants primarily processing liquids and solids:

$$Q \geq 60,000: C = 4320 N \left( \frac{Q}{s} \right)^{0.675} \quad (7.4)$$

$$Q < 60,000: C = 380,000 N \left( \frac{Q}{s} \right)^{0.3} \quad (7.5)$$

where:  $C$  = ISBL capital cost in US\$, US Gulf Coast, January 2010 basis (CEPCI = 532.9)

$Q$  = plant capacity in metric tons per year

$s$  = reactor conversion (= mass of desired product per mass fed to the reactor)

$N$  = number of functional units

(Note: The correlations have been updated from the original reference.)

A functional unit includes all the equipment and ancillaries needed for a significant process step or function, such as a reaction, separation, or other major unit operation. Pumping and heat exchange are not normally considered functional units unless they have substantial cost, for example, compressors, refrigeration systems, or process furnaces.

### 7.4.3 Reverse engineering methods

In some situations, a very rough estimate of capital cost can be backed out from operating costs or product prices.

#### Payback method

The difference between the sales price and the cost of production is the gross profit of the plant; see Section 8.2.4. If an investor were to build a new plant, then the gross profit minus taxes would have to be sufficient to give a return on the investment made in building the plant. Assuming the plant pays back in 3 to 5 years (average = 4 years), a very rough estimate of the plant cost would be:

$$\text{Plant cost} = 4 \times (\text{gross profit} - \text{taxes}) \quad (7.6)$$

For chemicals, raw materials costs are typically 80% to 90% of the cash cost of production, so the gross profit can be roughly estimated as:

$$\text{Gross profit} = \text{product value} - (1.2 \times \text{raw materials costs}) \quad (7.7)$$

This method obviously can only be used in situations where the designer believes that new plants are being built and generating a reasonable return on capital.

### **Turnover ratio method**

An even simpler (and less accurate) method is to estimate the capital investment based on the turnover ratio. The turnover ratio is the gross annual sales divided by the fixed capital investment. The turnover ratio can vary widely, but a typical value for the chemical industry is between 1.0 and 1.25 (Humphreys, 1991).

### **TCOP method**

For large-scale production of formed or assembled components (>500,000 pieces per year) a rule of thumb is:

$$\text{Total cost of production (TCOP)} = 2 \times \text{raw materials cost} \quad (7.8)$$

The TCOP is equal to the cost of raw materials, plus utility costs, fixed costs, and an annualized capital charge that is typically one-third to one-fifth of the capital cost (see Section 8.2.4, Equation 8.6). Equation 7.8 can thus be used to make a very approximate estimate of plant cost for manufactured products if fixed costs and utilities can be estimated.

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### **Example 7.1**

The process for making cyclohexane by saturation of benzene consists of a feed-effluent heat exchanger, a saturation reactor, and a product stabilizer column. Estimate the cost of a plant that produces 200,000 metric tons per year (200 kMTA) of cyclohexane using the correlation in Table 7.1 and Bridgewater's method.

#### **Solution**

From Table 7.1, the cost correlation for the Axens process for benzene saturation gives:

$$\begin{aligned} C &= 0.0061(S)^{0.6} \\ &= 0.0061(2 \times 10^5)^{0.6} \\ &= \underline{\underline{\$9.2 \text{ MM}}} \text{ expressed on a Jan. 2006 USGC basis} \end{aligned}$$

Using Bridgewater's method, we have two functional units (the reactor and product stabilizer—the heat exchanger doesn't count), and assuming that the reactor conversion is 1.0, we can substitute into Equation 7.4:

$$\begin{aligned} C &= 4320 \times 2(Q)^{0.675} \\ &= 4320 \times 2 \times (2 \times 10^5)^{0.675} \\ &= \underline{\underline{\$33 \text{ MM}}} \text{ expressed on a Jan. 2010 USGC basis} \end{aligned}$$

Note that we have obtained two very different answers. Bridgewater's correlation is known to be only an approximation; however, Table 7.1 is based on data from technology vendors that may be somewhat understated. With the level of information available, it is probably safe to say that the cost is in the range \$10 MM to \$20 MM. Note also that the costs are not on the same time basis. Methods for correcting costs on different time bases will be discussed in Section 7.7.

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## **7.5 Estimating purchased equipment costs**

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When more design information is available, the cost of a plant can be worked up from the cost of individual items of process equipment. Costs of single pieces of equipment are also often needed for minor revamp and de-bottlenecking projects.

### 7.5.1 Sources of equipment cost data

The best source of purchased equipment costs is recent data on actual prices paid for similar equipment. Engineers working for engineering, procurement, and construction (EPC) companies (often referred to as *contractors*) have access to large amounts of high-quality data, as these companies carry out many projects globally every year. Engineers working in operating companies may have access to data from recent projects, but unless they work for a large company that carries out many capital projects, they are unlikely to be able to develop and maintain current cost correlations for more than a few basic equipment types. Most large companies recognize the difficulty of making reliable cost estimates and employ a few experienced cost engineering specialists who collect data and work closely with the EPC companies on project budgets.

Actual prices paid for equipment and bulk items may differ substantially from catalog or list prices, depending on the purchasing power of the contractor or client and the urgency of the project. Discounts and surcharges are highly confidential business information and will be closely guarded even within EPC companies.

Those design engineers who are outside the EPC sector and do not have the support of a cost estimating department must rely on cost data from the open literature or use cost estimating software. The most widely used software tools for estimating chemical plant costs are the Aspen Capital Cost Estimator (ACCE) suite of tools licensed by Aspen Technology, Inc., and Cleopatra Enterprise owned by Cost Engineering Consultancy. The cost estimation software within ACCE is built on the ICARUS cost estimation algorithms. ACCE does not use the factorial method, but instead estimates equipment costs, bulk costs, and installation costs from the costs of materials and labor, following the practice used by cost engineers for detailed estimating. The models in ACCE are developed by a team of cost engineers based on data collected from EPC companies and equipment manufacturers. The models are updated annually. The ACCE software is included in the standard Aspen/Hysys academic package and is available in most universities. Cleopatra Enterprise is an integrated project cost management software that supports the entire project life cycle. It uses databases of equipment costs, bulk materials costs, and installation costs that were developed from vendor quotations and historic data and are updated annually. Cleopatra's database allows cost estimates to be built in different levels of detail, supporting the preparation of high-level, semi-detailed, and detailed estimates that cover the full scope of a project. The software also has a variety of features for analyzing and comparing different estimates and scenarios. Both ACCE and Cleopatra can give reasonably good estimates when used properly, and both are described in more detail in Section 7.10.

There is an abundance of equipment cost data and cost correlations in the open literature, but much of it is of very poor quality. The relationship between size and cost given in Equations 7.1 and 7.2 can also be used for equipment if a suitable size parameter is used. If the size range spans several orders of magnitude, then log-log plots usually give a better representation of the relationship than simple equations.

Some of the most reliable information on equipment costs can be found in the professional cost engineering literature. Correlations based on recent data are occasionally published in *Cost Engineering*, which is the journal of AACE International. AACE International also has an excellent website, [www.aacei.org](http://www.aacei.org), which has cost models that can be used by members. There is also an extensive listing of other web resources for cost estimating at <https://web.aacei.org/resources/publications>. The UK Association of Cost Engineers (ACostE) publishes the journal *The Cost Engineer* and also prints a guide to capital cost estimating (Gerrard, 2000), which gives cost curves for the main types of process equipment based on recent data. The prices are given in British pounds sterling on a UK basis, but this book is useful for making estimates of prices in northwest Europe. The International Cost Engineering Council website ([www.icoste.org](http://www.icoste.org)) provides links to 46 international cost engineering societies, several of which maintain databases of local costs.

Current prices for new and used equipment can be found on resale websites such as [www.equipnet.com](http://www.equipnet.com). It is not always easy to find the exact piece of equipment needed on such sites, but they can be used to give a good indication of the accuracy of cost correlations. A free web-based costing tool is available at [www.matche.com](http://www.matche.com). This tool appears to have not been updated since 2003, and the source of the cost correlations given is not clear, so it is not suitable for any use other than undergraduate design projects.

Many cost correlations can be found in chemical engineering textbooks, for example, Douglas (1988), Garrett (1989), Turton et al. (2003), Peters et al. (2003), Ulrich and Vasudevan (2004), and Seider et al. (2016). The references for such correlations should always be checked very carefully. When they are properly referenced, they are often found to be based on data published by Guthrie (1969, 1974) and updated using either cost indices (as described in Section 7.7) or a few recent data points. Guthrie's correlations were reasonably good when published, but there have been substantial changes in the relative contributions of material and fabrication costs of most process equipment since then. Academic authors usually do not have access to sufficient high-quality cost data to be able to make

reliable correlations, and most of the academic correlations predict lower costs than would be obtained using ACCE or other detailed estimating methods. These correlations are adequate for the purposes of university design projects but should not be used in real projects. It is to be hoped that the authors of these publications will benchmark their correlations against ACCE or Cleopatra in future editions, which will improve the accuracy of the correlations and make them more useful to those who do not have access to costing software.

### 7.5.2 Cost curves for purchased equipment costs

For those design engineers who lack access to reliable cost data or estimating software, the correlations given in Table 7.2 can be used for preliminary estimates. The correlations in Table 7.2 are of the form:

$$C_e = a + b S^n \quad (7.9)$$

where:  $C_e$  = purchased equipment cost on a U.S. Gulf Coast basis, Jan. 2010 (CEPCI = 532.9, NF refinery inflation index = 2281.6)

$a, b$  = cost constants in Table 7.2

$S$  = size parameter, units given in Table 7.2

$n$  = exponent for that type of equipment

The correlations in Table 7.2 are only valid between the lower and upper values of  $S$  indicated. The prices are all for carbon steel equipment except where noted in the table. Extrapolation to other materials is discussed in Section 7.6.3.

#### Example 7.2

Estimate the purchased equipment cost of a plain carbon steel shell and tube heat exchanger with area 400 m<sup>2</sup>.

#### Solution

From Table 7.2, the cost correlation for a shell and tube heat exchanger is:

$$C_e = 28,000 + 54 A^{1.2} \quad (7.9)$$

where  $A$  is the area in m<sup>2</sup> and  $C_e$  is on a January 2010 basis. Substituting for the area:

$$C_e = 28,000 + (54 \times 400^{1.2}) = \underline{\underline{\$99,600}}$$

### 7.5.3 Detailed method of cost estimating

When the method of design and construction of a piece of equipment are known, the cost can be estimated from the costs of the materials, parts, labor, and manufacturer's profit. This is the method preferred by professional cost estimators and procurement managers, as it allows the estimator to obtain an unbiased estimate of the real cost of the equipment, which can then be used in negotiations with the vendor to determine a fair price. This method is also used in many of the correlations in commercial cost estimating programs such as ACCE.

A detailed estimate requires an itemized list of the required parts, an understanding of the fabrication steps, knowledge of the machinery involved (so as to estimate the machine costs), and an understanding of the amount of labor needed for each step. The fabrication method is stated as a work breakdown structure (WBS) to arrive at an accurate estimate of the labor. Machine time costs are usually allocated by taking the annual cost of the machine, including allowances for capital recovery, maintenance, and electric power costs, and then dividing these costs by the number of hours in use to arrive at an hourly machine cost. The overall cost can then be made by summing the component costs, machine costs, and labor costs and adding suitable amounts for supervision, overhead, and manufacturer's profit. A hypothetical example of a cost breakdown structure for a shell and tube heat exchanger is given in Table 7.3.

The detailed method of cost estimating must be followed whenever cost or price data are not available, for example, when making an estimate of the cost of specialized equipment that cannot be found in the literature. For example, a reactor design is usually unique for a particular process but the design can be broken down into

TABLE 7.2 Purchased equipment cost for common plant equipment

Equipment	Units for size, <i>S</i>	<i>S<sub>lower</sub></i>	<i>S<sub>upper</sub></i>	<i>a</i>	<i>b</i>	<i>n</i>	Note
<i>Agitators and mixers</i>							
Propeller	driver power, kW	5.0	75	17,000	1130	1.05	
Spiral ribbon mixer	driver power, kW	5.0	35	30,800	125	2.0	
Static mixer	liters/s	1.0	50	570	1170	0.4	
<i>Boilers</i>							
Packaged, 15 to 40 bar	kg/h steam	5000	200,000	124,000	10.0	1.0	
Field erected, 10 to 70 bar	kg/h steam	20,000	800,000	130,000	53	0.9	
<i>Centrifuges</i>							
High-speed disk	diameter, m	0.26	0.49	57,000	480,000	0.7	
Atmospheric suspended basket	power, kW	2.0	20	65,000	750	1.5	
<i>Compressors</i>							
Blower	m <sup>3</sup> /h	200	5000	4450	57	0.8	
Centrifugal	driver power, kW	75	30,000	580,000	20,000	0.6	
Reciprocating	driver power, kW	93	16,800	260,000	2700	0.75	
<i>Conveyors</i>							
Belt, 0.5 m wide	length, m	10	500	41,000	730	1.0	
Belt, 1.0 m wide	length, m	10	500	46,000	1320	1.0	
Bucket elevator, 0.5 m bucket	height, m	10	30	17,000	2600	1.0	
<i>Crushers</i>							
Reversible hammer mill	t/h	30	400	68,400	730	1.0	
Pulverizers	kg/h	200	4000	16,000	670	0.5	
Jaw crusher	t/h	100	600	-8000	62,000	0.5	
Gyratory crusher	t/h	200	3000	5000	5100	0.7	
Ball mill	t/h	0.7	60	-23,000	242,000	0.4	
<i>Crystallizers</i>							
Scraped surface crystallizer	length, m	7	280	10,000	13,200	0.8	
<i>Distillation columns</i>							
See pressure vessels, packing, and trays							
<i>Dryers</i>							
Direct contact rotary	area, m <sup>2</sup>	11	180	15,000	10,500	0.9	1
Atmospheric tray batch	area, m <sup>2</sup>	3.0	20	10,000	7900	0.5	
Spray dryer	evap rate kg/h	400	4,000	410,000	2200	0.7	
<i>Evaporators</i>							
Vertical tube	area, m <sup>2</sup>	11	640	330	36,000	0.55	
Agitated Falling film	area, m <sup>2</sup>	0.5	12	88,000	65,500	0.75	2

Continued

TABLE 7.2 Purchased equipment cost for common plant equipment—cont'd

Equipment	Units for size, $S$	$S_{lower}$	$S_{upper}$	$a$	$b$	$n$	Note
<i>Exchangers</i>							
U-tube shell and tube	area, m <sup>2</sup>	10	1000	28,000	54	1.2	
Floating head shell and tube	area, m <sup>2</sup>	10	1000	32,000	70	1.2	
Double pipe	area, m <sup>2</sup>	1.0	80	1900	2500	1.0	
Thermosiphon reboiler	area, m <sup>2</sup>	10	500	30,400	122	1.1	
U-tube Kettle reboiler	area, m <sup>2</sup>	10	500	29,000	400	0.9	
Plate and frame	area, m <sup>2</sup>	1.0	500	1600	210	0.95	2
<i>Filters</i>							
Plate and frame	capacity, m <sup>3</sup>	0.4	1.4	128,000	89,000	0.5	
Vacuum drum	area, m <sup>2</sup>	10	180	-73,000	93,000	0.3	
<i>Furnaces</i>							
Cylindrical	duty, MW	0.2	60	80,000	109,000	0.8	
Box	duty, MW	30	120	43,000	111,000	0.8	
<i>Packings</i>							
304 ss Raschig rings	m <sup>3</sup>			0	8000	1.0	
Ceramic Intalox saddles	m <sup>3</sup>			0	2000	1.0	
304 ss pall rings	m <sup>3</sup>			0	8500	1.0	
PVC structured packing	m <sup>3</sup>			0	5500	1.0	
304 ss structured packing	m <sup>3</sup>			0	7600	1.0	3
<i>Pressure vessels</i>							
Vertical, cs	shell mass, kg	160	250,000	11,600	34	0.85	4
Horizontal, cs	shell mass, kg	160	50,000	10,200	31	0.85	4
Vertical, 304 ss	shell mass, kg	120	250,000	17,400	79	0.85	4
Horizontal, 304 ss	shell mass, kg	120	50,000	12,800	73	0.85	4
<i>Pumps and drivers</i>							
Single-stage centrifugal	flow liters/s	0.2	126	8000	240	0.9	
Explosion proof motor	power, kW	1.0	2500	-1100	2100	0.6	
Condensing steam turbine	power, kW	100	20,000	-14,000	1900	0.75	
<i>Reactors</i>							
Jacketed, agitated	volume, m <sup>3</sup>	0.5	100	61,500	32,500	0.8	2
Jacketed, agitated, glass lined	volume, m <sup>3</sup>	0.5	25	12,800	88,200	0.4	
<i>Tanks</i>							
Floating roof	capacity, m <sup>3</sup>	100	10,000	113,000	3250	0.65	
Cone roof	capacity, m <sup>3</sup>	10	4000	5800	1600	0.7	
<i>Trays</i>							
Sieve trays	diameter, m	0.5	5.0	130	440	1.8	5
Valve trays	diameter, m	0.5	5.0	210	400	1.9	
Bubble cap trays	diameter, m	0.5	5.0	340	640	1.9	



TABLE 7.2 Purchased equipment cost for common plant equipment—cont'd

Equipment	Units for size, $S$	$S_{lower}$	$S_{upper}$	$a$	$b$	$n$	Note
<i>Utilities</i>							
Cooling tower and pumps	flow liters/s	100	10,000	170,000	1500	0.9	6
Packaged mechanical refrigerator	evaporator duty, kW	50	1,500	24,000	3500	0.9	
Water ion exchange plant	flow m <sup>3</sup> /h	1	50	14,000	6200	0.75	

Notes:

1. Direct heated.

2. Type 304 stainless steel.

3. With surface area 350 m<sup>2</sup>/m<sup>3</sup>.

4. Not including heads, ports, brackets, internals, etc. (see Chapter 14 for how to calculate wall thickness).

5. Cost per tray, based on a stack of 30 trays.

6. Field assembly.

7. All costs are U.S. Gulf Coast basis, Jan. 2010 (CEPCI index = 532.9, NF refinery inflation index = 2281.6).

TABLE 7.3 Hypothetical cost breakdown for fabrication of a heat exchanger

Step	Materials	Parts	Machine	Labor
1. Shell fabrication				
1.1 Shell forming	Steel plate, $L_s \times \pi D_s \times t_s$		Cutting Rolling Welding	2 hr 2 hr 2 hr
1.2 End flange forming	Two steel plate discs, $D_f$ diameter $\times t_f$		Cutting Drilling (for bolt holes)	1 hr each 2 hr each
1.3 Flange attachment			Welding	2 hr each
1.4 Nozzle reinforcement	Two steel plate discs, $2 D_n$ diameter		Cutting/rolling Welding	2 hr 1 hr each
1.5 Nozzle attachment		Two flanged pipes	Welding	1 hr each
2. Head fabrication ( $\times 2$ ) (Note: Both nozzles may be on one head, depending on the exchanger type.)				
2.1 Head forming	Steel plate disc		Cutting Drop forging	1 hr
2.2 Hole cutting			Cutting	4 hr
2.3 End flange forming	Two steel plate discs, $D_f$ diameter $\times t_f$		Cutting Drilling (for bolt holes)	1 hr 2 hr
2.4 Flange attachment			Welding	2 hr
2.5 Nozzle reinforcement	Two steel plate discs, $D_n$ diameter		Cutting/rolling Welding	1 hr each 1 hr each
2.6 Nozzle attachment		Two flanged pipes	Welding	1 hr each
3. Tube bundle fabrication				
3.1 Tube sheet fabrication	Steel plate disc, $D_s$ diameter		Stamping	1 hr
3.2 Baffle cutting	$N_{baf}$ steel plate discs		Stamping	0.5 hr each
3.3 Tie rod cutting	10 to 12 $\times L$ , steel rod		Cutting Threading	1 hr 2 hr
3.4 Tube preparation	$N_t$ tubes cut to right length		Cutting Bending	0.25 hr each 0.25 hr each

Continued

TABLE 7.3 Hypothetical cost breakdown for fabrication of a heat exchanger—cont'd

Step	Materials	Parts	Machine	Labor
3.5 Bundle assembly				0.25 hr/tube
3.6 Tube sheet sealing			Welding/rolling	0.25 hr/tube
4. Exchanger assembly				
4.1 Bundle insertion			Large crane	1hr
4.2 Gasket cutting	Two sheets gasket material		Stamping	1 hr
4.3 Head attachment		$N_b$ bolts		2 hr

Notes:

1.  $L_s$  = shell length,  $D_s$  = shell diameter,  $t_s$  = shell thickness,  $D_f$  = flange diameter,  $t_f$  = flange thickness,  $D_n$  = nozzle diameter,  $N_{\text{baf}}$  = number of baffles,  $N_t$  = number of tubes,  $N_b$  = number of bolts.

2. Dimensions are not exact, labor hours are approximate, and some steps have been omitted to simplify the example. There could be considerable variation in labor times depending on the exchanger layout and complexity.

3. Costs would also include supervision, testing and inspection, overhead, and manufacturer's margin.

4. The components and fabrication of a shell and tube heat exchanger are discussed in detail in [Chapter 19](#).

standard components (vessels, heat-exchange surfaces, spargers, agitators, etc.), the cost of which can be found in the literature and used to build up an estimate of the reactor cost.

More information on estimating costs from a detailed breakdown of parts and labor is given by [Dysert \(2015\)](#) in the AACE International's training manual ([Hastak, 2015](#)). Breakdowns of the materials and labor components for many types of process equipment are given by [Page \(1996\)](#). [Pikulik and Diaz \(1977\)](#) give a method of costing major equipment items from cost data on the basic components: shells, heads, nozzles, and internal fittings. [Purohit \(1983\)](#) gives a detailed procedure for estimating the cost of heat exchangers.

#### 7.5.4 Use of vendor data in cost estimating

At some point in the development of a design it is always necessary to get a real price quotation from an equipment vendor. Although vendor quotations are always realistic, some caution is needed to ensure that such estimates are valid for the use intended.

A large amount of vendor information is now available online and can easily be found using any of the major search engines or by starting from directories such as [www.purchasing.com](http://www.purchasing.com). Online costs are usually manufacturer's catalog prices for small-order quantities. Large order sizes (as filled by contractors) are often heavily discounted, which is one reason why many operators subcontract plant construction to EPC companies that have greater purchasing power. Items requiring special fabrication, for example, large vessels or compressors, may experience discounts or surcharges depending on the state of the manufacturer's order books and the purchasing power of the customer.

When a vendor is contacted directly, the quality of estimate that they can provide depends very much on the quality of the information that they are given. Often the vendor will ask the customer to provide information on the process conditions and equipment duty, so as to be able to complete their own optimization and ensure that the model or design specified is correct for the purpose. It is very important to include any constraints on materials of construction in the specification provided to the vendor, as materials substitution can cause significant increases in cost; see Section 7.6.3.

As with any purchasing decision, it is always a good idea to contact several vendors and comparison shop for the best price. Care must be taken to be ethical in handling vendors' confidential information, and commercially sensitive information such as pricing should not be shared with other vendors.

### 7.6 Estimating installed costs: The factorial method

Capital cost estimates for chemical process plants are often based on an estimate of the purchase cost of the major equipment items required for the process, with the other costs being estimated as factors of the equipment cost. The accuracy of this type of estimate will depend on what stage the design has reached at the time the estimate is made

TABLE 7.4 Installation factors proposed by Hand (1958)

Equipment type	Installation factor
Compressors	2.5
Distillation columns	4
Fired heaters	2
Heat exchangers	3.5
Instruments	4
Miscellaneous equipment	2.5
Pressure vessels	4
Pumps	4

and on the reliability of the data available on equipment costs. In the later stages of the project design, when detailed equipment specifications are available and firm quotes have been obtained from vendors, a Class 3 estimate of the capital cost of the project can be made by this method, but generally this method leads to a Class 4 estimate.

### 7.6.1 Lang factors

Lang (1948) proposed that the ISBL fixed capital cost of a plant is given as a function of the total purchased equipment cost by the equation:

$$C = F \left( \sum C_e \right) \quad (7.10)$$

where:  $C$  = total plant ISBL capital cost (including engineering costs).

$\sum C_e$  = total delivered cost of all the major equipment items: reactors, tanks, columns, heat exchangers, furnaces, etc. The delivered cost is the purchased cost plus shipping or delivery costs.

$F$  = an installation factor, later widely known as a *Lang factor*.

Lang originally proposed the following values of  $F$ , based on 1940s economics:

$F = 3.1$  for solids processing plants

$F = 4.74$  for fluids processing plants

$F = 3.63$  for mixed fluids-solids processing plants

Hand (1958) suggested that better results are obtained by using different factors for different types of equipment. Examples of the factors proposed by Hand are given in Table 7.4. Hand also observed that this approach should only be used in the earliest stages of process design and in the absence of detailed design information.

Both Lang (1948) and Hand (1958) included home office costs but not offsite costs or contingency in their installation factors, so beware of double counting EPC costs when using this approach. The relative costs of materials and labor have changed substantially from when these factors were developed, and the accuracy of the correlation probably never warranted three significant figures for  $F$ . Most practitioners using this method therefore use a Lang factor of 3, 4, or 5, depending on the plant scale (larger plant = smaller factor) and type.

### 7.6.2 Detailed factorial estimates

Equation 7.10 can be used to make a preliminary estimate once the flowsheet has been drawn up and the main plant equipment has been sized. When more detailed design information is available, the installation factor can be estimated somewhat more rigorously by considering the cost factors that are compounded into the Lang factor individually.

The direct-cost items that are incurred in the construction of a plant in addition to the delivered cost of equipment are:

1. Equipment erection, including foundations and minor structural work
2. Piping, including insulation and painting
3. Electrical, power, and lighting
4. Instruments and automatic process control (APC) systems
5. Process buildings and structures
6. Ancillary buildings, offices, laboratory buildings, and workshops (if not costed separately as offsites)
7. Storage for raw materials and finished product (if not costed separately as offsites)
8. Utilities and provision of plant for steam, water, air, and firefighting services (if not costed separately as offsites)
9. Site preparation

The contribution of each of these items to the total capital cost is calculated by multiplying the total purchased equipment cost by an appropriate factor. As with the basic Lang factor, these factors are best derived from historical cost data for similar processes. Typical values for the factors are given in several references: [Happle and Jordan \(1975\)](#) and [Garrett \(1989\)](#). [Guthrie \(1974\)](#) splits the costs into the material and labor portions and gives separate factors for each.

The accuracy and reliability of an estimate can be further improved by dividing the process into subunits and using factors that depend on the function of the subunits; see [Guthrie \(1969\)](#). In Guthrie's detailed method of cost estimation, the installation, piping, and instrumentation costs for each piece of equipment are costed separately. Detailed costing is only justified if the cost data available are reliable and the design has been taken to the point where all the cost items can be identified and included. [Gerrard \(2000\)](#) gives factors for individual pieces of equipment as a function of equipment cost and complexity of installation.

Typical factors for the components of the capital cost are given in [Table 7.5](#). These can be used to make an approximate estimate of capital cost using equipment cost data published in the literature.

### 7.6.3 Materials factors

The installation factors given in [Tables 7.4 and 7.5](#) are for plants built from carbon steel. When more exotic materials are used, a materials factor  $f_m$  should also be introduced:

TABLE 7.5 Typical factors for estimation of project fixed capital cost

Item	Process type		
	Fluids	Fluids–solids	Solids
1. Major equipment, total purchase cost	$C_e$	$C_e$	$C_e$
$f_{er}$ Equipment erection	0.3	0.5	0.6
$f_p$ Piping	0.8	0.6	0.2
$f_i$ Instrumentation and control	0.3	0.3	0.2
$f_{el}$ Electrical	0.2	0.2	0.15
$f_c$ Civil	0.3	0.3	0.2
$f_s$ Structures and buildings	0.2	0.2	0.1
$f_l$ Lagging and paint	0.1	0.1	0.05
ISBL cost, $C = \Sigma C_e \times$	3.3	3.2	2.5
Offsites (OS)	0.3	0.4	0.4
Design and Engineering (D&E)	0.3	0.25	0.2
Contingency (X)	0.1	0.1	0.1
Total fixed capital cost $C_{FC} = C(1 + OS)(1 + D\&E + X)$			
$= C \times$	1.82	1.89	1.82
$= \Sigma C_e \times$	6.00	6.05	4.55

$$f_m = \frac{\text{purchased cost of item in exotic material}}{\text{purchased cost of item in carbon steel}} \quad (7.11)$$

Note that  $f_m$  is not equal to the ratio of the metal prices, as the equipment purchased cost also includes labor costs, overheads, fabricator's profit, and other costs that do not scale directly with metal price. Equation 7.10 can then be expanded for each piece of equipment to give:

$$C = \sum_{i=1}^{i=M} C_{e,i,CS} [(1+f_p)f_m + (f_{er}+f_{el}+f_i+f_c+f_s+f_l)] \quad (7.12)$$

Equation 7.12 should be used when the purchased equipment cost has been found on a carbon steel basis and the designer is estimating the cost for alloy construction. If the purchased equipment cost has been obtained on an alloy basis, then the designer should instead correct the other installation factors so as not to overestimate the cost of installation:

$$C = \sum_{i=1}^{i=M} C_{e,i,A} [(1+f_p) + (f_{er}+f_{el}+f_i+f_c+f_s+f_l) / f_m] \quad (7.13)$$

where:  $C_{e,i,CS}$  = purchased equipment cost of equipment  $i$  in carbon steel

$C_{e,i,A}$  = purchased equipment cost of equipment  $i$  in alloy

$M$  = total number of pieces of equipment

$f_p$  = installation factor for piping

$f_{er}$  = installation factor for equipment erection

$f_{el}$  = installation factor for electrical work

$f_i$  = installation factor for instrumentation and process control

$f_c$  = installation factor for civil engineering work

$f_s$  = installation factor for structures and buildings

$f_l$  = installation factor for lagging, insulation, or paint

Failure to properly correct installation factors for materials of construction is one of the most common sources of error with the factorial method. Typical values of the materials factor for common engineering alloys are given in [Table 7.6](#). A more detailed discussion of the relative cost of materials and sources of material cost factors is given in Section 6.6.

#### 7.6.4 Summary of the factorial method

Many variations of the factorial method are used. The method outlined here can be used with the data given in this chapter to make a quick, approximate estimate of the fixed capital investment needed for a project.

TABLE 7.6 Materials cost factors,  $f_m$ , relative to plain carbon steel

Material	$f_m$
Carbon steel	1.0
Aluminum and bronze	1.07
Cast steel	1.1
304 stainless steel	1.3
316 stainless steel	1.3
321 stainless steel	1.5
Hastelloy C	1.55
Monel	1.65
Nickel and Inconel	1.7

1. Prepare material and energy balances; draw up preliminary flowsheets; size major equipment items and select materials of construction.
2. Estimate the purchased cost of the major equipment items; see Section 7.5.
3. Calculate the ISBL installed capital cost, using the factors given in Table 7.5 and correcting for materials of construction using Equation 7.12 or 7.13 with the materials factors given in Table 7.6.
4. Calculate the OSBL, engineering, and contingency costs using the factors given in Table 7.5.
5. The sum of ISBL, OSBL, engineering, and contingency costs is the fixed capital investment.
6. Estimate the working capital as a percentage of the fixed capital investment; 10% to 20% is typical (or, better, calculate it from the cost of production if this has been estimated—see Section 9.2.3).
7. Add the fixed and working capital to get the total investment required.

### Example 7.3

A plant modification has been proposed that will allow recovery of a by-product. The modification consists of adding the following equipment:

Distillation column, height 30 m, diameter 3 m, 50 sieve trays, operating pressure 10 bar

- U-tube heat exchanger, area 60 m<sup>2</sup>
- Kettle reboiler, area 110 m<sup>2</sup>
- Horizontal pressure vessel, volume 3 m<sup>3</sup>, operating pressure 10 bar
- Storage tank, volume 50 m<sup>3</sup>
- Two centrifugal pumps, flow rate 3.6 m<sup>3</sup>/h, driver power 500 W
- Three centrifugal pumps, flow rate 2.5 m<sup>3</sup>/h, driver power 1 kW (two installed plus one spare)

Estimate the installed ISBL capital cost of the modification if the plant is to be built from type 304 stainless steel. Estimate the cost using both Hand's method and the factors given in Table 7.5.

### Solution

The first step is to convert the units to those required for the correlations and determine any missing design information. The distillation column can be costed as a combination of a vertical pressure vessel and internals. For both pressure vessels we need to know the wall thickness. The details of how to calculate vessel wall thickness in accordance with the ASME Boiler and Pressure Vessel Code are given in Section 14.5, and the equation to use is Equation 14.13.

The design pressure of the vessels should be 10% above the operating pressure (see Chapter 14), so the design pressure is 11 bar, or roughly  $1.1 \times 10^6$  N/m<sup>2</sup>. The maximum allowable stress for type 304 stainless steel at 500 °F (260 °C) is 12.9 ksi or roughly 89 N/mm<sup>2</sup> (see Table 14.2). Assuming the welds will be fully radiographed, the weld efficiency is 1.0. Substituting in Equation 14.13 for the column wall thickness,  $t_w$  then gives:

$$t_w = \frac{1.1 \times 10^6 \times 3}{(2 \times 89 \times 10^6 \times 1.0) - (1.2 \times 1.1 \times 10^6)} \quad (14.13)$$

$$= 0.0187 \text{ m, say 20 mm}$$

We can now calculate the shell mass, using the density of 304 stainless steel (= 8000 kg/m<sup>3</sup>, from Table 6.2).

$$\text{Shell mass} = \pi D_c L_c t_w \rho$$

where:  $D_c$  = vessel diameter, m

$L_c$  = vessel length, m

$t_w$  = wall thickness, m

$\rho$  = metal density, kg/m<sup>3</sup>

So the shell mass for the distillation column is:

$$\text{Shell mass} = \pi \times 3.0 \times 30 \times 0.02 \times 8000 = 46685 \text{ kg}$$

For the horizontal pressure vessel we need to convert the volume into a length and diameter. Assuming that the vessel is a cylinder with  $L_c = 2D_c$  then we can follow the same method as for the column and find  $t_w = 8$  mm and shell mass = 636 kg.

Using the correlations in Table 7.2, we obtain the following purchased costs for the stainless steel pressure vessels:

$$\text{Distillation column shell, cost} = 17,400 + 79 (46685)^{0.85} = \$753,000$$

$$\text{Horizontal pressure vessel, cost} = 12,800 + 73 (636)^{0.85} = \$30,400$$

For the remaining equipment we obtain the following purchased costs from the correlations in Table 7.2 based on carbon steel construction:

Distillation column trays, cost per tray = $130 + 440(3.0)^{1.8}$	= \$3310
Cost for 50 trays	= \$165,500
U – tube heat exchanger, cost = $28,000 + 54(60)^{1.2}$	= \$35,300
Kettle reboiler, cost = $29,000 + 400(110)^{0.9}$	= \$56,500
Storage tank (conical head), cost = $5800 + 1600(50)^{0.7}$	= \$30,500
Centrifugal pump, $3.6 \text{ m}^3/\text{h} = 1\text{L/s}$ , SO:	
cost each = $8000 + 240(1.0)^{0.9} = \$8240$ , cost for two pumps	= \$16,480
driver (electric motor) cost each = $-1100 + 2,100(0.5)^{0.6}$	= \$285
cost for two drivers	= \$570
Centrifugal pump, $2.5 \text{ m}^3/\text{h} = 0.694 \text{ L/s}$ , so	
cost each = $8000 + 240(0.694)^{0.9} = \$8170$ , cost for three	= \$24,520
driver (electric motor) cost each = $-1100 + 2100(1.0)^{0.6}$	= \$1000
cost for three drivers	= \$3000

Note that the pumps and drivers are at the lower end of the range of validity of the cost correlations and the driver cost for the smaller motors looks suspect; however, the pump and motor costs are small compared with the other costs, and the error introduced is therefore negligible given the overall accuracy of  $\pm 30\%$ .

Following Hand's method, the installed cost of the distillation column is then:

$$C = 4 \times 753,000 = \$3,012,000$$

The cost of the trays can be converted to type 304 stainless steel by multiplying by the appropriate materials factor from Table 7.6, giving:

$$C = 1.3 \times 165,500 = \$215,150$$

This then gives a total cost for the column plus internals of  $3,012,000 + 215,150 = \$3,227,150$ .

The installed cost of the horizontal pressure vessel is  $4 \times 30,400 = \$121,600$ .

The installed cost for the exchangers and storage tank in carbon steel construction is:

$$C = 3.5(35,300 + 56,500) + 2.5(30,500) = \$397,550$$

so the cost in type 304 stainless steel is  $1.3 \times 397,550 = \$516,800$ .

For the pumps, we need to add the cost of the pump and driver before determining the installed cost. Only the cost of the pump needs to be converted to stainless steel. For the first set of pumps:

$$C = 4 \times (570 + (1.3 \times 16,480)) = \$88,000$$

For the second set of pumps only two are installed (the other is a warehouse spare), so the total installed cost is:

$$C = (1.3 \times 8170) + 1000 + (4 \times 2 \times (1000 + (1.3 \times 8170))) = \$105,000$$

The total installed ISBL cost of the plant is then:

$$C = 3,227,150 + 121,600 + 516,800 + 88,000 + 105,000 = \$4,058,550$$

or \$4.1 MM  $\pm 30\%$  within the accuracy of the method.

If instead we use the factors given in Table 7.5, then using Equation 7.12, the installed cost for the exchangers, tank, and pumps is equal to:

$$C = (35,300 + 56,500 + 30,500 + 16,480 + 8170)[(1 + 0.8) \times 1.3 + (0.3 + 0.3 + 0.2 + 0.3 + 0.2 + 0.1)]$$

$$C = (155,120)[3.74] = \$580,150$$

The installed cost for the pump drivers (which do not require a materials conversion factor) is:

$$C = (1140 + 6,000)[1 + 0.8 + 0.3 + 0.3 + 0.2 + 0.3 + 0.2 + 0.1]$$

$$C = (7140)[3.2] = \$22,900$$

The installed cost for the pressure vessels can be found using Equation 7.13:

$$C = (753,000 + 30,400)[1 + 0.8 + (0.3 + 0.3 + 0.2 + 0.3 + 0.2 + 0.1) / 1.3]$$

$$C = (783,400)[2.88] = \$2,256,200$$

In addition to this, we require the cost of the trays in stainless steel and the cost of the spare pump and driver:

$$C = 1000 + 1.3(165,500 + 8170) = \$226,800$$

The total installed ISBL cost of the plant is then:

$$C = 580,150 + 22,900 + 2,256,200 + 226,800 = \$3,086,050$$

or \$3.1 MM ± 30% within the accuracy of the method.

Note that although the answers obtained by the two methods are different, the second answer is within the range of accuracy of the first, and the first is very close to the upper end of the range predicted by the second method. Both estimates should be stated as being on a U.S. Gulf Coast basis, January 2010, as this is the basis for the correlations in Table 7.2.

## 7.7 Cost escalation

All cost estimating methods use historical data and are themselves forecasts of future costs. The prices of the materials of construction and the costs of labor are subject to inflation. Some method has to be used to update old cost data for use in estimating at the design stage and to forecast the future construction cost of the plant.

The method usually used to update historical cost data makes use of published cost indices. These relate present costs to past costs and are based on data for labor, material, and energy costs published in government statistical digests.

$$\text{Cost in year A} = \text{Cost in year B} \times \frac{\text{Cost index in year A}}{\text{Cost index in year B}} \quad (7.14)$$

To get the best estimate, each job should be broken down into its components, and separate indices should be used for labor and materials. It is often more convenient to use the composite indices published for various industries in the trade journals. These are weighted average indices combining the various components of costs in proportions considered typical for the particular industry.

A composite index for the U.S. process plant industry is published monthly in the journal *Chemical Engineering*; this is the Chemical Engineering Plant Cost Index (CEPCI), often referred to as the *CE index*. *Chemical Engineering* also used to publish the Marshall and Swift index (M&S equipment cost index), but that index has been discontinued since 2010.

For oil refinery and petrochemicals projects, the *Oil and Gas Journal* used to publish the Nelson-Farrar Refinery Construction Index (NF index). The index is now available as a subscription service from <https://www.bakerrisk.com/products/nelson-farrar-cost-index/>. The index is updated monthly, and indices for 40 types of equipment are updated quarterly. The Nelson-Farrar index is on a U.S. Gulf Coast basis rather than U.S. average, and is more reliable than the CE index for the types of equipment used in hydrocarbon processing.

The journal *Engineering News Record* publishes a monthly construction cost index. This is based on civil engineering projects and is sometimes used for updating offsite costs. This index has been published since 1904 and is the oldest of all the indices.

For international projects, the journal *Process Engineering* publishes monthly cost indices for several countries, including the United States, United Kingdom, Japan, Australia, and many of the EU countries.

All cost indices should be used with caution and judgment. They do not necessarily relate the true make-up of costs for any particular piece of equipment or plant, nor the effect of supply and demand on prices. The longer



the period over which the correlation is made, the more unreliable the estimate. Between 1970 and 1990 prices rose dramatically. Prices then grew at a more or less steady 2% to 3% per year until 2003, when high demand for fuels projects and high energy prices caused another period of steeper price inflation. Prices then decreased during the 2008–2010 recession before another period of growth driven by major projects associated with the U.S. shale gas expansion and downstream petrochemical investments. At time of writing, the indices have not yet shown any impact of the 2020 COVID-19 pandemic and resulting economic recession and oil price crash; however, it is likely that there will be a substantial decrease in the indices in 2020 and that recovery to 2019 levels will not occur until there is renewed capital investment in the fuels and chemicals sector. The major cost indices are plotted in Fig. 7.2. Fig. 7.3 shows the same data plotted relative to the 1990 value of each index. Fig. 7.3 shows that the NF index is usually a leading indicator compared with the M&S and CE indices, as fuels-sector activity has been a big component of price inflation from 2000 onwards.

To estimate the future cost of a plant, some prediction has to be made of the future annual rate of cost inflation. This can be based on the extrapolation of one of the published indices, tempered by the engineer’s own assessment of what the future may hold. Inflation is difficult to forecast, and allowance for inflation is often included in the contingency charges added to the project cost.

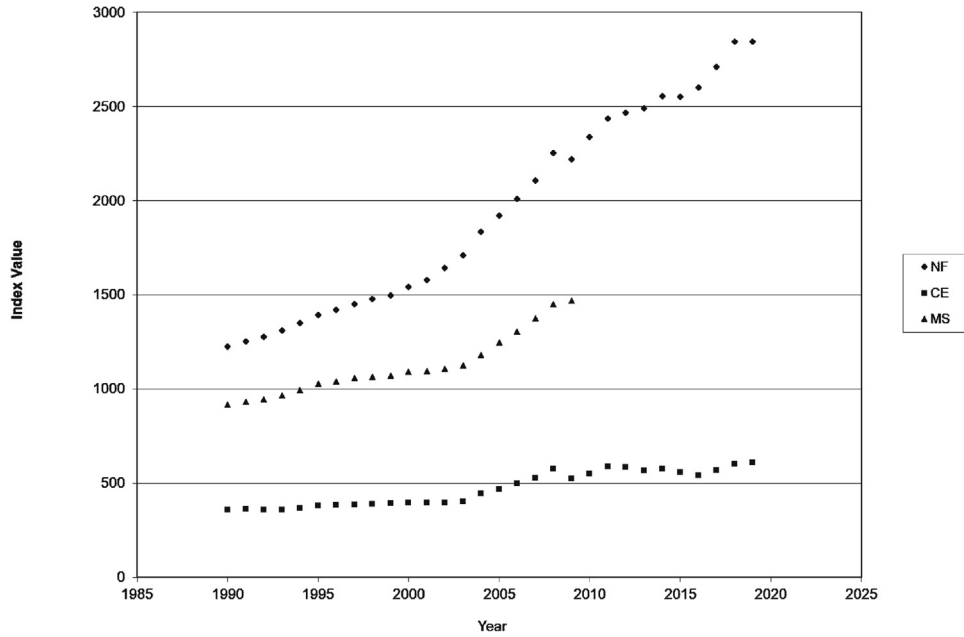


FIG. 7.2 Variation of major cost indices.

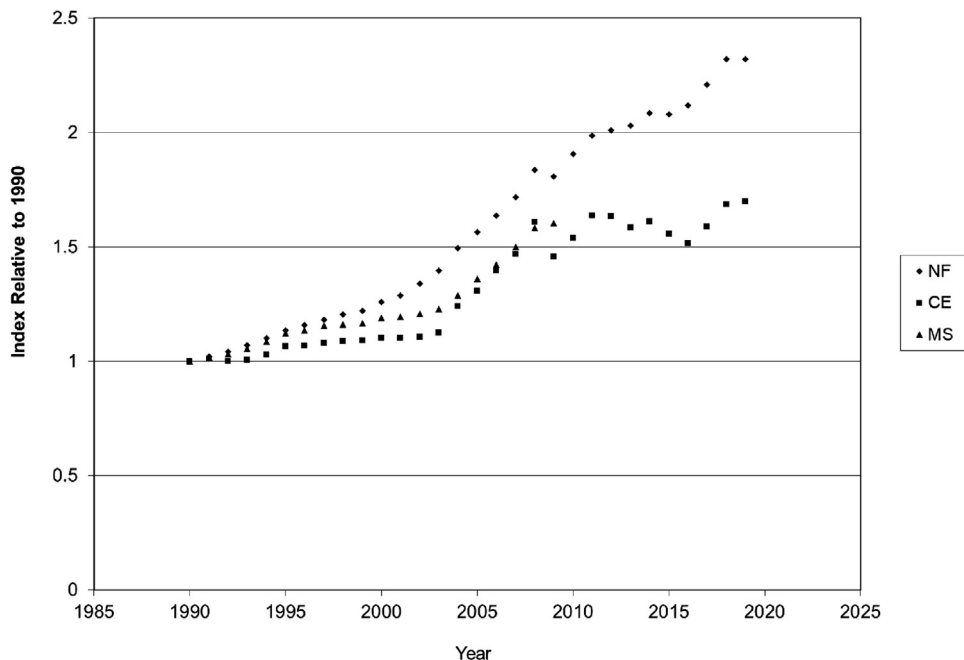


FIG. 7.3 Variation of major cost indices relative to 1990 = 1.0.

**Example 7.4**

The purchased cost of a shell and tube heat exchanger, carbon shell, 316 stainless steel tubes, heat transfer area 500 m<sup>2</sup>, was \$64,000 in January 2003; estimate the cost in January 2020. Use the CEPCI Equipment Cost Index.

**Solution**

From Fig. 7.2 (or by looking up the index in *Chemical Engineering*):

Index in 2003 = 402

Index in January 2020 = 596

So, estimated cost in January 2020 =  $\$64,000 \times 596 / 402 = \underline{\underline{\$95,000}}$

**Example 7.5**

The purchased cost of a distillation column was \$136,000 in 2004. Estimate the cost in 2024. Use the Nelson-Farrer index.

**Solution**

From Fig. 7.2 (or by looking up the index in *Oil and Gas Journal*):

Index in 2004 = 1833.6

The index in 2024 is difficult to extrapolate numerically, as it is not yet clear how deep or long the 2020 recession will be, nor whether we will see another sharp increase afterwards, as experienced in 2015–2018. Looking at Fig. 7.2, it seems likely that the NF index will have a value in the range of 3000 to 3200. A conservative approach would be to make a higher estimate, so let us assume 3200.

So, estimated cost in January 2022 =  $\$136,000 \times 3200 / 1833.6 = \underline{\underline{\$237,000}}$

## 7.8 Location factors

Most plant and equipment cost data are given on a U.S. Gulf Coast (USGC) or northwest Europe (NWE) basis, as these were historically the main centers of the chemical industry, for which the most data were available. The cost of building a plant in any other location will depend on:

- Local fabrication and construction infrastructure
- Local labor availability and cost
- Costs of shipping or transporting equipment to site
- Import duties or other local tariffs
- Currency exchange rates, which affect the relative cost of locally purchased items such as bulk materials, when converted to a conventional pricing basis such as U.S. dollars

These differences are captured in cost estimating by using a location factor:

$$\text{Cost of plant in location A} = \text{cost of plant on USGC} \times LF_A \quad (7.15)$$

where:  $LF_A$  = location factor for location A relative to USGC basis

Location factors for international locations are a strong function of currency exchange rates and hence fluctuate with time. Cran (1976a,b), Bridgewater (1979), Soloman (1990), and Gerrard (2000) give location factors for international locations from which this variation can be seen. It can be argued that as a result of globalization, all international installation factors are trending closer to 1.0 (Gerrard, 2000). Location factors within a country are somewhat

TABLE 7.7 Location factors

Country	Region	Location factor
United States	Gulf Coast	1.00
	East Coast	1.04
	West Coast	1.07
	Midwest	1.02
Canada	Ontario	1.00
	Fort McMurray	1.60
Mexico		1.03
Brazil		1.14
China	imported	1.12
	indigenous	0.61
Japan		1.26
Southeast Asia		1.12
Australia		1.21
India		1.02
Middle East		1.07
France		1.13
Germany		1.11
Italy		1.14
Netherlands		1.19
Russia		1.53
United Kingdom		1.02

easier to predict, and Bridgewater (1979) suggested a simple rule of thumb: add 10% for every 1000 miles from the nearest major industrial center.

Table 7.7 gives example location factors. These are based on data from Aspen Richardson's *International Construction Cost Factor Location Manual* (2003). More recent versions of this manual can be found by searching for Richardson Engineering Services at [www.aspentech.com](http://www.aspentech.com). The values in Table 7.7 give costs on a local basis in U.S. dollars. The location factors in Table 7.7 are based on 2003 data and can be updated by dividing by the ratio U.S. dollar/local currency in 2003 and multiplying by the ratio U.S. dollar/local currency in the year of interest. If a cost estimate for a future year is being made, then currency variation will have to be forecasted.

### Example 7.6

The cost of constructing a 30,000 metric tons per year (30 kMTA) acrolein plant was estimated as \$80 million (\$80MM) on a 2006 U.S. Gulf Coast basis. What would be the cost in U.S. dollars on a 2006 Germany basis?

### Solution

From Table 7.7, the 2003 location factor for Germany was 1.11.

The exchange rate in 2003 averaged about €1 = \$1.15, and in 2006 it averaged about €1 = \$1.35.

The 2006 location factor for Germany is thus  $1.11 \times 1.35/1.15 = 1.30$

The cost of building the acrolein plant in Germany in 2006 is  $\$80 \text{ MM} \times 1.30 = \underline{\underline{\$104 \text{ MM}}}$

## 7.9 Estimating off-site capital costs

Improvements to the site infrastructure are almost always needed when a new plant is added to a site or a major expansion is carried out. The cost of such improvements is known as the *off-site or OSBL investment*, as described in Section 7.2.1.

In the early stages of designing a new process, the off-site requirements are usually not precisely known, and an allowance for off-site costs can be made by assuming that they will be a ratio of the ISBL investment. A typical number is 30% to 50% of ISBL investment, depending on the process and site conditions. Table 7.8 gives some guidelines for making approximate estimates of off-site capital costs as a function of plant complexity and site conditions.

TABLE 7.8 Guidelines for estimating approximate OSBL costs as a percentage of ISBL cost

Process complexity	Site condition		
	Existing: underused	Existing: tight capacity	New site
Typical large-volume chemical	30%	40%	40%
Low-volume specialty chemical	20%	40%	50%
High solids-handling requirement	40%	50%	100%

As the design details are established and the requirements for utilities such as steam, electricity, and cooling water are determined, the site requirements can also be determined. If there is insufficient spare capacity in the existing site infrastructure, potential modifications can be designed to accommodate the new plant. For existing sites, the design engineer must always beware of scope creep, in which the new project is used to justify all kinds of (possibly long overdue) improvements to the site infrastructure. If scope creep occurs, excessive off-site costs may make the project economics unattractive.

Many of the off-site items are designed as “packaged” plants or systems that are purchased from specialized suppliers. In some cases, the supplier may even offer an *over-the fence* contract, in which the supplier builds, owns, and operates the off-site plant and contracts to supply the site with the desired utility stream or service. Over-the-fence contracts are widely used for industrial gases such as nitrogen, oxygen, and hydrogen, and most plants also import electricity from the local utility company. Over-the-fence contracts for steam, cooling water, and effluent treatment are less common, but are sometimes used in smaller plants or where several companies share a site.

The question of whether to build a self-contained infrastructure for a plant or contract for off-site services is an example of a *make-or-buy* problem. The over-the-fence price will usually be higher than the cost of producing the utility or service internally, because the supplier needs to make a profit and recover their capital investment. On the other hand, contracting for the service reduces the project capital investment and fixed costs, because the supplier must take on the costs of labor, maintenance, and overheads. The make-or-buy decision is usually made by comparing annualized costs, as described in Section 9.7.2. Correlations for costs of utility plants and other off-sites are given in the sources listed in Section 7.5.

## 7.10 Computer tools for cost estimating

It is difficult for engineers outside the EPC sector to collect recent cost data from a large set of real projects and maintain accurate and up-to-date cost correlations. Instead, the most common method for making preliminary estimates in industry is to use commercial cost estimating software.

A wide variety of cost estimating programs is available. These include ACCE (Aspen Technology, Inc.), Cleopatra Enterprise (Cost Engineering Consultancy), CostLink/CM (Building Systems Design, Inc.), Cost Track (OnTrack Engineering Ltd.), PRISM Project Estimator (ARES Corp.), Success Estimator (U.S. Cost), Visual Estimator (CPR International Inc.), WinEst (Win Estimator), and others that can be found by searching on the Web or looking at the listings provided by AACE International at [www.aacei.org](http://www.aacei.org). The discussion in this section will focus on the most widely used programs: Aspen Technology’s ACCE software and Cost Engineering Consultancy’s Cleopatra Enterprise. Both of these programs have academic licenses available for university use and are also available in most chemical companies.

The ACCE and Cleopatra cost estimating tools are simple to use and give quick, defensible estimates without requiring a lot of design data. Design information can be uploaded from major flowsheet simulation programs or entered manually. The programs allow the design to be updated as more information on design details becomes available, so that a more accurate estimate can be developed. Costs can be estimated for a whole plant or for one piece of equipment at a time. Both programs include a large number of equipment types, and these can be designed in a broad range of materials, including U.S., UK, German, and Japanese standard alloys.

Both ACCE and Cleopatra use a combination of mathematical models and expert systems to develop cost estimates. Costs are based on the materials and labor required (following the practice used for detailed estimates) rather than installation factors. If design parameters are not specified by the user, then they are calculated or set to default values by the program. The user should always review the design details carefully to make sure that the default values make sense for the application. If any values are not acceptable, they can be manually adjusted and a more realistic estimate can be generated.

A detailed description of how to run ACCE or Cleopatra is beyond the scope of this book. Both programs have user manuals available from the licensor or as .pdf files downloadable from the “help” button in the program, see, for example, [AspenTech \(2002a,b\)](#). Some of the common issues that arise in using these programs are discussed next. These or similar problems are also faced when using other cost estimating software.

### 7.10.1 Mapping simulation data

Instructions on loading data from a process simulation are given in the ACCE User’s Guide ([AspenTech, 2002a](#)). When a simulator report file is loaded, ACCE generates a block-flow diagram with each unit operation of the simulation shown as a block. These blocks must then be “mapped” to ICARUS project components (pieces of equipment or bulk items).

Unless the user specifies otherwise, each simulator block is mapped to a default ICARUS project component. The mapping defaults need to be understood properly, as large errors can be introduced if unit operations are mapped incorrectly. The default mapping specifications are given in section 3 of the user’s guide ([AspenTech, 2002a](#)). Some mappings that commonly cause problems include:

1. Reactors: Plug-flow reactor models (PLUG in Hysys and ProII, RPLUG in AspenPlus) are mapped to a packed tower, which is fine for fixed-bed catalytic reactors, but not for other types of plug-flow reactors. All other reactor models (Gibbs, stoichiometric, equilibrium, and yield) are mapped to agitated tank reactors. Reactors that are not suitable for these mappings can be mapped to other ICARUS project components or set up as user models (see later).
2. Heaters, coolers, and heat exchangers: The default mapping for all heat transfer equipment is the floating head heat exchanger. ICARUS contains several different heat exchanger types, including a generic TEMA heat exchanger that can be customized to the other types, as well as fired heater and air cooler components. It is often worthwhile to change the default mapping to the TEMA exchanger to allow the exchangers to be customized in ICARUS.
3. Distillation columns: The simulator column models include not just the column itself but also the reboiler, condenser, overhead receiver drum, and reflux pump (but not bottoms pump). ICARUS has 10 possible configurations to which a column can be mapped. Alternatively, the column can be mapped to a packed or trayed tower, and the ancillary items can be created as separate ICARUS project components.
4. Dummy items: Process simulations often contain models of items that are not actual plant equipment (see [Chapter 4](#)). For example, heat exchangers are sometimes modeled as a series of heaters and coolers linked by a calculator block as a means of checking for internal pinch points or allowing for heat losses to the ambient environment. When the simulation is mapped into ICARUS, dummy items should be excluded from the mapping process. In the previous example, only the heaters should be mapped, so as to avoid double counting the heat transfer area.

The default mapping can be edited by right-clicking on “Project Component Map Specifications” in the Project Basis/Process Design folder. A simulator model can be excluded from the mapping by selecting the item and then selecting “Delete All Mappings.” New mappings can be specified by selecting a simulator item and adding a new mapping.

To map loaded simulator data, click the map button on the toolbar (which maps all items) or right-click on an area or plant item in the process view window (which allows items to be mapped individually). If individual items are selected, then the user is given an option to use simulator data to override the default mapping in the Component

Map Specs file. This is useful for heat exchangers and other equipment where the simulator allows the equipment type to be specified.

The procedure for mapping equipment in Cleopatra is somewhat simpler and is illustrated in Example 7.7.

### 7.10.2 Design factors in ACCE

All good designs include an appropriate degree of overdesign to allow for uncertainties in the design data and method; see Section 1.6. For some equipment, the design factor or margin is specified by design codes and standards, for example, in the design of pressure vessels, as described in Chapter 14. In other cases, the design engineer must specify the degree of overdesign or margin based on experience, judgment, or company policy.

The equipment sizes calculated by a process simulator will be at the design flow rate unless a higher throughput was specified by the user, and hence include no design margin. The ACCE software adds an “equipment design allowance” to the equipment cost to allow for the design factor that will be introduced when the equipment is designed in detail. The equipment design allowance is based on the process description as follows:

New and unproven process	15%
New process	10%
Redesigned process	7%
Licensed process	5%
Proven process	3%

The process description is entered by right-clicking on “General Specs” in the Project Basis/Basis for Capital Costs folder.

The equipment design allowance is only applied to system-developed costs. If different design margins are needed for different equipment types, then the default should be set to “proven process,” and the equipment can then be oversized appropriately. Design margins can also be added to components using the ACCE custom model tool. Care should be taken to avoid adding more design margin than is necessary.

### 7.10.3 Pressure vessels

When costing pressure vessels such as reactors and distillation columns, care must be taken to ensure that the wall thickness is adequate. The default method in both ACCE and Cleopatra calculates the wall thickness required based on the ASME Boiler and Pressure Vessel Code Section VIII Division 1 method for the case where the wall thickness is governed by containment of internal pressure (see Chapter 14 for details of this method). If other loads govern the design, then the ACCE software can significantly underestimate the vessel cost. This is particularly important for vessels that operate at pressures below 5 bara, where the required wall thickness is likely to be influenced by dead weight loads and bending moments from the vessel supports, and for tall vessels such as distillation columns and large packed-bed reactors, where combined loading under wind loads may govern the thickness. Similarly, if the vessel is designed under a different section of the Boiler and Pressure Vessel Code, which is usually the case for vessels operated at high pressures, then ACCE can overestimate the vessel cost. It is important to always remember to enter the design pressure and temperature of the vessel, not the operating pressure and temperature.

The best approach to costing pressure vessels using commercial costing software is to enter all of the dimensions after completing the mechanical design of the vessel using the methods given in Chapter 14, or using suitable pressure vessel design software.

### 7.10.4 Nonstandard components in ACCE

Although ACCE contains over 250 equipment types, many processes require equipment that is not on the list of available project components. Also, in some cases the user will want to specify a certain make or model of equipment that may only be available in discrete sizes (for example, gas turbine engines or large pumps and compressors). In these situations, the nonstandard equipment can be included by setting up an Equipment Model Library (EML). Many companies maintain standard EMLs listing equipment that they often specify.

A new EML can be created by selecting the “Libraries” tab in the palette and opening the folder Cost Libraries/ Equipment Model Library. Right-clicking on either of the subfolders then allows the user to create a new EML in the appropriate set of units. Once an EML has been created, equipment items can be added to it. When a new item is added, a dialog box opens in which the user has to specify the sizing or costing method (linear, log-log, semi-log or discrete) and primary sizing parameters. Two costs and sizes must also be entered to establish the cost correlation.

Equipment model libraries are useful for completing an ACCE model of a process that contains nonstandard items. Care must be taken to update the EML costs so that they remain current.

---

### Example 7.7

Estimate the cost of a waste heat boiler designed to produce 4000 lb/h of steam. The exchanger area has been estimated as 1300 ft<sup>2</sup>.

#### Solution in ACCE

Starting from the ACCE project explorer window (on the far left of the screen), right-click on the Main Area and select Add Project Component (Fig. 7.4a).

Select Process Equipment, then Heat Exchangers (Fig. 7.4b). Select Waste Heat Boiler and enter a name (Fig. 7.4c).

Enter the size parameters and then click the Evaluate button (Fig. 7.4d). This runs the evaluator program and gives the results screen shown in Fig. 7.4e. The purchased equipment cost is \$145,900 on a Jan. 2006 USGC basis. The installed cost is \$196,225. Note that the installed cost is calculated directly by estimating bulk materials and labor rather than using an installation factor.

#### Solution in Cleopatra Enterprise

Starting from the Database Explorer (top-right menu), create a new document for the estimate (Fig. 7.4f, label A). Select the Estimating tab and then select the Equipment Knowledgebase tab, and in the Level tab select Unit-rates (Fig. 7.4f, label B). Then work through the drop-down menus to select Process Equipment / Special packaged mechanical equipment / Boilers / Steam boiler / Water tube packaged boiler (Fig. 7.4f, label C).

Clicking on the waste heat boiler then opens the menu for that item (Fig. 7.4g) and allows the user to add it to the shopping cart (label A) and open the data entry window by clicking the pyramid plus icon (label B). The equipment data can then be entered (Fig. 7.4h). The flow rate of 4000 lb/h corresponds to a duty of roughly 1220 kW, and we can select 40 bar as the operating pressure (a reasonable value for high-pressure steam). Cleopatra Enterprise then returns an equipment cost of \$119,000 on a Jan. 2018 U.S. basis, and we can add this into an overall estimate by clicking the pyramid button at the bottom right to give the results page shown in Fig. 7.4i.

The cost generated by Cleopatra Enterprise is a bare-module cost for the equipment. We would need to enter more details about the plant layout and associated instrumentation and piping to generate an installed cost (or use a suitable Hand factor).

Note that the two programs do not give identical answers. This is a common occurrence when using costing software, as the different programs are based on different data sets. An experienced cost engineer will benchmark the results from the software against recent cost data for equipment they have purchased and tune the costing models to more accurately fit their experience of local costs. Tuning of the cost models is beyond the scope of this book, but is typically covered in the software user manuals.

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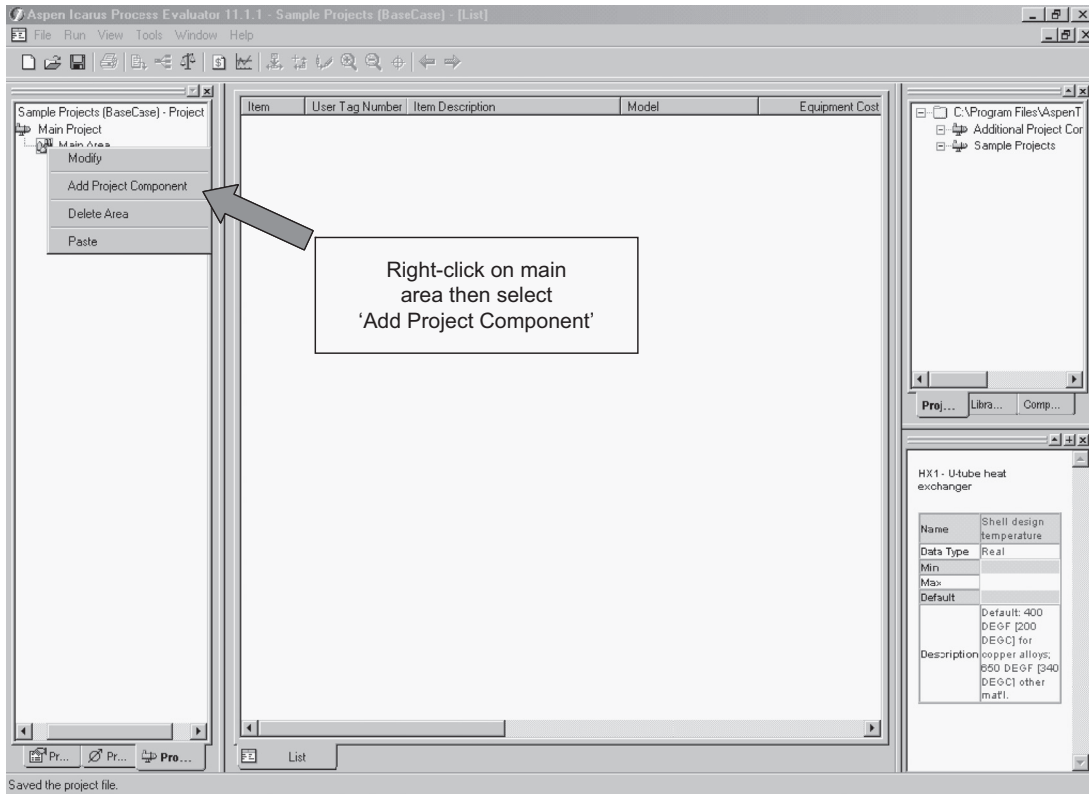
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## 7.11 Validity of cost estimates

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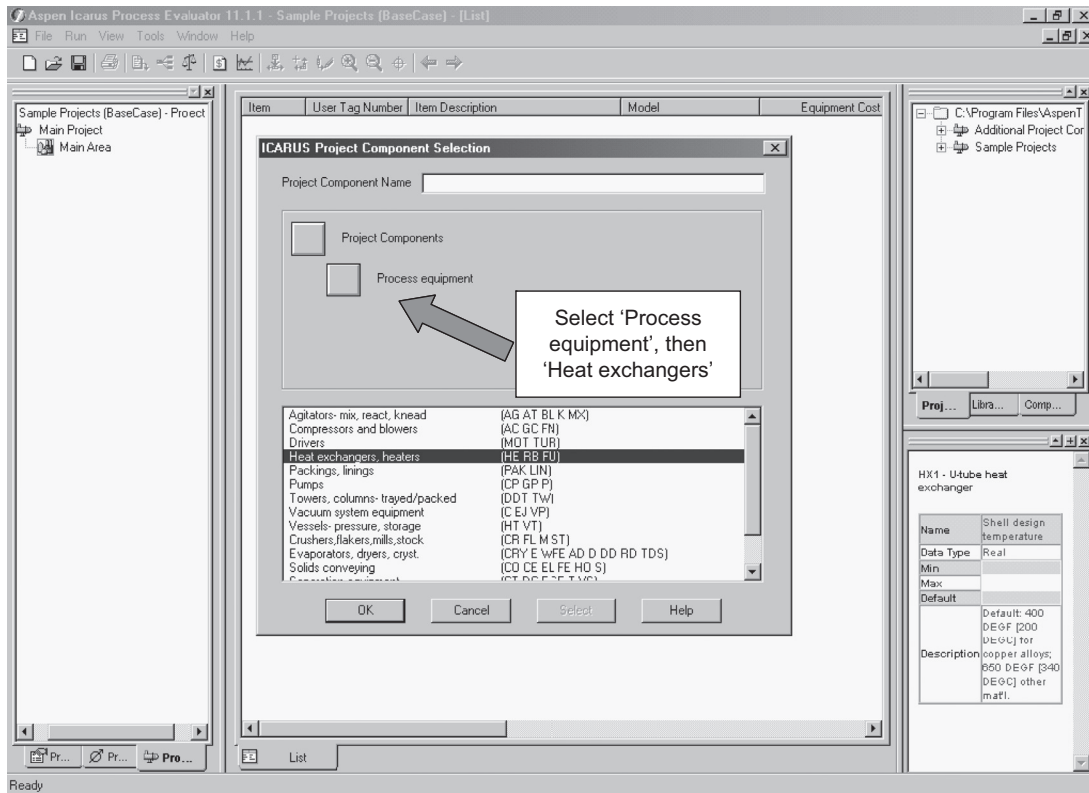
It should always be remembered that cost estimates are only estimates and are subject to error. An estimate should always indicate the margin of error. The error in a cost estimate is primarily determined by the degree of design detail that is available, and even a skilled estimator cannot estimate an accurate cost for an incomplete design.

When more design information has been developed, a professional cost engineer will be able to develop a more accurate estimate. The process design engineer should compare this estimate with the preliminary estimate to gain a better understanding of where the preliminary estimate could have been improved (either through capturing missing plant items or using better costing methods). This will help the design engineer to produce better preliminary estimates in the future.



(a)

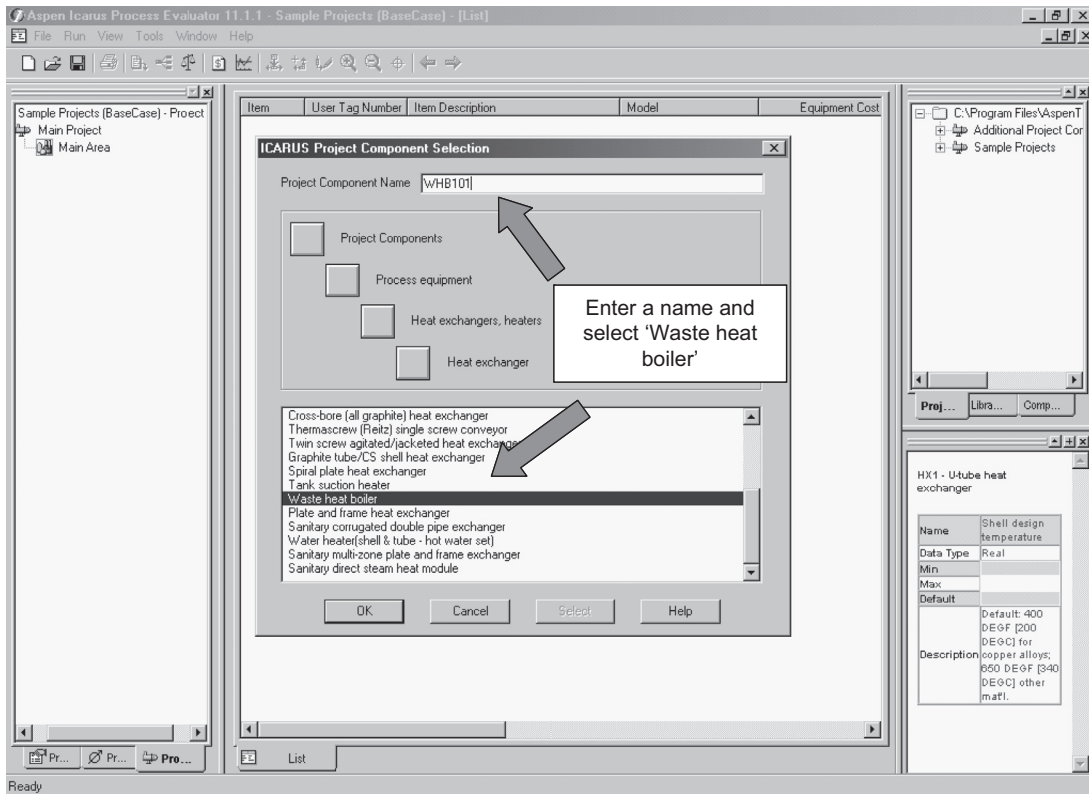
FIG. 7.4 (a–e) ACCE example. (f–i) Cleopatra Enterprise example.



(b)

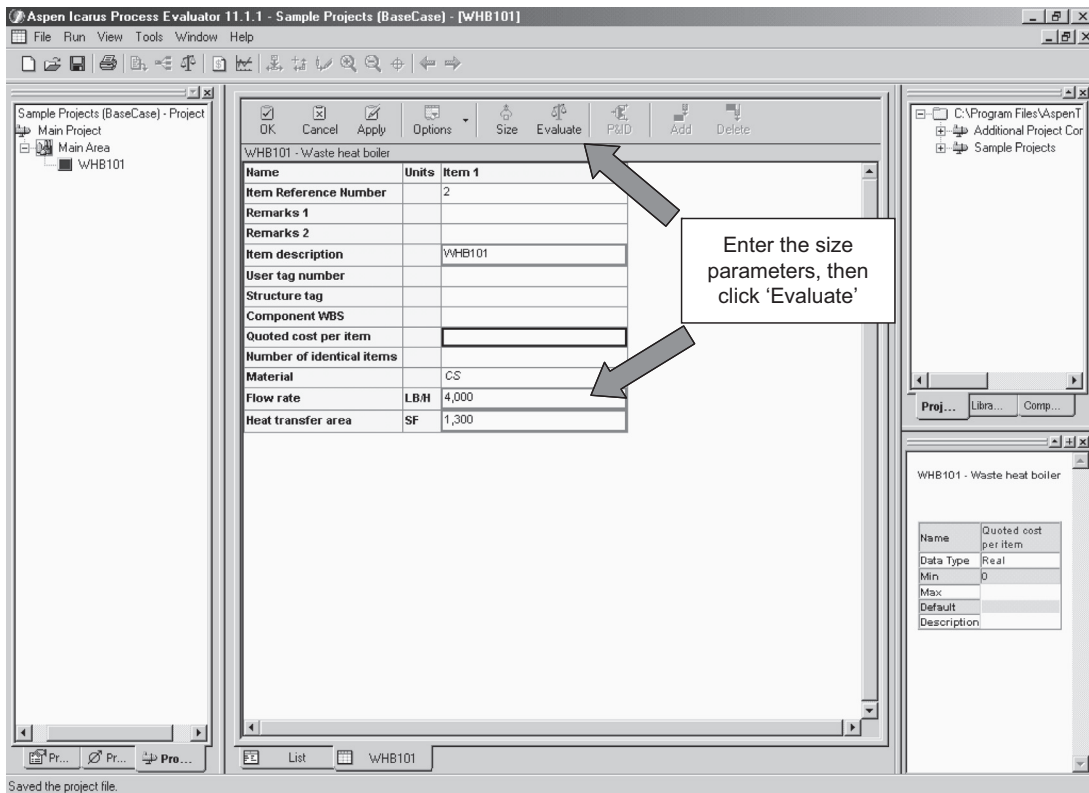
FIG. 7.4 Cont'd.





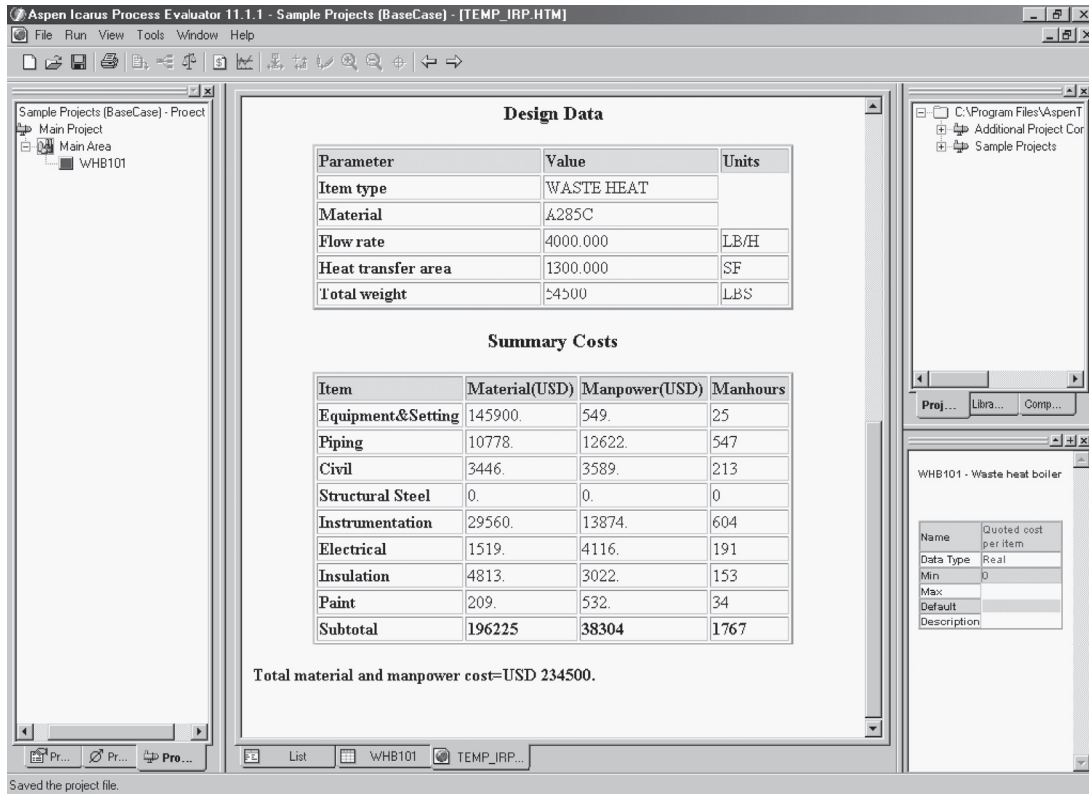
(c)

FIG. 7.4 Cont'd.



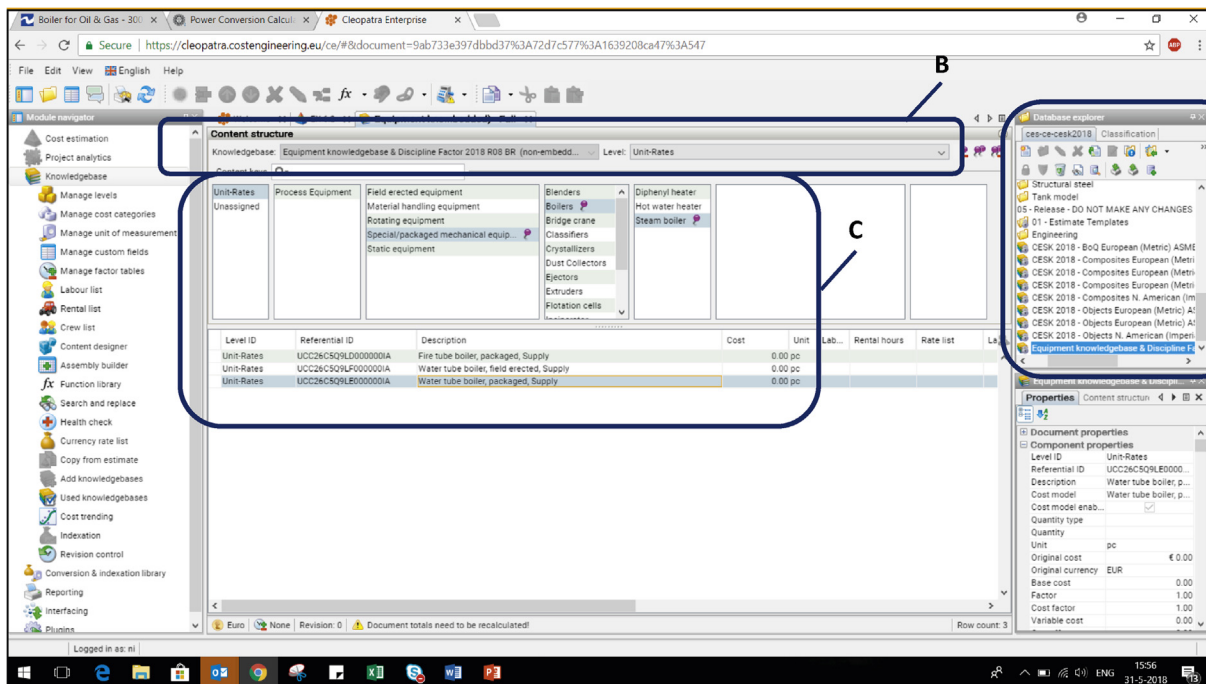
(d)

FIG. 7.4 Cont'd.



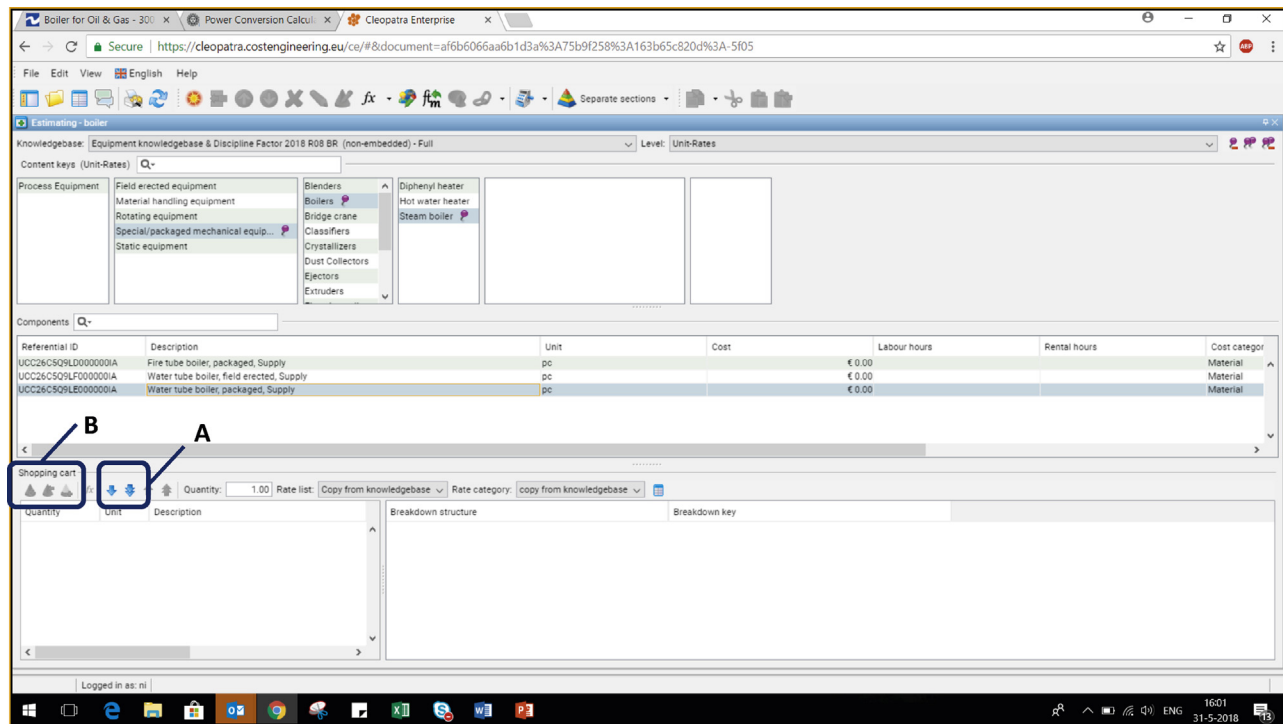
(e)

FIG. 7.4 Cont'd.



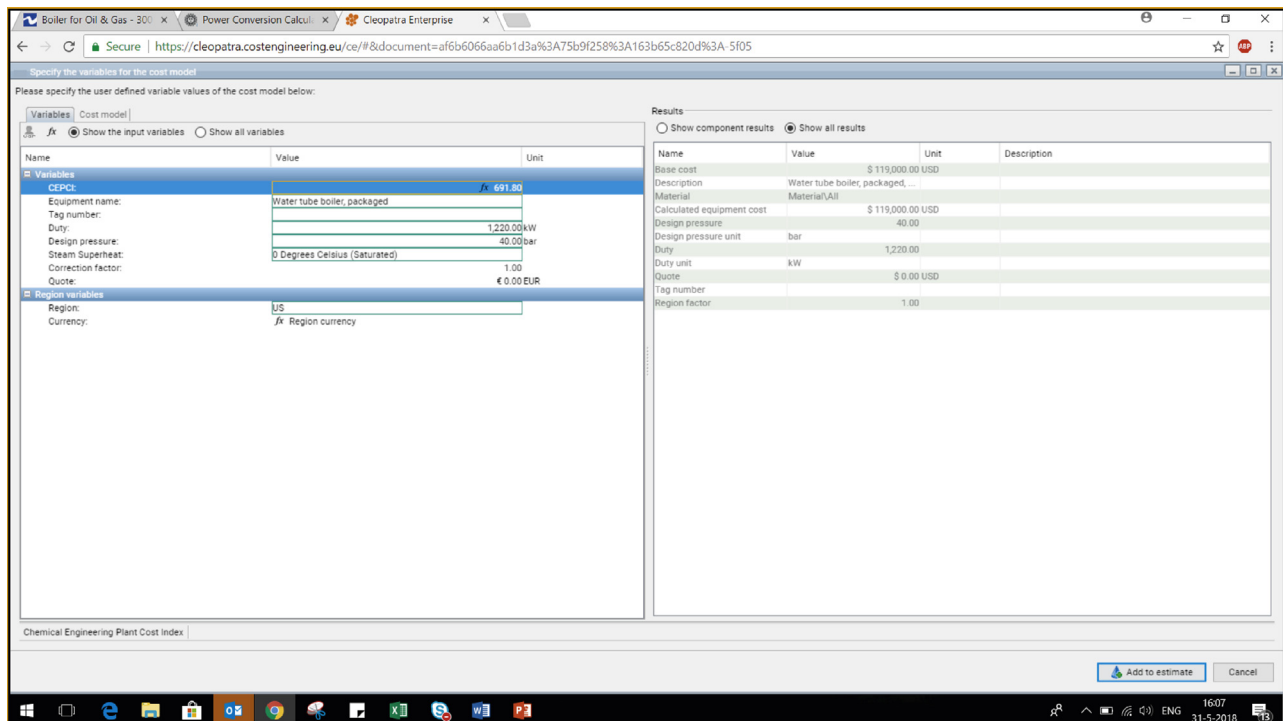
(f)

FIG. 7.4 Cont'd.



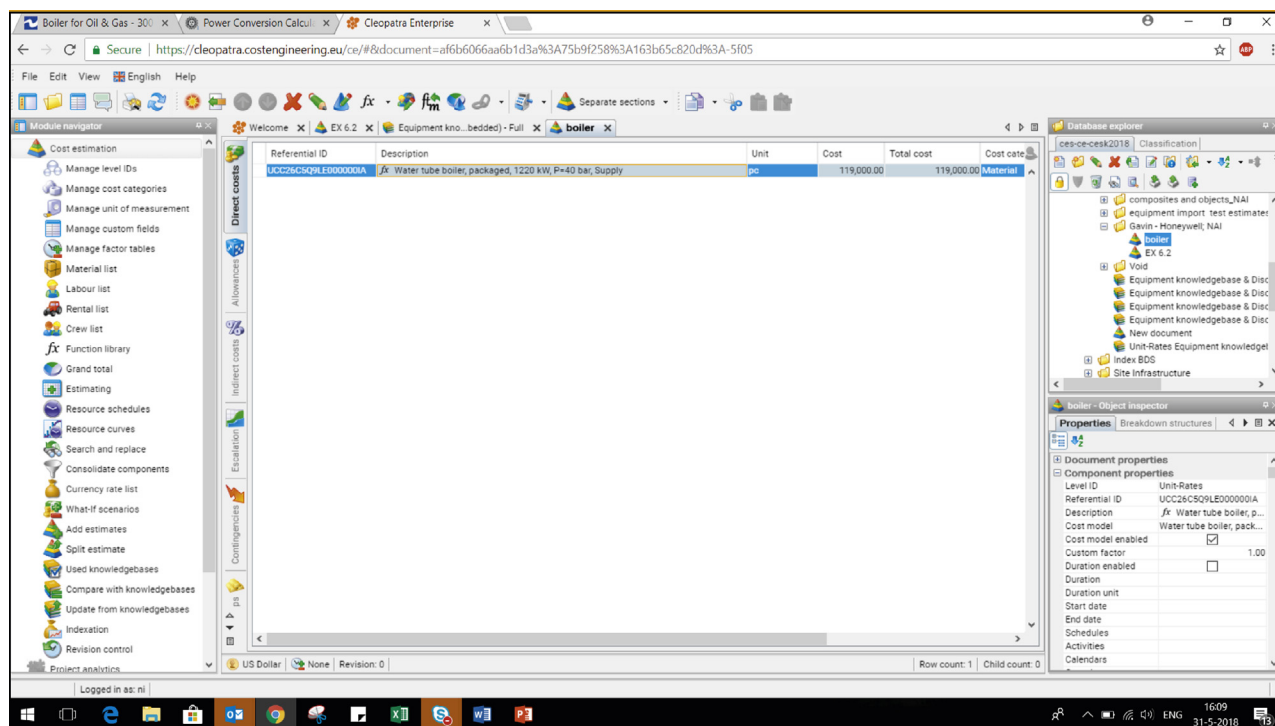
(g)

FIG. 7.4 Cont'd.



(h)

FIG. 7.4 Cont'd.



(i)

FIG. 7.4 Cont'd.

Additional resources for cost estimating are available from the various cost estimating associations: the AACE International ([www.aacei.org](http://www.aacei.org)); the Project Management Institute ([www.pmi.org](http://www.pmi.org)); the UK Association of Cost Engineers ([www.acoste.org.uk](http://www.acoste.org.uk)); and the International Cost Engineering Council (ICEC) ([www.icoste.org](http://www.icoste.org)). The ICEC website has links to cost engineering societies in 46 countries.

### Example 7.8

Adipic acid is used in the manufacture of nylon 6,6. It is made by hydrogenation of phenol to a mixture of cyclohexanol and cyclohexanone (known as *KA oil*—ketone and alcohol), followed by oxidation with nitric acid. Estimate the fixed capital cost for a 400,000 metric ton per year (400 kMTA) adipic acid plant located in northeast Asia in 2020.

### Solution

The capital cost of the process can be estimated based on historic data using the correlation given in Table 7.1. The correlation is based on the plant capacity in MMlb/y, so we need to convert the capacity: 400 kMTA is equal to 880 MMlb/y:

$$\text{ISBL capital cost} = 3.533 S^{0.6} = 3.533(880)^{0.6} = \$206.5 \text{ MM}$$

The ISBL cost is on a 2006 U.S. Gulf Coast basis, so we need to convert to a 2020 northeast Asia basis. If we look up the location factor in Table 7.7, it is not clear what factor we should use. The location factor for Japan is 1.26, whereas for China it varies from 0.6 to 1.1, depending on the amount of indigenous vs. imported equipment used. Because the exact location of the plant has not yet been specified, we are not able to make a definitive assessment of what the location factor should be. As a first approximation we therefore assume it is 1.0 and note that this should be revisited as part of the sensitivity analysis.

The OSBL capital cost is estimated as 40% of ISBL cost. The engineering cost and contingency are estimated as 10% and 15% of the sum (ISBL + OSBL) cost, respectively, giving a total fixed capital cost on a 2006 basis of \$361.3 MM.

Note that this cost is on a January 2006 basis, as that was the basis of the correlation in Table 7.1. We can update this to a 2020 basis using the CEPCI index. From Fig. 7.2 (or by looking up the index in *Chemical Engineering*):

$$\text{Index in 2006} = 499.6$$

Index for full year 2020 is not yet known, but is likely to be in the range of 570 to 600,

so as a conservative estimate we can assume Index in 2020 = 600.

$$\text{So, estimated total fixed capital cost in 2020} = \$361.3 \text{ MM} \times 600/499.6 = \underline{\underline{\$433.9 \text{ MM}}}.$$

The impact of updating this estimate is further explored in Problems 7.9 and 9.6.

## 7.12 References

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## 7.13 Nomenclature

		Dimensions in \$MLT <sup>0</sup>
$A$	Area	$L^2$
$a$	Constant	\$
$b$	Constant	\$
$C$	Capital cost	\$
$C_e$	Purchased equipment cost	\$
$C_{e,i,A}$	Purchased cost of equipment $i$ in alloy	\$
$C_{e,i,CS}$	Purchased cost of equipment $i$ in carbon steel	\$
$C_1$	Capital cost of plant with capacity $S_1$	\$
$C_2$	Capital cost of plant with capacity $S_2$	\$
$D_c$	Diameter of distillation column	L
$D_f$	Flange diameter	L
$D_n$	Nozzle diameter	L

Continued

		Dimensions in \$MLT <sup>θ</sup>
$D_s$	Shell diameter	L
$F$	Installation (Lang) factor	—
$f_c$	Installation factor for civil engineering work	—
$f_{el}$	Installation factor for electrical work	—
$f_{er}$	Installation factor for equipment erection	—
$f_i$	Installation factor for instrumentation and control	—
$f_l$	Installation factor for lagging, insulation and paint	—
$f_m$	Materials factor	—
$f_p$	Installation factor for piping	—
$f_s$	Installation factor for structures and buildings	—
$L_c$	Vessel length	L
$L_s$	Shell length	L
$LF_A$	Location factor for location <i>A</i> relative to U.S. Gulf Coast basis	—
$M$	Total number of pieces of equipment	—
$N$	Number of significant processing steps (functional units)	—
$N_b$	Number of bolts	—
$N_{baf}$	Number of baffles	—
$N_t$	Number of tubes	—
$n$	Capital cost exponent	—
$Q$	Plant capacity	MT <sup>-1</sup>
$S$	Plant or equipment capacity	*
$S_{lower}$	Lower end of capacity range over which correlation is valid	*
$S_{upper}$	Upper end of capacity range over which correlation is valid	*
$S_1$	Capacity of plant	1 *
$S_2$	Capacity of plant	2 *
$s$	Reactor conversion	—
TCOP	Total cost of production	\$M <sup>-1</sup> or \$T <sup>-1</sup>
$t$	Time, project life in years	T
$t_f$	Flange thickness	L
$t_s$	Shell thickness	L
$t_w$	Vessel wall thickness	L
$\rho$	Metal density	ML <sup>-3</sup>

Asterisk (\*) indicates that the dimensions are dependent on the type of equipment or process.

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## Acronyms

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AACEI Association for the Advancement of Cost Estimating International  
APC Automatic Process Control  
CEPCI Chemical Engineering Plant Cost Index, published monthly in *Chemical Engineering*  
EPC Engineering, Procurement and Construction  
FEED Front End Engineering Design  
ISBL Inside Battery Limits, referring to parts of the plant itself  
MM Million (abbreviation commonly used in industry in United States)  
M&S Marshall and Swift cost index, published monthly in *Chemical Engineering*  
NF Nelson-Farrer refinery cost index, published monthly in *Oil and Gas Journal*  
NWE Northwest Europe, often used as a location basis in capital costing  
OSBL Outside Battery Limits, referring to site infrastructure, also known as off-sites  
TCOP Total cost of production, including allowance for capital charges  
USGC U.S. Gulf Coast, often used as a location basis in capital costing  
WBS Work Breakdown Structure  
WC Working Capital

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## 7.14 Problems

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- 7.1. Estimate the capital cost of a plant that produces 80,000 metric tons per year of caprolactam.
- 7.2. The process used in the manufacture of aniline from nitrobenzene is described in Appendix F, design problem F.8. The process involves six significant stages:
- Vaporization of the nitrobenzene
  - Hydrogenation of the nitrobenzene
  - Separation of the reactor products by condensation
  - Recovery of crude aniline by distillation
  - Purification of the crude nitrobenzene
  - Recovery of aniline from wastewater streams
  - Estimate the capital cost of a plant to produce 20,000 metric tons per year.
- 7.3. A reactor vessel cost \$365,000 in June 1998; estimate the cost in January 2020.
- 7.4. The cost of a distillation column was \$225,000 in early 1998; estimate the cost in January 2024.
- 7.5. Using the data on equipment costs given in this chapter or commercial cost estimating software, estimate the cost of the following equipment:
1. A shell and tube heat exchanger, heat transfer area 50 m<sup>2</sup>, floating head type, carbon steel shell, stainless steel tubes, operating pressure 25 bar.
  2. A kettle reboiler: heat transfer area 25 m<sup>2</sup>, carbon steel shell and tubes, operating pressure 10 bar.
  3. A horizontal, cylindrical, storage tank, 3 m diameter, 12 m long, used for liquid chlorine at 10 bar, material carbon steel.
  4. A plate column: diameter 2 m, height 25 m, stainless clad vessel, 20 stainless steel sieve plates, operating pressure 5 bar.
- 7.6. Compare the cost of the following types of heat exchangers, each with a heat transfer area of 10 m<sup>2</sup>. Take the construction material as carbon steel.
1. Shell and tube, fixed head.
  2. Double-pipe.
- 7.7. Estimate the cost of the following items of equipment:
1. A packaged boiler to produce 20,000 kg/h of steam at 40 bar.
  2. A centrifugal compressor, driver power 75 kW.
  3. A plate and frame filter press, filtration area 10 m<sup>2</sup>.
  4. A floating roof storage tank, capacity 50,000 m<sup>3</sup>.
  5. A cone roof storage tank, capacity 35,000 m<sup>3</sup>.
- 7.8. A storage tank is purged continuously with a stream of nitrogen. The purge stream leaving the tank is saturated with the product stored in the tank. A major part of the product lost in the purge could be recovered by

installing a scrubbing tower to absorb the product in a solvent. The solution from the tower could be fed to a stage in the production process and the product and solvent recovered without significant additional cost. A preliminary design of the purge recovery system has been made. It would consist of:

1. A small tower 0.5 m diameter, 4.0 m high, packed with 25 mm ceramic saddles, packed height 3.0 m.
2. A small storage tank for the solution, 5 m<sup>3</sup> capacity.
3. The necessary pipe work, pump, and instrumentation.

All the equipment can be constructed from carbon steel.

Using the following data, evaluate how long it would take for the cost savings to pay for the capital cost of installing the recovery system:

1. Cost of product \$5 per lb.
2. Cost of solvent \$0.5 per lb.
3. Additional solvent make-up 10 kg/d.
4. Current loss of product 0.7 kg/h.
5. Anticipated recovery of product 80%.
6. Additional utility costs, negligible.

Other operating costs will be insignificant.

- 7.9. Example 7.8 developed the cost for an adipic acid plant on a northeast Asia January 2006 basis. Estimate the cost of the plant on a 2024 basis if the plant is located in Japan, South Korea, or China. In the case of China, assume that 85% of the plant cost can be sourced indigenously.
- 7.10. The production of methyl ethyl ketone (MEK) is described in Appendix F, Problem F.3. A preliminary design has been made for a plant to produce 10,000 metric tons per year (t/y). The major equipment items required are listed here. The plant operating rate will be 8000 hours per year. Estimate the capital required for this project.  
The plant will be built on an existing site with adequate infrastructure to provide the ancillary requirements of the new plant (no off-site investment is needed).

Major equipment items:

1. Butanol vaporizer: shell and tube heat exchanger, kettle type, heat transfer area 15 m<sup>2</sup>, design pressure 5 bar, material carbon steel.
2. Reactor feed heaters (two): shell and tube, fixed head, heat transfer area 25 m<sup>2</sup>, design pressure 5 bar, material stainless steel.
3. Reactors (three): shell and tube construction, fixed tube sheets, heat transfer area 50 m<sup>2</sup>, design pressure 5 bar, material stainless steel.
4. Condenser: shell and tube heat exchanger, fixed tube sheets, heat transfer area 25 m<sup>2</sup>, design pressure 2 bar, material stainless steel.
5. Absorption column: packed column, diameter 0.5 m, height 6.0 m, packing height 4.5 m, packing 25 mm ceramic saddles, design pressure 2 bar, material carbon steel.
6. Extraction column: packed column, diameter 0.5 m, height 4 m, packed height 3 m, packing 25 mm stainless steel pall rings, design pressure 2 bar, material carbon steel.
7. Solvent recovery column: plate column, diameter 0.6 m, height 6 m, 10 stainless steel sieve plates, design pressure 2 bar, column material carbon steel.
8. Recovery column reboiler: thermosyphon, shell and tube, fixed tube sheets, heat transfer area 4 m<sup>2</sup>, design pressure 2 bar, material carbon steel.
9. Recovery column condenser: double pipe, heat transfer area 1.5 m<sup>2</sup>, design pressure 2 bar, material carbon steel.
10. Solvent cooler: double pipe exchanger, heat transfer area 2 m<sup>2</sup>, material stainless steel.
11. Product purification column: plate column, diameter 1 m<sup>2</sup>, height 20 m, 15 sieve plates, design pressure 2 bar, material stainless steel.
12. Product column reboiler: kettle type, heat transfer area 4 m<sup>2</sup>, design pressure 2 bar, material stainless steel.
13. Product column condenser: shell and tube, floating head, heat transfer area 15 m<sup>2</sup>, design pressure 2 bar, material stainless steel.
14. Feed compressor: centrifugal, rating 750 kW.
15. Butanol storage tank: cone roof, capacity 400 m<sup>3</sup>, material carbon steel.
16. Solvent storage tank: horizontal, diameter 1.5 m, length 5 m, material carbon steel.
17. Product storage tank: cone roof, capacity 400 m<sup>3</sup>, material carbon steel.