Kirk-Othmer Encyclopedia of Chemical Technology.

PROCESS ENERGY CONSERVATION

The main driving force for increased energy conservation, which continues in times of both rising and falling energy prices, is broadscale technological process. Advances in technology are responsible for the historical rise in energy efficiency of 1–3% per year achieved by process industries. A wide range of big and little steps have contributed to these advances, such as improved gas turbine efficiency, structured packing in distillation (qv), computer control (see Process control), variable speed drives, computer design tools, and improved catalysts (see Catalysis) and synthetic processes for a variety of materials, eg, low density polyethylene (see Olefin polymers), acrylonitrile (qv), ammonia (qv), and acetic acid (see Acetic acid and derivatives).

The second force that has driven increased energy conservation is the trade of capital for energy. This trade is optimized within an existing technology and nets large increases when energy prices rise rapidly compared to capital price as in the 1975–1985 time period. The effect of energy usage on total cost is shown in Figure 1. If proper design is used, total costs are relatively tolerant of large deviations from the optimum design. For example, in Figure 1**a**, if the piping pressure drop is anywhere between one-third and three times the optimal, the penalty in total cost is $\leq 10\%$. In piping systems (qv), capital costs dominate over energy costs (1).

1. Energy Balance

Historically, an energy balance has been prepared for the components of a process primarily to ensure that heat exchangers and utility supply are adequate (see Energy management; Power generation; Process control). Often, an overall process energy balance was not developed. However, beginning in the mid-1980s, the energy balance for the overall process has become a document almost as important as the material balance. The overall energy balance serves as an evergreen framework during design to highlight the areas having greatest potential for improvement. Moreover, this document serves as a tool for plant-operating personnel after start-up, to aid optimization of energy use.

The energy balance should analyze the energy flows by type and amount, ie, present summaries of electricity, fuel gas, steam level, heat rejected to cooling water, etc. It should include realistic loss values for turbine inefficiencies and heat losses through insulation.

1.1. Exergy, Lost Work, and Second-Law Analysis

When energy is critically important to process economics, the simple energy balance is sometimes carried into an analysis of lost work. This compares the actual design against the theoretical ideal at each step and defines where the true energy use, or lost work, is occurring. In the discussions herein of reaction, separation, heat exchange, compression, refrigeration, and steam systems, the importance of this concept is illustrated. A few terms are defined below.

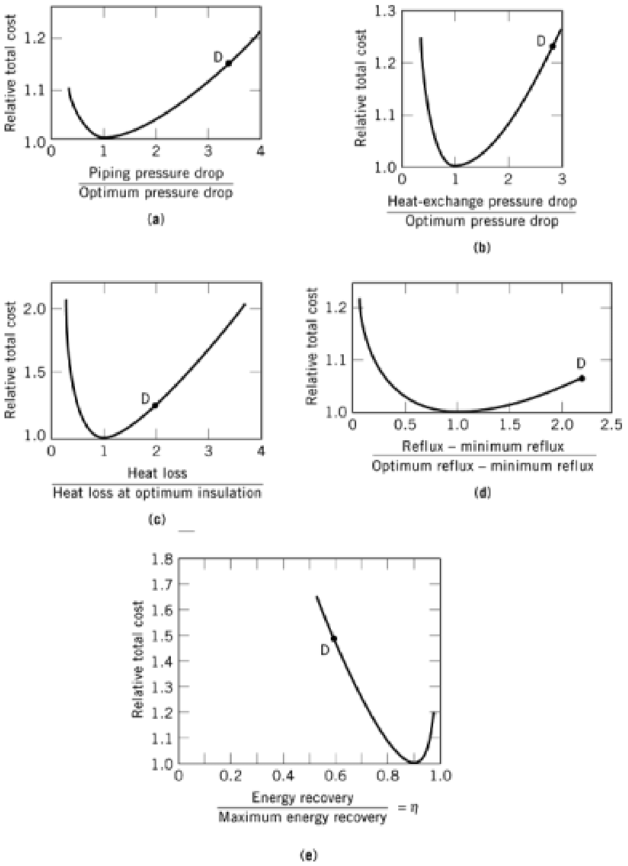

Fig. 1. Effect of energy use on total cost where total cost is the sum of capital and energy costs for the lifetime of the plant, discounted to present value. Point D corresponds to the design point if the designer uses an energy price that is low by a factor of four in projected energy price. Effects on costs of (**a**) pressure drop in piping, (**b**) pressure drop in exchangers, (**c**) heat loss through insulation, (**d**) reflux use, and (**e**) energy recovery through waste-heat boiler use.

Exergy, *E,* is the potential to do work. It is also sometimes called availability or work potential. Thermodynamically, this is the maximum work a stream can deliver by coming into equilibrium with its surroundings:

$$E = (H - H_0) - T_0\,(S - S_0)$$

where E = the maximum theoretical work potential; H and S = enthalpy and entropy of the stream at its original conditions; H_0 and S_0 = enthalpy and entropy of the same stream at equilibrium with the surroundings; and T_0 = temperature of the surroundings (sink).

Free energy, *G,* is a related thermodynamic property. It is most commonly used to define the condition for equilibrium in a processing step. It is identical to ΔE if the processing step occurs at T_0.

$$\Delta G = \Delta H - T\Delta S$$

Lost work, *LW,* is the irreversible loss in exergy that occurs because a process operates with driving forces or mixes material at different temperatures or compositions.

$$LW = E_{\text{in}} - E_{\text{out}}$$

Second-law analysis looks at the individual components of an overall process to define the causes of lost work. Sometimes it focuses on the efficiency of a step and ratios the theoretical work needed to accomplish a change, eg, a separation, to that actually used.

Sometimes it is more cost-effective simply to compare the design against a second-law violation checklist covering items such as mixing streams at different temperatures and compositions, high pressure drops in control valves, reactions running far from equilibrium, high temperature differentials, and pump-discharge recirculation (2).

2. Reactor Design for Energy Conservation

How closely a design approaches minimum energy is largely determined by the raw materials and catalyst system chosen. However, if reaction temperature, residence time, and diluent are the only variables, there is still a tremendous opportunity to influence energy use via the effect on yield. Even given none of these, there is still wide freedom to optimize the heat interchange system (see Reactor technology).

2.1. Design Variables

2.1.1. Maximizing Yield

Often the greatest single contribution to reduced energy cost is increased yield. High yield reduces the amount of material to be pumped, heated, and cooled while also simplifying downstream separation. This says nothing about the indirect energy reduction achieved through reduced raw material use. On average, the chemical industry uses almost as much energy in its raw materials as it does in direct purchases of fuel.

2.1.2. Minimizing Diluent

The case concerning diluent is less clear. A careful balance must be made of the benefits a diluent gives in higher yield against the costs needed for mass handling and separation.

2.1.3. Optimizing Temperature

Temperature is usually dictated by yield considerations. The choice of temperature for yield often overrides any desire to choose a temperature that minimizes the energy bill.

2.1.4. Heat Recovery and Feed Preheating

The objective is to bring the reactants to and from reaction temperature at the least utility cost, and to recover maximum waste heat at maximum temperature. The impact of feed preheating merits a more careful look. In an exothermic reaction, preheated feed permits the reactor to act as a heat pump, ie, to buy low and sell high. The most common example is combustion-air preheating for a furnace.

2.1.5. Batch vs Continuous Reactors

Usually, continuous reactors yield much lower energy use because of increased opportunities for heat interchange. Sometimes the savings are even greater in downstream separation units than in the reaction step itself. Especially for batch reactors, any use of refrigeration to remove heat should be critically reviewed. Batch processes often evolve little from the laboratory-scale glassware setups where refrigeration is a convenience.

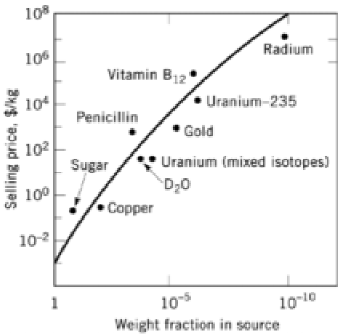

Fig. 2. Commercial selling prices of some separated materials (3). (Courtesy of McGraw-Hill Book, Co., Inc.)

3. Separation

About one-third of the chemical industry's energy is used for separation. A correlation exists between selling price and feed concentration (Fig. 2) as well as between selling price and product purity.

Concentration and purity can both be traced to the minimum work of separation, W, where T_0 is the sink temperature; N_i is the number of moles of a species present in the feed; $x_i = N_i / \Sigma N_i$; and γ_i is an activity coefficient. The value of W provides a target that is easily calculated and approachable in practice. For example, work calculated from this expression closely approaches the performance of a real-world distillation after inefficiencies for driving forces are taken into account.

$$W = RT_0 \Sigma N_i \ln (x_i \gamma_i)$$

For ideal solutions ($\gamma_i = 1$) of a binary mixture, the equation simplifies to the following, which applies whether the separation is by distillation or by any other technique.

$$W = RT_0 \left[x_1 \ln x_1 + (1 - x_1) \ln (1 - x_1) \right]$$

When a separation is not completed, less work is required. For x_1 equal to 0.5,

Product purity, %	Relative work
100	1
99.9	0.99
99.0	0.92
90.0	0.53

This relative work is an important consideration when comparing separation techniques. Some leave much of the work undone, as, for example, in crystallization (qv) involving an unseparated eutectic mixture.

3.1. Distillation

Distillation (qv) is by far the most common separation technique because of its inherent advantages. Its phase separation is clean, its equilibrium is closely approached in each stage, and its multistage countercurrent device is relatively easy to build.

Minimum work for an ideal separation at first glance appears unrelated to the slender vertical vessel having a condenser at the top and a reboiler at the bottom. The connection becomes evident when one calculates the work embedded in the heat flow that enters the reboiler and leaves at the condenser. An ideal engine can extract work from this heat.

$$E = QT_0 \left(\frac{1}{T_{\text{condenser}}} - \frac{1}{T_{\text{reboiler}}} \right)$$

Comparison of actual use of work potential against the minimum allows calculation of an efficiency relative to the best possible separation:

$$\eta = \frac{RT_0 \left[x_1 \ln x_1 + (1 - x_1) \ln (1 - x_1) \right]}{QT_0 \left(\frac{1}{T_{\text{condenser}}} - \frac{1}{T_{\text{reboiler}}} \right)}$$

There is still no obvious reason to believe that the efficiency of separating a mixture and an α (relative volatility) of 1.1 is related to that for an α of 2; however, it is known that when α is small, the required reflux and Q are large, but $(T_{\text{condenser}} - T_{\text{reboiler}})$ is small (see Distillation).

The two effects almost cancel one another to yield an approximation for the minimum work potential used in a distillation (3, 4).

$$E = RT_0 \left(1 + [\alpha - 1] x_1 \right)$$

When this is combined with the definition of minimum separation work, an approximation for distillation efficiency for an ideal binary can be obtained:

$$\eta = \frac{x_1 \ln x_1 + (1 - x_1) \ln (1 - x_1)}{1 + (\alpha - 1) x_1}$$

This efficiency is high and shows only minor dependence on α over a broad range of α. For $x_1 = 0.5$:

α	η
1.1	0.66
1.5	0.55
2.0	0.46

The dependence on x_1 is greater:

	$\alpha = 1.05$	$\alpha = 2$
η for $x_1 = 0.1$	0.32	0.30
η for $x_1 = 0.01$	0.056	0.053

These values, which match experience, suggest that distillation should be the preferred separation method for feed concentrations of 10–90%, but is probably a poor choice for feed concentrations of less than 1%. Techniques such as adsorption (qv), chemical reaction, and ion exchange (qv) are chiefly used to remove impurity concentrations of $<1\%$.

The high η values above conflict with the common belief that distillation is always inherently inefficient. This belief arises mainly because past distillation practices utilized such high driving forces for pressure drop, reflux ratio, and temperature differentials in reboilers and condensers. A real example utilizing an ethane–ethylene splitter follows, in which the relative number for the theoretical work of separation is 1.0, and that for the net work potential used before considering driving forces is 1.4.

$$\eta = \frac{\text{theoretical work}}{\text{net work potential used}} = \frac{1.0}{1.4} = 0.7$$

losses for driving forces:	
reflux above the minimum	0.1
exchanger ΔT	2.1
ΔP in tower	0.5
ΔP in condenser and tower	0.8
Total	*3.5*

$$\eta_{\text{includinglosses}} = \frac{1.0}{1.4 + 3.5} = 0.2$$

These numbers show that, first, the theoretical work can be closely approached by actual work after known inefficiencies are identified and, second, the dominant driving force losses are in pressure drop and temperature difference. This is a characteristic of towers having low relative volatilities.

3.1.1. Optimum Design

3.1.1.1. Condenser and Reboiler ΔT. The losses for ΔT are typically far greater than those for reflux beyond the minimum. The economic optimum for temperature differential is usually under 15°C, in contrast to the values of over 50°C often used in the past. This is probably the biggest opportunity for improvement in the practice of distillation. A specific example is the replacement of direct-fired reboilers with steam (qv) heat.

3.1.1.2. Adjusting Process to Optimize ΔT. At first glance, there appear to be only three or four utility levels (temperatures), and these can be 50°C apart. Different ways to increase the options include using multieffect distillation, which spreads the ΔT across two or three towers; using waste heat for reboil; and recovering energy from the condenser. To make these options possible, the pressure in a column may have to be raised or lowered.

3.1.1.3. Reflux Ratio. Generally, the optimum reflux ratio is below 1.15 and often below 1.05 minimum. At this point, excess reflux is a minor contributor to column inefficiency. When designing for this tolerance, correct vapor–liquid equilibrium (VLE) and adequate controls are essential.

3.1.1.4. State-of-the-Art Control. Computer control using feed-forward capability can save 2–20% of a unit's utilities by reducing the margin of safety (5). Unless the discipline of a controller forces the reduction of the safety margin, operators typically opt for increased safety. Operators are probably correct to do so when a proper set of analyzers and controllers has not been provided and maintained.

3.1.2. Right Feed Enthalpy

Often it is possible to heat the feed with a utility considerably less costly than that used for bottom reboiling. Sometimes the preheating can be directly integrated into the column-heat balance by exchange against the condensing overhead or against the net bottoms from the column. Simulation and careful examination of the overall process are required to assess the value of feed preheating.

A vapor feed is favored when the stream leaves the upstream unit as a vapor or when most of the column feed leaves the tower as overhead product. The use of a vapor feed was a key component in the high efficiency cited previously for the C_2 splitter, where most of the feed goes overhead.

3.1.3. Low Column Pressure Drop

The penalty for column pressure drop is an increase in temperature differential:

$$\Delta T = \left(\frac{dT}{dP}\right)\Delta P$$

$$\frac{dT}{dP} = \frac{R}{\Delta H}\,\frac{T^2}{P}$$

As this suggests, the penalty becomes large for low vapor pressure materials, ie, for components that are distilled at or below atmospheric pressure. The work penalty associated with this ΔT is approximately defined by the following ratio.

$$\frac{\Delta T_{\text{pressure drop}}}{T_{\text{reboiler}} - T_{\text{condenser}}} = \text{fraction of } W \text{ for } \Delta P$$

This penalty is greatest for close-boiling mixtures. A powerful technique for cutting ΔP is the use of packing. Conventional packings such as 5-cm (2-in.) pall rings can achieve a factor of four reductions over trays, and structured packing can achieve a factor of 10 reduction. Structured packing is more vulnerable to mistakes in detailed engineering and much less tolerant of fouling than trays. Almost 50% of the installations have encountered serious performance problems (6). It is also 2 to 10 times as expensive as the trays it typically replaces. However, despite these obstacles, structured packing is the biggest innovation in energy-saving hardware in the chemical processing industries. The overhead line and condenser pressure drop should be considered as well. (Note the high loss in the C_2 splitter example.)

3.1.4. Intermediate Condenser

As shown in Figure 3, an intermediate condenser forces the operating line closer to the equilibrium line, thus reducing the inherent inefficiencies in the tower. Using intermediate condensers and reboilers, it is possible to raise the efficiency above that for a simple reboiler–condenser system, particularly when the feed composition is far from 50:50 in a binary mixture.

	Maximum efficiency of heavy component in feed	
	50%	95%
one condenser, one reboiler	67	20
two condensers, one reboiler	73	47
three condensers, one reboiler	77	62

The intermediate condenser is most effective when a less costly coolant can be substituted for refrigeration.

3.1.5. Intermediate Reboiler

Inclusion of an intermediate reboiler moves the heat-input location up the column to a slightly colder point. It can permit the use of waste heat for reboil when the bottoms temperature is too hot for the waste heat.

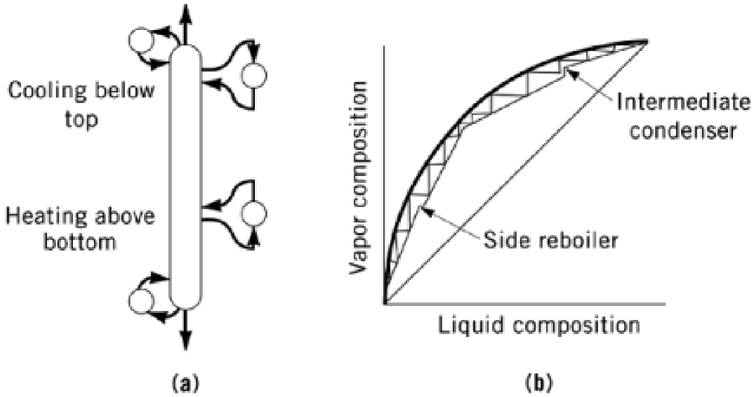

Fig. 3. (**a**) Schematic of an intermediate condenser and reboiler; (**b**) the corresponding vapor–liquid equilibrium.

3.1.6. Heat Pumps

Because of added capital and complexity, heat pumps are rarely economical, although they were formerly commonly used in ethylene/ethane and propylene/propane splitters. Generally, the former splitters are integrated into the refrigeration system; the latter are driven by low level waste heat, cascading to cooling water.

3.1.7. Lower Pressure

Usually, relative volatility increases as pressure drops. For some systems, a 1% drop in absolute pressure cuts the required reflux by 0.5%. Again, if operating at reduced pressure looks promising, the process can be evaluated by simulation. In a complete study of distillation processes, other questions that need to be asked include, Is the separation necessary? Is the purity necessary? Are there any recycles that could be eliminated? Can the products be sent directly to downstream units, thereby eliminating intermediate heating and cooling?

3.2. Other Separation Techniques

Under some circumstances, distillation is not the best method of separation. Among these instances are the following: when relative volatility is <1.05; when $<1\%$ of a stream is removed, as in gas drying (adsorption or absorption) or C_2H_2 removal (reaction or absorption); when thermodynamic efficiency of distillation is $<5\%$; and when a high boiling point pushes thermal stability limits. A variety of other techniques may be more applicable in these cases.

3.2.1. Reaction

Purification by reaction is relatively common when concentrations are low (ppm) and a high energy but low value molecule is present. Some examples are the hydrogenation of acetylene and the oxidation of waste hydrocarbons:

$$C_2H_2 + H_2 \longrightarrow C_2H_4$$

$$\text{waste hydrocarbon} + O_2 \longrightarrow H_2O + CO_2$$

3.2.2. Absorption

As a separation technique, absorption (qv), also called extractive distillation, starts with an energy deficit because the process mixes in a pure material (solvent) and then separates it again. This process is nevertheless quite common because it shares most of the advantages of distillation. Additionally, because it separates by molecular type, it can be tailored to obtain a high α. The following ratios are suggested for equal costs (7):

$\alpha_{\text{distillation}}$	$\alpha_{\text{extractiondistillation}}$	$\alpha_{\text{extraction}}$
1.2	1.4	2.5
1.4	1.9	5
1.6	2.3	8

In practice, most of the applications have come where a small part ($<5\%$) of the feed is removed. Examples include H_2S/CO_2 removal and gas drying with a glycol (see Distillation, azeotropic and extractive).

3.2.3. Extraction

The advantage of extraction is that a liquid is purified rather than a vapor, allowing operation at lower temperatures and the removal of a series of similar molecules at the same time, even though these molecules differ widely in boiling point. An example is the extraction of aromatics from hydrocarbon streams (see Extraction, liquid–liquid).

The disadvantage of extraction relative to extractive distillation is the greater difficulty of getting high efficiency countercurrent processing.

3.2.4. Adsorption

Adsorbents can achieve even more finely tuned selectivity than extraction. The most common application is the fixed bed with thermal regeneration, which is simple, attains essentially 100% removal, and carries little penalty for low feed concentration. An example is gas drying. A variant is pressure-swing adsorption. Here, regeneration is attained by a drop in pressure. By using multiple stages, high impurity rejection can be achieved, but at the expense of losing part of the desired product (see Adsorption).

Another approach is the simulated moving-bed system, which has large-volume applications in normal-paraffin separation and *para*-xylene separation. Since its introduction in 1970, the simulated moving-bed system has largely displaced crystallization in xylene separations. The unique feature of the system is that, although the bed is fixed, the feed point shifts to simulate a moving bed (see Adsorption, liquid separation).

3.2.5. Melt Crystallization

Crystallization (qv) from a melt is inherently more attractive than distillation because the heat of fusion is much lower than that of evaporation. It also benefits from lower operating temperature. In addition, organic crystals are virtually insoluble in each other so that a pure product is possible in a one-stage operation.

However, crystallization has unique disadvantages that outweigh its virtues and have sharply limited its application. Industry practice suggests the use of a workable alternative, if one exists. The disadvantages of melt crystallization include the following. (*1*) Difficulty of physical separation. Impure liquid is trapped as occlusion, and wets all crystal surfaces. (*2*) Requirement of a second separating process for eutectic mixture. The process thus resembles formation of two liquid phases. Although little energy is required to get the two phases, a great amount of it is required to finish the purification. (*3*) Difficulty of adding or removing heat on account of the thermal resistance of the crystal. (*4*) Difficulty of moving the liquid countercurrent to the crystals.

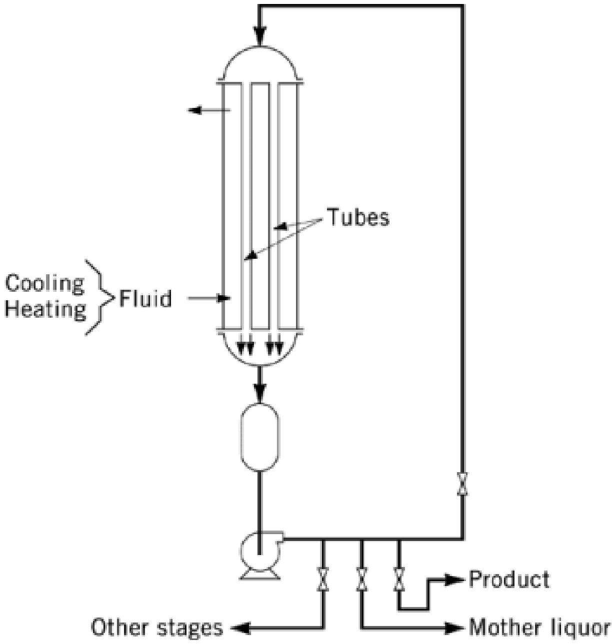

Fig. 4. Falling-film crystallizer, semibatch. Tube length is 12.2 m; tube diameter is 7.6 cm.

Thermodynamic efficiency is hurt by the large ΔT between the temperatures of melting and freezing. In an analogy to distillation, the high α comes at the expense of a big spread in reboiler and condenser temperature. From a theoretical standpoint, this penalty is smallest when freezing a high concentration (ca 90%) material.

One process, shown in Figure 4, is a semibatch operation in which liquid falls down the walls of long tubes. This permits both staged operation and sweating of crystals. Sweating is the removal of impurities by melting a small portion of crystals after mother liquor is first drained. The sweating operation washes residual mother liquor off the remaining crystals, and also removes some impurities from within the crystals. Typically, the sweating and staged operations require melting 5 kg of material for each kg of product (8).

3.2.6. Membranes

Liquid separation via membranes, ie, reverse osmosis (qv), is used in production of pure water from seawater. The chief limit to broader use of reverse osmosis is the high pressure required as the concentration of reject rises.

Mole fraction of reject	Minimum ΔP, MPa (psi)
0.05	7.6 (1100)
0.10	15.2 (2200)
0.20	31.8 (4600)

As a result, most systems are limited to achieving a mole fraction reject of 0.1 or less (see Membrane technology).

Membranes are also used to separate gases, for example, the production of N_2 and O_2 from air and the recovery of hydrogen from ammonia plant purge gas. The working principle is a membrane that is chemically tuned to pass a molecular type.

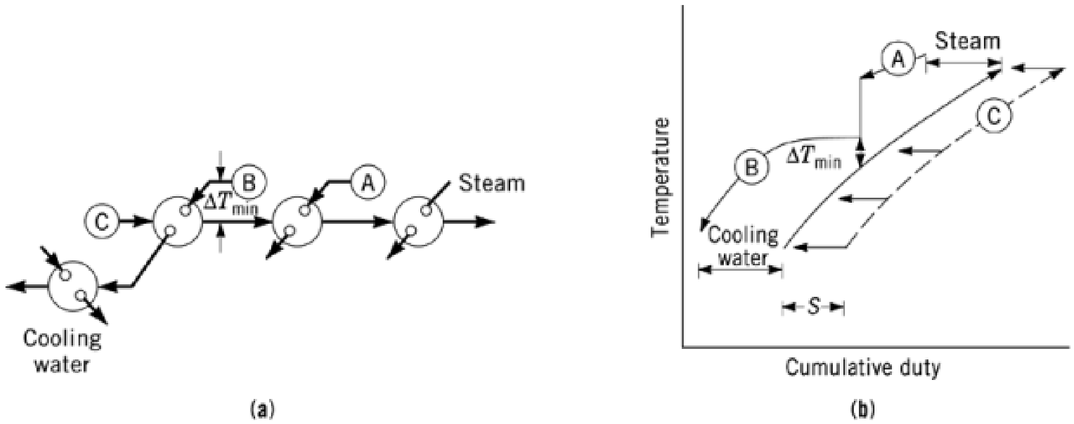

Fig. 5. Simple heat-exchange network where stream C is heated, and streams A and B are cooled: (**a**) schematic; (**b**) temperature–cumulative duty curve where S=savings in system and cooling water owing to driving to ΔT_{min}.

4. Heat Exchange

Most processing is thermal. Reaction systems and separation systems are typically dominated by the associated heat exchange. Optimization of this heat exchange has tremendous leverage on the ultimate process efficiency.

Heat exchangers use energy two ways: as frictional pressure drop, and as the loss in ability to do work when heat is degraded.

$$LW = QT_{\text{sink}}\left(\frac{1}{T_{\text{cold}}} - \frac{1}{T_{\text{hot}}}\right) + \text{frictional work for } \Delta P$$

In an optimized system, the lifetime value of the lost work associated with ΔT typically exceeds the cost of the heat exchanger. The lifetime value of the ΔP lost work in an optimized system is typically one-third as great as the heat exchanger capital (9). This means that when the costs for pumping power to overcome the heat exchanger ΔP (for the lifetime of the heat exchanger) are discounted to the time of heat exchanger purchase, their sum approximates one-third the heat exchanger cost. (This assumes a large heat exchanger designed for optimum pressure drop.)

The selection of design numbers for ΔP and ΔT is frequently the most important decision the process designer makes. The designer commonly becomes lost in the detail of tube length and baffle cut in an effort to optimize the hardware to meet a target, and spends far too little time on choosing that target.

4.1. Heat-Exchange Networks

A basic theme of energy conservation is to look at a process broadly, ie, to look at how best to combine process elements. The heat-exchange network analysis can be a useful part of this optimization. Figure 5 illustrates the basic concept of what the network analysis does. The analysis builds cumulative heating and cooling curves, pinching them together until a minimum ΔT is reached. This is discussed in greater detail elsewhere.

Network analysis, or pinch technology, has become an increasingly powerful approach to process design that includes most of the virtues of second-law analysis. For example, pinch technology can be broadened to include process revisions such as changing the temperature and pressure of distillation columns to fit into the natural cascade of high level heat dropping down to ambient (10). Other extensions of the concept include analysis of distillation column profiles, total site integration, and batch processing (11). The approach yields a quantitative estimate of readily achievable improvement. For example, in Figure 5, note the reduction in steam and cooling water obtained by driving the design to the pinch, ΔT_{min}.

4.2. Overdesign

Overdesign has a great impact on the cost of heat exchange and sometimes is confused with energy conservation, through lower ΔT and ΔP. The best approach is to define clearly what the objective of overdesign is and then to specify it explicitly. If the main concern is a match to other units in the system, a multiplier is applied to flows. If the concern is with the heat balance or transfer correlation, the multiplier is applied to area. If the concern is fouling, a fouling factor is called for. If low ΔT or ΔP is the principal concern, however, that should be specified. Adding extra surface saves energy only if the surface is configured to do so. Doubling the area may do nothing more than double the ΔP, unless it is configured properly.

4.3. ΔT and ΔP Optimization

Ideally, ΔT and ΔP are optimized by trying several values, making preliminary designs, and finding the point where savings in utility costs just balance the incremental surface costs. Where the sums at stake are large, this should be done. However, for many cases the simple guidelines given below are adequate. The primary focus is the impact of surface and utility prices; a secondary focus is the impact of fluid properties on heat-transfer coefficient (9).

4.3.1. Optimum ΔT

There are three general cases of high importance: the waste-heat boiler, in which only one fluid involves sensible heat transfer, ie, a temperature change; the feed–effluent exchanger, in which both fluids involve sensible heat transfer and are roughly balanced, ie, undergo essentially the same temperature change; and the reboiler, in which neither fluid involves a temperature change, ie, one fluid condenses and the other boils.

4.4. Waste-Heat Boiler

In a waste-heat boiler (Fig. 6), the approach ΔT sets both the amount of the unrecovered energy and the amount of heat-exchange surface. When terms are added for energy value, K_{v}, and surface cost, K_{l}, the optimum occurs when

$$\Delta T_{\mathrm{approach}} = \frac{K_l}{K_v}\,\frac{1.33}{U}$$

where K_l = annualized cost per unit of surface, $\$/(\mathrm{m}^2{\cdot}\mathrm{yr})$; K_v = annualized cost per unit of utility saved, $\$/(\mathrm{W}{\cdot}\mathrm{yr})$; and U = heat − transfer coefficient, $\mathrm{W}/(\mathrm{m}^2{\cdot}\mathrm{K})$. The factor 1.33 includes the value of the pressure drop for the added surface.

For example, the optimum $\Delta T_{\mathrm{approach}}$ is computed as follows:

$$K_l = \frac{\$215/\mathrm{m}^2}{2\ \mathrm{yr}} = \frac{\$107.5}{\left(\mathrm{m}^2{\cdot}\mathrm{yr}\right)}$$

$$K_v = \frac{0.017}{\mathrm{kW{\cdot}h}}{\cdot}8322\ \mathrm{h/yr} = \frac{\$142}{(\mathrm{kW{\cdot}yr})} = \frac{\$0.142}{(\mathrm{W{\cdot}yr})}$$

$$\Delta T_{\mathrm{approach}} = \frac{107.5/\left(\mathrm{m}^2{\cdot}\mathrm{yr}\right)}{0.142/(\mathrm{W{\cdot}yr})}\ \frac{1.33}{56.8\ \mathrm{W}/\left(\mathrm{m}^2{\cdot}\mathrm{K}\right)} = 17.7\ \mathrm{K}$$

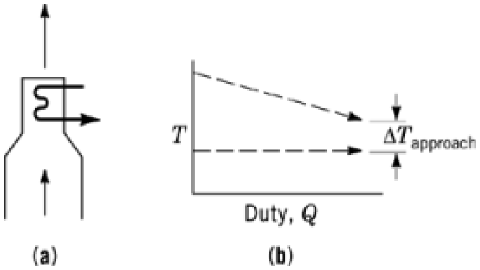

Fig. 6. ΔT in a waste-heat boiler: (**a**) schematic; (**b**) corresponding graphic representation.

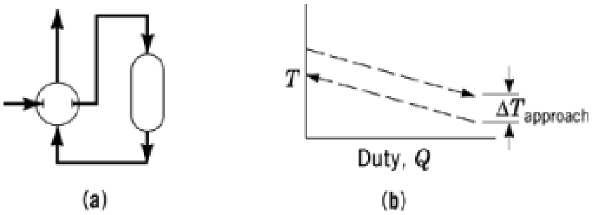

Fig. 7. ΔT in a feed–effluent exchanger: (**a**) schematic; (**b**) corresponding graphic representation.

where $U = 56.8\ \mathrm{W/(m^2{\cdot}K)}$ (10 Btu/(h·ft^2·°F)); surface cost = \$215/m^2 (\$20/ft^2); payout time = 2 yr; energy price = \$0.017/(kW·h) (\$5/10^6 Btu); and onstream time = 8322 h/yr. This case underlines a dramatic change in process design. Note that $\Delta T_{\text{approach}}$ varies with the first power of the ratio of surface price to energy price. The most visible result has been a change in typical fired heater design efficiency from 65–75% to 92–94%. A secondary result has been the appearance of waste-heat recovery units in many processes at the point where air coolers were once used.

4.5. Feed–Effluent Exchanger

The detailed solution for the optimum ΔT in a feed–effluent exchanger (Fig. 7) involves a quadratic equation for $\Delta T_{\text{approach}}$, but within the following restrictions.

$$0.8 < \frac{\Delta T_{\text{hot}}}{\Delta T_{\text{cold}}} < 1.25$$

$$\frac{T_{\text{hot}_{\text{in}}} - T_{\text{cold}_{\text{in}}}}{\Delta T_{\text{log mean}}} < 10$$

An excellent approximation is given by Reference 9.

$$\Delta T_{\text{log mean}} = \left[\frac{K_l}{K_v}\,\frac{1.33}{U}\left(T_{\text{hot}_{\text{in}}} - T_{\text{cold}_{\text{in}}}\right)\right]^{1/2}$$

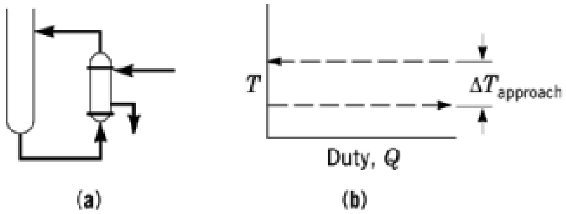

Fig. 8. ΔT in the reboiler: (**a**) schematic; (**b**) corresponding graphic representation.

For example, the optimum $\Delta T_{\text{logmean}}$ for a feed–effluent exchanger is computed as follows:

$$K_l = \frac{\$107.5/\text{m}^2}{2\ \text{yr}} = \$53.8/\left(\text{m}^2{\cdot}\text{yr}\right)$$

$$K_v = \frac{\$0.027}{(\text{kW}{\cdot}\text{h})}{\cdot}8322\ \frac{\text{h}}{\text{yr}}{\cdot}1000\ \frac{\text{kW}{\cdot}\text{h}}{\text{W}} = \frac{\$0.227}{(\text{W}{\cdot}\text{yr})}$$

$$\Delta T_{\text{log mean}} = \left(\frac{53.8}{0.227}\ \frac{1.33}{284}\,(200-100)\right)^{1/2} = 10.5^{\circ}\text{C}$$

where $T_{\text{hot}_{\text{in}}} = 200^{\circ}\text{C}$; $T_{\text{cold}_{\text{in}}} = 100^{\circ}\text{C}$; $U = 284\ \text{W}/(\text{m}^2{\cdot}\text{K})(50\ \text{Btu}/(\text{h}{\cdot}\text{ft}^2{\cdot}{}^{\circ}\text{F}))$; surface cost = $\$107.5/\text{m}^2$ ($\$10/\text{ft}^2$); payout time = 2 yr; energy price = $\$0.027/(\text{kW}{\cdot}\text{h})$ ($\$8/10^6$ Btu); onstream time = 8322 h/yr; and $\Delta T_{\text{hot}}/\Delta T_{\text{cold}} = 1.20$.

4.6. Reboiler

The case shown in Figure 8 is common for reboilers and condensers on distillation towers. Typically, this ΔT has a greater impact on excess energy use in distillation than does reflux beyond the minimum. The capital cost of the reboiler and condenser is often equivalent to the cost of the column they serve.

The concept of an optimum reboiler or condenser ΔT relates to the fact that the value of energy changes with temperature. As the gap between supply and rejection widens, the real work in a distillation increases. The optimum ΔT is found by balancing this work penalty against the capital cost of bigger heat exchangers.

If the Carnot cycle is used to calculate the work embedded in the thermal flows with the assumption that the heat-transfer coefficient, U, is constant and the process temperature is much greater than ΔT, a simple derivation yields the following:

$$\Delta T_{\text{optimum}} = T_p\left[\frac{K_l}{K_v U T_{\text{sink}}}\right]^{1/2}$$

where T_{sink} = temperature (absolute) at which heat is rejected; T_p = process temperature (absolute); K_l = annualized cost per unit of surface; and K_v = annualized value of power.

For utilities above ambient temperature,

$$K_v = K_p\ \text{(turbine efficiency)}$$

where K_{p} is the annualized cost of purchased power. The above relations typically give ΔT values in the 10–20°C range.

One strong caution is that the assumption of a constant U is usually inaccurate for boiling applications. Simulation is generally needed to fix ΔT accurately, particularly at ΔT values below 15°C.

4.7. Optimum Pressure Drop

For most heat exchangers there is an optimum pressure drop. This results from the balance of capital costs against the pumping (or compression) costs. A common prejudice is that the power costs are trivial compared to the capital costs. The total cost curve is fairly flat within ±50% of the optimum (see Fig. 1**b**), but the incremental costs of power are roughly one third of those for capital on an annualized basis. This simple relationship can be extremely useful in quick design checks.

The best approach is to have a computer program check a series of pressure drops and see how energy requirements decrease as surface increases. If this option is not available, the following simple method can be used to obtain specification sheet values. Start with a pressure drop of 6.9 kPa (1 psi), and apply three correction factors, $F_{\Delta T}$, F_{cost}, and F_{prop}, as follows.

$$\Delta P_{\text{optimum}} = 6.9\,(F_{\Delta T})\,(F_{\text{cost}})\,\left(F_{\text{prop}}\right)$$

The correction for temperature difference is given by the following.

$$F_{\Delta T} = \frac{T_{\text{in}} - T_{\text{out}}}{(T_{\text{hot}} - T_{\text{cold}})_{\text{mean}}}$$

This term is a measure of the unit's length. Sometimes it is referred to as the number of transfer units. This simply says that the optimum pressure drop increases as the heat exchanger gets longer, ie, has more transfer units. The forms of F_{cost} and F_{prop} both follow from the fact that in turbulent flow the heat-transfer coefficient varies approximately with (power dissipated/volume)$^{0.25}$ (9). The preexponential terms are empirical. They net a 35% higher $\Delta P_{\text{optimum}}$ than the (power dissipated/volume) correlation for heat transfer in a 1.5-cm diameter tube. The correction for costs is

$$F_{\text{cost}} = 0.017\left(\frac{\$/\left(\text{m}^2\cdot\text{yr}\right)}{\$/\text{kW}\cdot\text{h}}\right)^{0.75}$$

The correction for physical properties is

$$F_{\text{prop}} = \left(\frac{c}{c_w}\right)^{0.6}\left(\frac{k_w}{k}\right)^{0.6}\left(\frac{\mu}{\mu_w}\right)^{0.35}\left(\frac{\rho}{\rho_w}\right)^{0.5}$$

where c = specific heat, k = thermal conductivity, μ = viscosity, ρ = density, and c_{w}, k_{w}, μ_w, and ρ_w are the same properties for water at 25°C.

From these equations, the optimum ΔP for a feed–effluent exchanger, where the fluid has the physical properties of water and the following values:

$$T_{\text{in}} - T_{\text{out}} = 20°\text{C}$$

$$\Delta T = (T_{\text{hot}} - T_{\text{cold}})_{\text{mean}} = 10°\text{C}$$

$$\left\{\begin{array}{c}\text{surface cost} = \$215/\text{m}^2\\ \text{payout time} = 2\text{ yr}\end{array}\right\}\$107.5/\left(\text{m}^2\cdot\text{yr}\right)$$

$$\text{power cost} = \$0.03/(\text{kW}\cdot\text{h})$$

is calculated by

$$F_{\Delta T} = \frac{20}{10} = 2$$

$$F_{\text{cost}} = 0.017\left(\frac{107.5}{0.03}\right)^{0.75} = 7.8$$

$$F_{\text{prop}} = 1$$

$$\Delta P_{\text{optimum}} = 6.9\,(2)\,(7.8)\,(1) = 107.67\ \text{kPa}\ \ (15.6\ \text{psi})$$

If all else remains the same as above, except that the process fluid is a gas with $\mu = 0.02\,\mu_w$, $\rho = 0.00081\,\rho_w$, $c = 0.25\,c_w$, and $k = 0.066\,k_w$, then

$$F_{\text{prop}} = (0.25)^{0.6}\left(\frac{1}{0.066}\right)^{0.6}(0.02)^{0.35}\,(0.00081)^{0.5} = 0.016$$

$$\Delta P_{\text{optimum}} = 6.9\,(2)\,(7.8)\,(0.016) = 1.7\ \text{kPa}\ \ (0.25\ \text{psi})$$

The great impact of density in this example and in Table 1 should be noted. Probably the most common specification error is to use the large ΔP values characteristic of liquids in low density gas systems.

Table 1. Impact of Fluid Density on Optimum ΔP

Fluid	$\frac{\rho}{\rho_w}$	$\left(\frac{\rho}{\rho_w}\right)^{0.5}$	=	Relative optimum ΔP
water	1	$(1)^{0.5}$		1
oil	0.8	$(0.8)^{0.5}$		0.9
gas				
high pressure	0.05	$(0.05)^{0.5}$		0.22
atm pressure	0.001	$(0.001)^{0.5}$		0.03
vacuum	0.0002	$(0.0002)^{0.5}$		0.014

4.8. Fired Heaters

The fired heater is first a reactor and second a heat exchanger. Often, in reality, it is a network of heat exchangers.

4.8.1. Fired Heater as a Reactor

When viewed as a reactor, the fired heater adds a unique set of energy considerations, such as, Can the heater be designed to operate with less air by O_2 and CO analyzers? How does air preheating affect fuel use and efficiency? How can a lower cost fuel (coal) be used? Can the high energy potential of the fuel be used upstream in a gas turbine?

4.8.1.1. Air Preheating. Use of unpreheated air in the combustion step is probably the biggest waste of thermodynamic potential in industry (Table 2).

Table 2. Lost-Work Analysis for a Fired Heater

Parameter	Lost-work potential, %
combustion step	54
radiant section ΔT	7
convection section ΔT	24
stack losses (exit temp 225°C)	13
wall losses	2

Air preheating has the unique benefit of giving a direct cut in fuel consumed. It also can increase the heat-input capability of the firebox because of the hotter flame temperature. The drawback is that it tends to increase NO_x formation.

The most common type of air preheater on new units is the rotating wheel. On retrofits, heat pipes or hot-water loops are often more cost-effective because of ductwork costs or space limits.

Limitations in the material of construction make it difficult to use the high temperature potential of fuel fully. This restriction has led to the insertion of gas turbines into power generation steam cycles and even to the use of gas turbines in preheating air for ethylene-cracking furnaces.

4.8.2. Fired Heater as a Heat-Exchange System

Improved efficiency in fired heaters has tended to focus on heat lost with the stack gases. When stack temperatures exceed 150°C, such attention is proper, but other losses can be much bigger when viewed from a lost-work perspective. For example, a reformer lost-work analysis by Monsanto gave the breakdown shown in Table 2.

Losses for ΔT in the convection section are almost twice those for the hot exit flue gas. Furnace optimization is the clearest illustration of the benefits of lost-work analysis. If losses from a stack are nearly transparent, the losses embedded in an excessive ΔT in a convection section are even harder to identify. They do not show up on the energy balance that highlights the hot stack. These losses can be cut by adding surface to the convection section and shifting load from the radiant section, as well as by looking at the overall process (including steam generation) for streams to match the cooling curve of the flue gases.

Concern over corrosion from sulfuric acid when burning sulfur-bearing fuels often governs the temperature of the exit stack gas. However, the economics of heat recovery is so strong that flue gases are sometimes designed into the condensing range of weak sulfuric acid, recognizing that tube replacement will be required in the future.

Simple heat losses through the furnace walls are also significant. This follows from the high temperatures and large size of fired heaters, but these losses are not inevitable. In an optimized system, losses through insulation (1) are roughly proportional to

$$\left(\frac{\text{refractory price}}{\text{energy price}}\right)^{1/2}$$

This means that if the price ratio has decreased by a factor of four, then losses should be down by a factor of two. If the old optimum allowed a 2% loss, the new optimum would be closer to 1%.

4.9. Dryers

A drying operation (see Drying agents) needs to be viewed as both a separation and a heat-exchange step. When it is seen as a separation, the obvious perspective is to cut down the required work. This is accomplished by

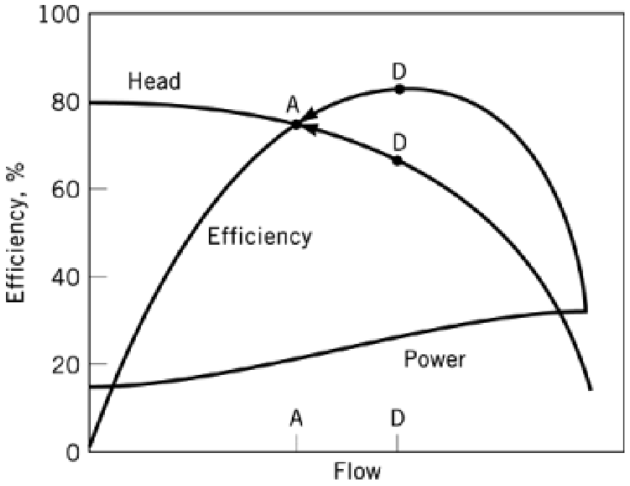

Fig. 9. Impact of excess design capacity on pump energy use, where A corresponds to actual operating flow, and D to the design point if the designer calls for 50 % more flow than the actual operating flow value.

mechanically squeezing out the water. The objective is to cut the moisture in the feed to the thermal operation to less than 10%. In terms of hardware, this requires centrifuges and filters, and may involve mechanical expression or a compressed air blow. In terms of process, it means big crystals.

When the dryer is seen as a heat exchanger, the obvious perspective is to cut down on the enthalpy of the air purged with the evaporated water. Minimum enthalpy is achieved by using the minimum amount of air and cooling as low as possible. A simple heat balance shows that for a given heat input, minimum air means a high inlet temperature. However, this often presents problems with heat-sensitive material and sometimes with materials of construction, heat source, or other process needs. All can be countered somewhat by exhaust-air recirculation.

Minimum exhaust-air enthalpy also means minimum temperature. If this cannot be attained by heat exchange within the dryer, preheating the inlet air is an option. The temperature differential guidelines of the feed–effluent interchange apply.

Like the fired heater, the dryer is physically large, and proper insulation of the dryer and its allied ductwork is critical. It is not uncommon to find 10% of the energy input lost through the walls in old systems.

5. Optimum Design of Pumping, Compression, and Vacuum Systems

5.1. Pumping

Many companies have optimum pipe-sizing programs, but in the absence of one, a good rule of thumb is that, in an optimized system, the annualized cost for pumping power should be one-seventh the annualized cost of piping (1). Piping is always a significant cost component and should be optimized (see Piping systems). Similarly, for an optimized heat exchanger, the annualized cost for pumping should be one-third the annualized cost of the surface for the thermal resistance connected with that stream.

The pump should be specified for the right flow. As Figure 9 shows, a 50% overdesign factor increases power by 35% in a combination of higher head and lower efficiency (see Pumps).

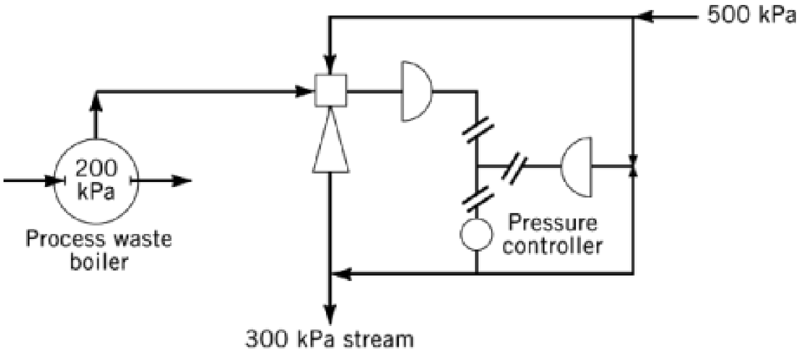

Fig. 10. A thermocompressor. To convert kPa to psi, multiply by 0.145.

If the allowance for control can be reduced, it should be. One option is the use of variable-speed drives. This eliminates the control valve and its pressure drop and piping. Its best application is where a large share of the head is required for friction and where process demands cause the required flow to vary.

5.2. Compression

The work of compression is typically compared against the isentropic–adiabatic case.

$$\eta_{\text{comp}} = \frac{W_{\min}}{E_{\text{out}} - E_{\text{in}}}$$

For an ideal gas, this can be expressed in terms of temperatures

$$\eta_{\text{comp}} = \frac{W_{\min}}{W_{\text{actual}}} = \frac{T_{\text{in}}\left(\left(\frac{P_{\text{out}}}{P_{\text{in}}}\right)^{R/c} - 1\right)}{T_{\text{out}} - T_{\text{in}}}$$

where *R*/*c* is the ratio of gas constant to molar specific heat. Minimum work is directly proportional to suction temperature. This means that close temperature approaches are justified on suction coolers.

Sometimes $W_{\min}$ for compression is expressed for the isothermal case, which is always lower than that for the adiabatic case. The difference defines the maximum benefit from interstage cooling.

The measuring of temperature rise permits monitoring efficiency for a fixed pressure ratio and suction temperature. Efficiencies should always exceed 0.6, and 1.00 is approachable in reciprocating devices. Their better efficiency needs to be balanced against their greater cost, greater maintenance, and lower capacity.

5.3. Thermocompressors

A thermocompressor is a single-stage jet using a high pressure gas stream to supply the work of compression. One application is in boosting waste-heat-generated steam to a useful level. An example is shown in Figure 10. Thermocompressors can also be used to boost a waste combustible gas into a fuel system by using high pressure natural gas. The mixing of the high energy motive stream with the low energy suction stream inherently involves lost work, but as long as the pressures are fairly close, the net efficiency for the device can be respectable (25–30%). Here, efficiency is defined as the ratio of isentropic work done on the suction gas to the isentropic work of expansion that could have been obtained from the motive gas. The thermocompressor has the advantage of no moving parts and low capital cost.

5.4. Vacuum Systems

The most common vacuum system is the vacuum jet. Because of the high ratio of motive pressure to suction pressure, the efficiency of vacuum systems is generally only 10–20%. The optimum system often employs several stages with intercondensers. Steam use in this range varies roughly as $(1/P)^{0.3}$, where P is absolute suction pressure. Like the thermocompressor, the steam jet has the advantage of no moving parts. However, the velocity in the steam nozzle (the part through which the steam discharges) is extremely high, which makes it subject to erosion and replacement every few years to maintain efficiency (see Vacuum technology).

Because of the low efficiency of steam-ejector vacuum systems, there is a range of vacuum above 13 kPa (100 mm Hg) where mechanical vacuum pumps are usually more economical. The capital cost of the vacuum pump goes up roughly as $(\text{suction volume})^{0.6}$ or $(1/P)^{0.6}$. This means that as pressure falls, the capital cost of the vacuum pump rises more swiftly than the energy cost of the steam ejector, which increases as $(1/P)^{0.3}$. Usually below 1.3 kPa (10 mm Hg), the steam ejector is more cost-effective.

Other factors that favor the choice of the steam ejector are the presence of process materials that can form solids or require high alloy materials of construction. Factors that favor the vacuum pump are credits for pollution abatement and high cost steam. The mechanical systems require more maintenance and some form of backup vacuum system, but these can be designed with adequate reliability.

6. Refrigeration

Refrigeration is a high value utility (see Refrigeration and refrigerants). The value of heat in a hot stream is the amount of work it can surrender:

$$\frac{W}{Q} = \left(\frac{T - T_{\text{sink}}}{T}\right) \eta_{\text{turbine}}$$

and the value of refrigeration is the work required to heat-pump it to the sink temperature:

$$\frac{W}{Q} = \left(\frac{T_{\text{sink}} - T}{T}\right) \frac{1}{\eta_{\text{compressor}}} \frac{1}{\eta_{\text{fluid}}}$$

The value of refrigeration is compared to heating in Figure 11 for $\eta_{\text{turbine}} = \eta_{\text{compressor}} = 0.7$ and for $\eta_{\text{fluid}} = 0.8$. In Figure 12, η_{fluid} accounts for cycle inefficiencies such as the letdown valve.

Because of its value, refrigeration justifies thicker insulation, lower ΔT values in heat exchange, and generally much more care in engineering (12). The designer should ensure that the capital cost of the refrigeration users has been optimized and integrated with the cost of the refrigeration system, and that the cost of supplying power to the refrigeration machine driver has been integrated with the refrigeration system optimization. It is also good to ask, Is refrigeration really necessary? Can river water or cooling-tower water be used directly for part of the year? Can part of the refrigeration be replaced? Can the refrigerant-condensing temperature be reduced during part of the year? Can the system be designed to operate without the compressor during cold weather? Is a central system more efficient than scattered independent systems? Does the control system cut required power for part-load operations? Are enough gauges and meters provided to monitor operation? Is there an abundance of waste heat (above 90°C) available from the plant? If so, refrigeration can be supplied by an absorption system.

Absorption chiller units (Fig. 13) need 1.6–1.8 J (0.38–0.43 cal) of waste heat per joule (0.24 cal) of refrigeration. Commercially available LiBr absorption units are suitable for refrigeration down to 4.5°C. For low level waste heat (90–120°C), absorption chillers utilize waste heat as efficiently as steam turbines using mechanical

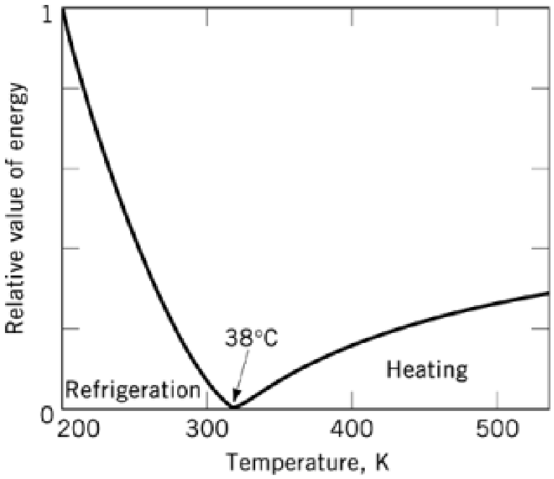

Fig. 11. Relative value of energy at various temperatures.

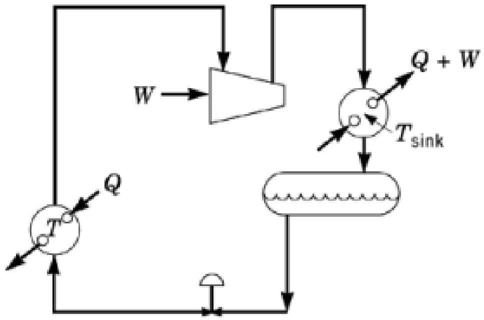

Fig. 12. Schematic of compression refrigeration.

refrigeration units. Absorption refrigeration using 120°C saturated steam delivers 4.5°C refrigeration, having an efficiency, with respect to the work potential in the steam of 35%.

7. Steam and Condensate Systems

In the process industry, steam (qv) serves much the same role as money does in an economy, ie, it is the medium of exchange. If its pricing fails to follow common sense or thermodynamics, strange design practices are reinforced. For example, many process plants employ accounting systems where all steam carries the same price regardless of temperature or pressure. This may be appropriate in a polymer or textile unit where there is no special use for the high temperature; it is wrong in a petrochemical plant.

Some results of the constant-value pricing system are as follow: generation in a central unit at relatively low pressure, <4.24 MPa (600 psig); tremendous economic pressure to use turbines rather than motors for drives; lack of incentive for high efficiency turbines; excessively high temperature differentials in steam users; tremendous incentive to recover waste heat as low pressure steam; and a large plume of excess low pressure steam vented to the atmosphere.

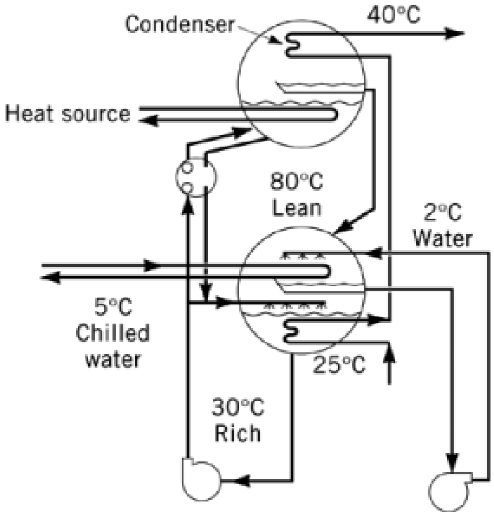

Fig. 13. Absorption refrigeration.

A number of alternative pricing systems have been proposed that hinge on turbine efficiency and the relative pricing of fuel and electricity. All pricing systems have problems, which suggests that the best system is a simpler one that relates the value of steam to that at the generation pressure by its work potential (exergy content).

$$\frac{\text{value at pressure}}{\text{value at highest generation pressure}} = \frac{\text{exergy at pressure}}{\text{exergy at highest generation pressure}}$$

Design of a central power steam system is beyond the scope of this discussion, but the interaction between the steam system and the process must be considered at all stages of design. There is a long list of factors to consider in designing a steam system (see also Energy management):

Is there a computer program that monitors the system and advises on what turbines to operate and how to minimize steam venting?

Can a gas turbine be utilized for power generation upstream of the boiler?

Can steam-use pressure be lowered? (If ΔT in the heater is above 20°C, the steam pressure is probably above the economic optimum.)

Are there any turbines under 65% efficiency? (Turbines are often limited to sizes above 500 kW, where good efficiencies can be obtained; they are used for smaller drives only where they are essential to the safe shutdown of the unit.)

Are there waste streams with unutilized fuel value that can be burned in the boilers?

Is there a program to monitor turbine efficiency by checking temperatures in and out?

Is condensate recovered?

Is the flash steam from condensate recovered?

Is feedwater heating optimized?

Is there any pressure letdown without power recovery?

Has enough flexibility been built into the overall condensing turbine system? (The balance changes over the history of a unit as a process evolves, generally in the direction of less condensing demand.)

Is steam superheat maintained at the maximum level permitted by mechanical design?
Can a thermocompressor be used to increase steam pressures from waste heat?
Are all users metered?
Is low level process heat used to preheat deaerator makeup?
Are ambient sensing valves used to turn off steam tracer systems?

8. Cooling-Water Systems

Cooling water is a surprisingly costly utility. On the basis of price per unit energy removed, it can cost one-fifth as much as the primary fuel. Roughly half of this cost is in delivery (pump, piping, and power). This fact has several important implications for design. Heat exchangers should be designed to use the available pressure drop. A heat exchanger that is designed for 10 kPa (1.45 psi) when 250 kPa (36 psi) is available will have five times the design flow. If an exchanger cannot be economically designed to use available ΔP, orifices should be provided to balance the system. This can be done without compromising the guidelines that no unit should be designed for less than 0.6 m/s on tubeside or less than 0.3 m/s on shellside. If temperature requirements permit, the system will cost less to operate with exchangers in series. An installed flow-measuring element is usually justified. If only part of the system requires a high head, this could be supplied by a booster pump. The whole system need not be designed for the high head.

Other energy considerations for cooling towers include the use of two-speed or variable-speed drives on cooling-tower fans, and proper cooling-water chemistry to prevent fouling in users (see Water, industrial water treatment). Air coolers can be a cost-effective alternative to cooling towers at 50–90°C, just below the level where heat recovery is economical.

9. Special Systems

9.1. Heat Pumps

A heat pump is a refrigeration system that raises heat to a useful level. The most common application is the vapor recompression system for evaporation (qv) (Fig. 14). Its application hinges primarily on low cost power relative to the alternative heating media. If electricity price per unit energy is less than 1.5 times the cost of the heating medium, it merits a close look. This tends to occur when electricity is generated from a cheaper fuel (coal) or when hydroelectric power is available.

Use in distillation systems are rare. The reason is the recognition that almost the same benefits can be achieved by integrating the reboiling–condensing via either steam system (above ambient) or refrigeration system (below ambient).

In an optimized system, where T_{hot} and T_{cold} are in absolute units, K, the following is true.

$$\frac{Q}{W} = \frac{T_{\text{hot}}\eta_{\text{compressor}}}{T_{\text{hot}} - T_{\text{cold}}}$$

This provides another criterion for testing whether a heat-pump system may be cost-effective. A power plant takes three units of Q to yield one unit of W. Therefore, to provide any incentive for less overall energy use, Q/W must be far in excess of 3.

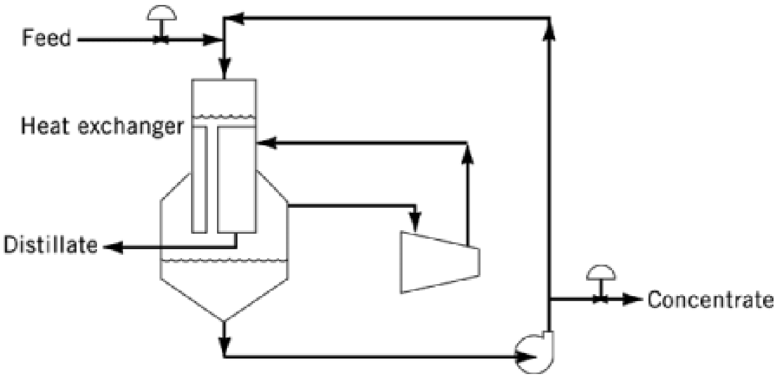

Fig. 14. Vapor recompression evaporator system.

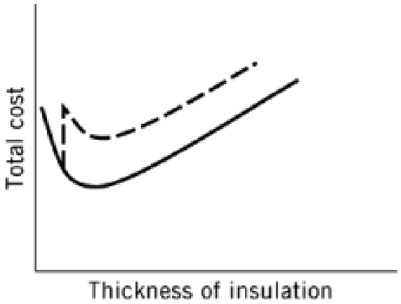

Fig. 15. Tank insulation costs: existing (‐) vs new (▬). Total cost represents the sum of heat loss and insulation.

9.2. Energy Management Systems

The reduction in computing costs has made it possible to do a wide range of routine monitoring and controlling. For example, a distillation system can be monitored continuously, the energy use can be compared against an optimum, and the cost-per-hour deviation from the optimum setpoint can be displayed.

10. Existing Plants

Good design ideas for new plants are also good for existing plants, but there are three basic differences. (*1*) Because a plant already exists, the capital–operating cost curve differs. Usually, this makes it more difficult to reduce utility costs to as low a level as in a new plant. (*2*) The real economic justification for change is more likely to be obscured by the plant accounting system and other nontechnical inputs. (*3*) The real process needs are measurable and better defined.

An example in support of the first point is the case of optimum insulation thickness. A tank, optimally insulated when first installed, can fall below optimal if the value of heat is quadrupled. This change can justify twice the old insulation thickness on a new tank. However, the old tank may have to function with its old insulation. The reason is that there are large costs associated with preparation to insulate. This means that the cost of an added increment of insulation is much greater than assumed in the optimum insulation thickness formulas (Fig. 15).

An example of the second difference is that many things appear to be strongly justified by savings in low pressure steam if the steam is valued artificially high. A designer of a new plant has the advantage of focusing attention on savings in the primary budget items, ie, fuel and electricity at the plant gate, rather than on cost-sheet items such as steam at battery limits.

The third difference is that many process details are relatively uncertain when a plant is designed. For example, inert loading for vacuum jets is rarely known to within 50%. Although the first two differences are negative, the third provides a unique opportunity to measure the true need and revise the system accordingly.

11. Energy Audit

The energy audit has seven components: as-it-is balance, field survey, equipment tests, checking against optimum design, idea-generation meeting, evaluation, and follow-up.

11.1. As-It-Is Balance

This is a mandatory first step for the energy audit. It permits the targeting of principal potentials; checking of use against design; checking of use against optimum, ie, how a new plant would be designed; definition of possible hot or cold interchanges; definition of unexpected uses, eg, the large steam purge to process or high pressure drop exchanger; and contribution from specialists not familiar with the unit.

11.2. Field Survey

This is often done by a team of two: one who knows what to look for and one who knows the process. They should look and listen for things such as air leaks, high pressure drops across values, frost on piping, lights of the wrong type or lights on at the wrong time, steam plumes (a reason to climb to the top of the unit), and minimum-flow bypasses in use. They should also talk to the operators and ask such questions as, What runs when the unit is down? What happens when the reflux is cut? Where are the guidelines for steam–feed ratio? How close is the unit run to these guidelines? The field survey should develop detailed repair lists for leaking traps, uninsulated metal, lighting, and steam leaks.

11.3. Equipment Tests

Procedures for rigorous, detailed efficiency determination are available (ASME Test Codes) but are rarely used. For the objective of defining conservation potentials, relatively simple measurements are adequate. For fired heaters, stack temperature and excess O_2 in stack should be measured; for turbines, pressures (in and out) and temperatures (in and out) are needed.

11.4. Checking Against Optimum Design

This attempts to answer the question whether a balance needs to be as it is. The first thing to compare against is the best current practice. Information is available in the literature (13) for large-volume chemicals such as NH_3, CH_3OH, urea, and ethylene. The second step is to look for obvious violations of good practice on individual pieces of equipment. Examples of violations are $\text{stack temperatures} > 150^\circ\text{C}$; $\text{process streams} > 120^\circ\text{C}$, cooled by air or water; $\text{process streams} > 65^\circ\text{C}$, heated by steam; $\eta_{\text{turbine}} < 65\%$; $\text{reflux ratio} > 1.15 \text{ times minimum}$; and $\text{excess air} > 10\%$ on clean fuels.

11.5. Idea-Generation Meeting

This meeting has the following guidelines: gather people with expertise and experience, discuss the as-it-is balance for each area, record all ideas, and assign follow-up responsibilities.

11.6. Evaluation

The evaluation of each idea should include a technical description as well as its economic impact and technical risk. Ideas should be ranked for implementation. A report should provide a five-year framework for energy projects.

11.7. Follow-Up

If no savings result, the effort has been wasted. Thus the audit leader must ensure that the potential of every good design idea is recognized by management and the project-generation channels of the company.

11.8. Perspective

Often, what seems like negligence can be a tried-and-proven practice. Lists offering huge savings are frequently not accurate. The process has to work, and safety cannot be compromised for energy savings. If a change to save energy is justified, the control and hardware revisions that permit it to be implemented safely are also justified. Similarly, present utility savings may or may not compensate for future repair bills or lost products. For example, an idling turbine may be necessary to permit a safe plant shutdown if a power failure occurs; a cooling-water flow that is throttled to below 0.6 m/s in winter is likely to require a heat-exchanger cleaning in late spring; a furnace that runs too low on excess air may run into after-burning; and a column run too close to the minimum reflux ratio without adequate controls runs a risk of off-specification product.

The plant accounting system often needs questioning. All energy is not created equal. The energy that is recovered from flashed steam or that is shaved off a reboiler's duty may not be worth its cost-sheet value. The meters that matter are the primary meters at the plant gate. Only if the recovered energy reduces the plant gate meters does it save the plant money. Solutions to an energy waste problem must fit into the over plant–energy balance.

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