

Chapter 2

Process Specification

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2.1 Introduction

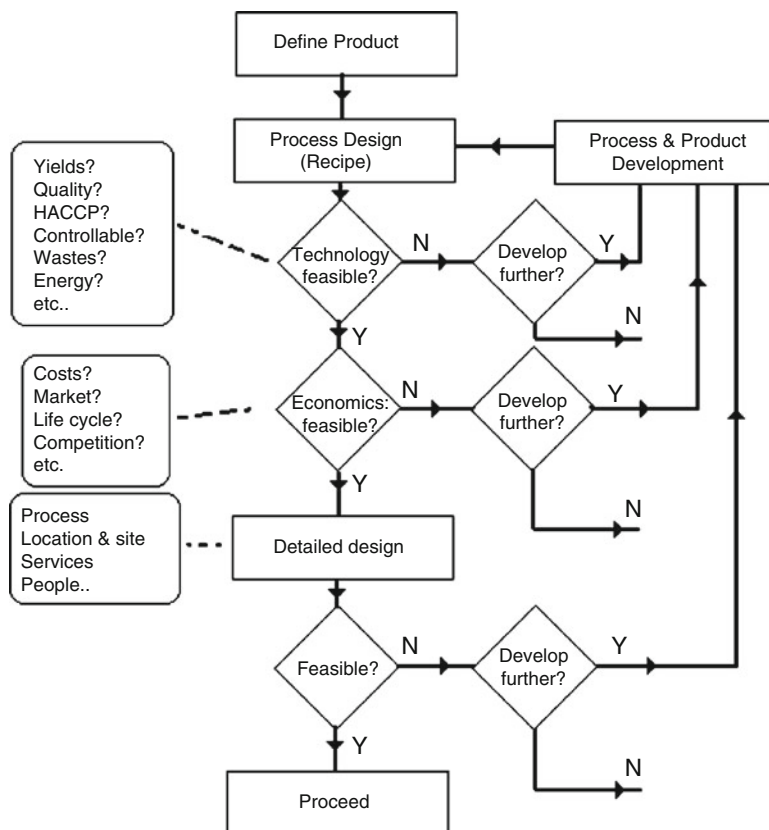
2.1.1 *Evolution of the Design*

Food processing is concerned with transforming raw materials into edible, safe, and nutritious products to meet a human and/or market need. The design problem is to establish and specify the mix of operations (i.e., machines) and material requirements, which, with appropriate scheduling, can produce defined quantities of the required products with assured quality and form. Very few factories produce only one product or an unchanged product mix day in, day out: different raw materials are available at different times of the year; the market demands variety, and few products are made on a sufficient scale to merit a dedicated line. Thus, it is not uncommon to find many recipe and/or product changes on a production line. This may involve using the same equipment (after a cleaning cycle); it may involve changes in the food processing operations or their sequence. The “recipe” is thus the specification of the materials and the operating sequence. The products must meet defined quality measures, implying defined levels of consistency, hygiene, and control in the production process. The processes should therefore be flexible and robust. In other words, they must be able to cope with variations in raw materials and other disturbances. The production system should also be efficient in the use of materials, energy and other services; rapid and efficient product changeover will be important; materials and other aspects of processing history should be traceable. Ideally the factories will be flexible enough to cope with new products. Above all, the process must meet defined economic objectives within the resource constraints on people, equipment and services.

Many food operations are inherently risky and it is therefore very important that, at defined intervals, the process equipment can be thoroughly and reliably cleaned. Hence, CIP must be included as part of the design remit from the outset. Also, most processing lines involve a mix of continuous and batch or semi-batch operations. This poses special problems for process operability, scheduling, and control.

The design of a production facility evolves through a series of iterations (Fig. 2.1), beginning with a definition of the products and their recipes. This leads to a simplified flowsheet. From this, preliminary estimates of the materials, energy and service requirements can be produced. The flowsheet can also

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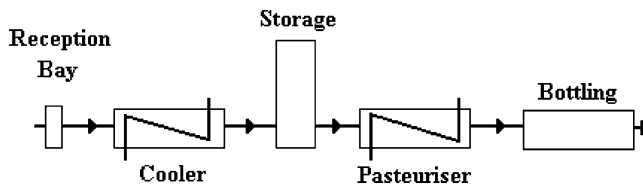
Fig. 2.1 The design cycle

be used at an early stage for the preliminary study of important aspects such as microbial (and other) hazards and their control, process control and operational feasibility. These form the basis for a preliminary economic analysis. From this, the flowsheet is further refined and more detailed analysis of all these aspects can be pursued. The design specification thus develops through a hierarchy of levels. Here we focus mainly on the first stages of design, i.e., defining a process to a stage at which specification of the details of the process and its operation can proceed.

2.1.2 Flowsheeting

Usually the specification of an outline flowsheet is relatively straight forward, in that there are relatively few really novel products. Most “new” food products are developments from existing products and processes, involving either new ingredients and operating methods or sometimes new technology at one stage of the sequence. Of course, at a more detailed level, the differences between one company’s process and another’s may be considerable, even if this is not immediately obvious from the outline flowsheet. A key stage in developing successful new products is to solve the scaling problem, i.e., to find and use the rules which ensure satisfactory development from the kitchen or product development laboratory to the industrial scale, producing hundreds of kilograms or tonnes of products. Although many of the basic rules of scale up are understood, the complexity (and rapid product cycle) associated with food processing means that it is often unsafe to jump straight from laboratory to production scale. In other words, be careful before you drop the pilot-scale trials.

Fig. 2.2 Milk bottling line



Two examples of flowsheets are shown here. Figure 2.2 is a simplified, outline flowsheet of a line to produce and bottle pasteurized milk. The milk is received from a tanker where it is held in cool storage until the pasteurizing and bottling line becomes available or the production schedule (of which, more later) demands it. Then the milk is pasteurized continuously en route to the bottling plant. Note that this flowsheet is extremely basic: it does not show any of the necessary CIP features, waste streams, services, or alternative feed streams. At this level, the flowsheet is simply a representation of the processing sequence; no further implications (e.g., that there is only one storage tank, that milk from only one source is to be used, or that the pasteurizer is dedicated to this line, etc.) should be drawn. On the other hand, together with the product specification, it does embody sufficient features of the process to enable *preliminary* estimation of the material and energy requirements (i.e., how many bottles, how much steam and cooling water or refrigerant are required per tanker load) to be made.

The second example is a more detailed flowsheet of a (hypothetical) potato-frying process. Figure 2.3 is a block diagram listing the sequence of operations, while Fig. 2.4 shows the main process vessels, lines, and service supplies, but not the detailed instrumentation. This flowsheet is sufficiently detailed to permit a reasonably accurate assessment of the material and energy requirements, and equipment sizing. Also the flowsheet can be used to form the basis for HACCP (*Hazard Analysis and Critical Control Points*) and related quality and process control studies. In the case of the frying process, many of the process steps shown in Fig. 2.4 can be identified as critical control points (CCPs). However, the principal concern is for the post-frying steps, since snacks are susceptible to post-processing contamination, and none of the processing steps after the fryer can positively reduce or eliminate the hazards. This has clear implications for *process* control since the HACCP analysis is based on a presumption that the various stages are operated in the way the design team intends.

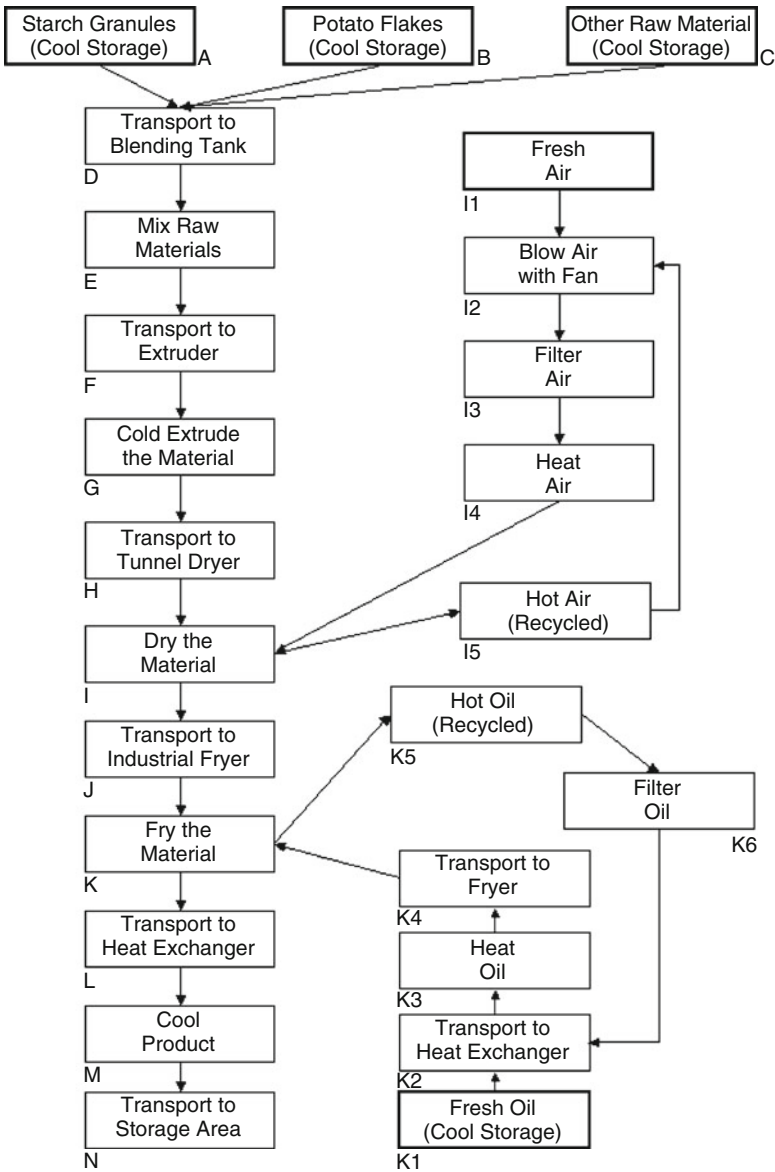
Traditionally, HACCP analyses have been monitored and recorded manually, despite the fact that several good computer-driven HACCP analysis programs are available. Table 2.1 lists a number of Web sites that advertize such software. This list is not intended to be comprehensive and the reader is advised to undertake a detailed and up-to-date search to satisfy his requirements. In 2005, the International Organization for Standards published the ISO 22000 Food Safety Management Systems Standard, which supersedes the HACCP principles promulgated by the Codex Alimentarius Commission in 1993. The principal differences between this Standard, which was designed for easy incorporation into the ISO 9001 quality management system, and HACCP have been described by Blanc (2006).

Today we are in a position where all the process monitoring—including all the HACCP-driven, etc., actions—can be monitored and recorded electronically. This must be the way forward in developing efficient, integrated, and traceable systems.

2.2 Batch Processes: Scheduling and Its Implications

As already noted, batch and semi-batch operations are common in the food industry. This is because it makes a wide range of products on demand, often requiring relatively short processing runs, and because regular cleaning cycles are needed to maintain hygienic conditions. Equipment is shared between different products; there are frequent start-ups and shutdowns. Decisions have to be made

Fig. 2.3 Potato frying process



constantly as to which product to make, which tanks and processing equipment are to be used, and so on. The combination of many different but similar products and their perishability implies that time spent in the warehouse and distribution networks should be minimized; this intensifies the pressure on the production system.

It is important that these features are fully recognized at the design stage and here we concentrate on one or two simple examples to illustrate the methods and issues involved. The easiest way of visualizing a batch sequence is by means of a Gantt chart. In this diagram, the usage (including filling and emptying) of all the principal items of equipment is plotted versus time. In a multiproduct plant, different colors can be used to show their processing history.

Consider first a plant where a single product is made in a sequence of repeated batches using the same equipment, i.e., on a committed production line. For simplicity, the times to fill, empty, and

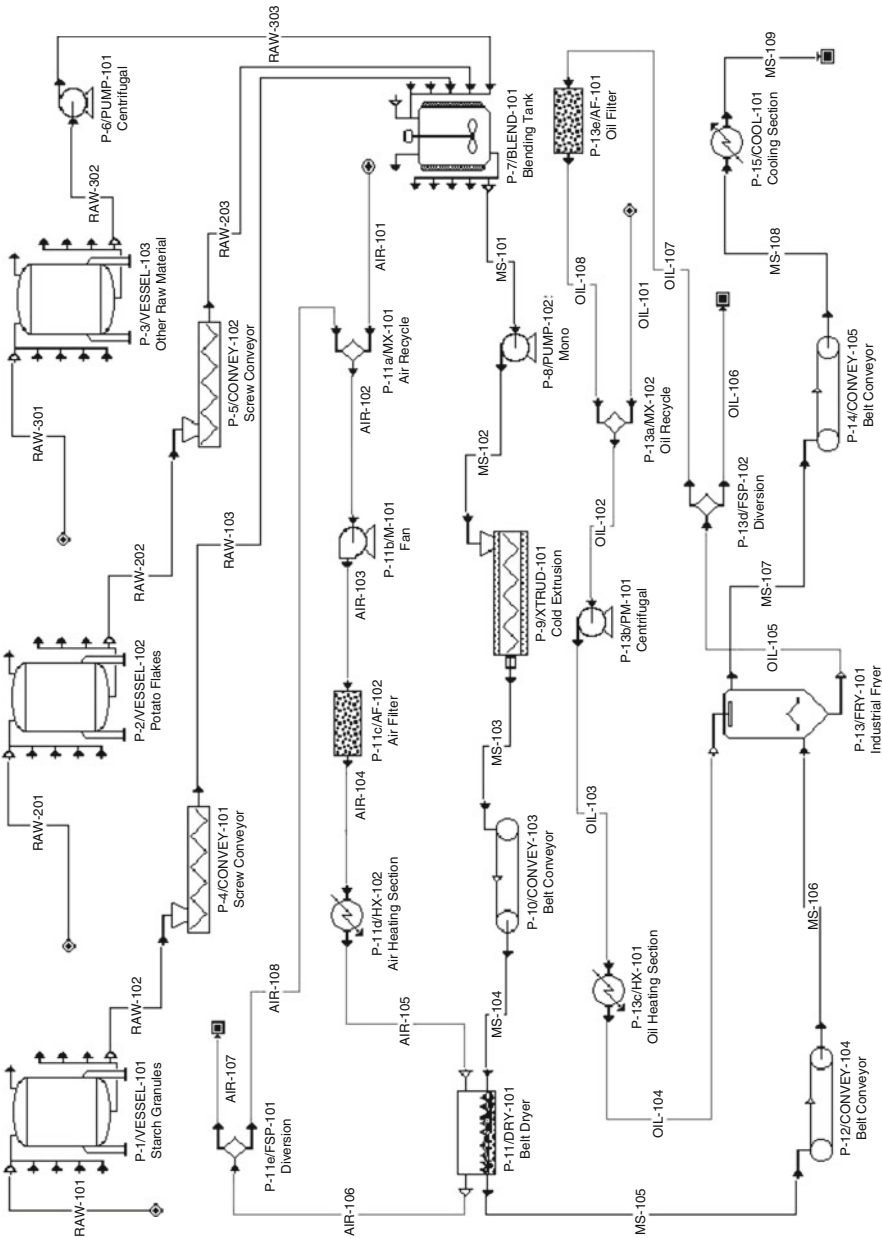
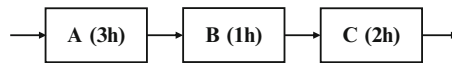


Fig. 2.4 Flowsheet of potato-frying process

Table 2.1 Selection of Web sites that advertize HACCP software

Software	Organization	URL
Food Safety Management System (FSMS)	EtQ, New York, NY, USA	http://www.etq.com/haccp
HACCPweb	One World Learning Ltd, Cork, Eire	http://www.haccpweb.com
HaccpHelp	HACCPHelp Software, Inc., Caledon, ON, Canada	http://haccphelp.com/haccphelp_software.htm
doHACCP	Norback, Ley & Associates, Middleton, WI, USA	http://www.norbackley.com
HACCP Control Point	Vertical Software International, London, ON, Canada	http://www.haccp.ca
HACCP Software	HACCP Builder, St. Paul, MN, USA	http://www.haccpbuilder.com
HACCP Now	HACCP Now, UK	http://www.haccpnow.co.uk
HACCP Software	HACCP Software, Dublin, Eire	http://www.Haccpsoftware.com

Last accessed 14 September 2011

**Fig. 2.5** Flowsheet of hypothetical 3-stage batch process

clean between operations are neglected. The flowsheet (Fig. 2.5) shows a hypothetical process with three batch stages A, B, and C lasting 3, 1, and 2 h respectively. Figure 2.6a shows the Gantt chart for nonoverlapping operation. The chart shows that none of the equipment is used continuously; batches (indicated by “P”) are produced every 6 h. Figure 2.6b shows how this plant is used with overlapping operation. Now Stage A is fully occupied and batches are produced every 3 h. Since Stage A is used continuously the only way of further speeding up the process would be to remove the bottleneck by duplicating this piece of equipment (which of course raises a cost issue). The result is shown in Fig. 2.6c. This can be speeded up further by duplicating Stage C (Fig. 2.6d). The minimum batch time and maximum equipment utilization could be achieved by using three Stage A units, one Stage B unit and two Stage C units.

The “rules” governing the batch cycle times are relatively easily seen from this example. For nonoverlapping operations, the batch cycle time is the sum of the individual stage batch times (i.e., $3 + 1 + 2 = 6$ h). For overlapping operations, with no duplicate equipment, the batch cycle time is the time needed to complete the slowest individual stage (i.e., here 3 h). With overlapping operation the (average) batch cycle time is the maximum value of the ratio of the batch time to the number of duplicates as calculated for each stage. Thus, referring to Fig. 2.6c—with Stage A duplicated—the batch cycle time = $\max \{3/2, 1, 2\} = 2$ h. When both Stages A and C are duplicated, the batch cycle time is reduced to $\max \{3/2, 1, 2/2\} = 1.5$ h, as may be confirmed from Fig. 2.6d, which shows two batches every 3 h. Note that the cycle time could be reduced to 1 h by using a third Stage A.

If the total daily production of this particular product is fixed, then the strategies outlined above have implications for the batch scale and therefore the size of the equipment. This is illustrated in Table 2.2 in which the four strategies are compared, assuming 24-h production per day, and a daily production of 1,000 kg product. The economic and operating implications of these choices must be explored at an early stage of the design process.

Consider now the typical situation where the factory manufactures more than one product. For similar products this is likely to imply the use of the same stages in the same sequence. For dissimilar products it is likely that different stages and/or sequences will be used. For these situations the cycle times can usually only be obtained by computation. To illustrate the problem, we use a simple

Fig. 2.6 3-Stage batch process: (a) Gantt chart—nonoverlapping operation, (b) Gantt chart—overlapping operation, (c) Gantt chart—overlapping operation, Stage A duplicated, (d) Gantt chart—overlapping operation, Stages A and C duplicated

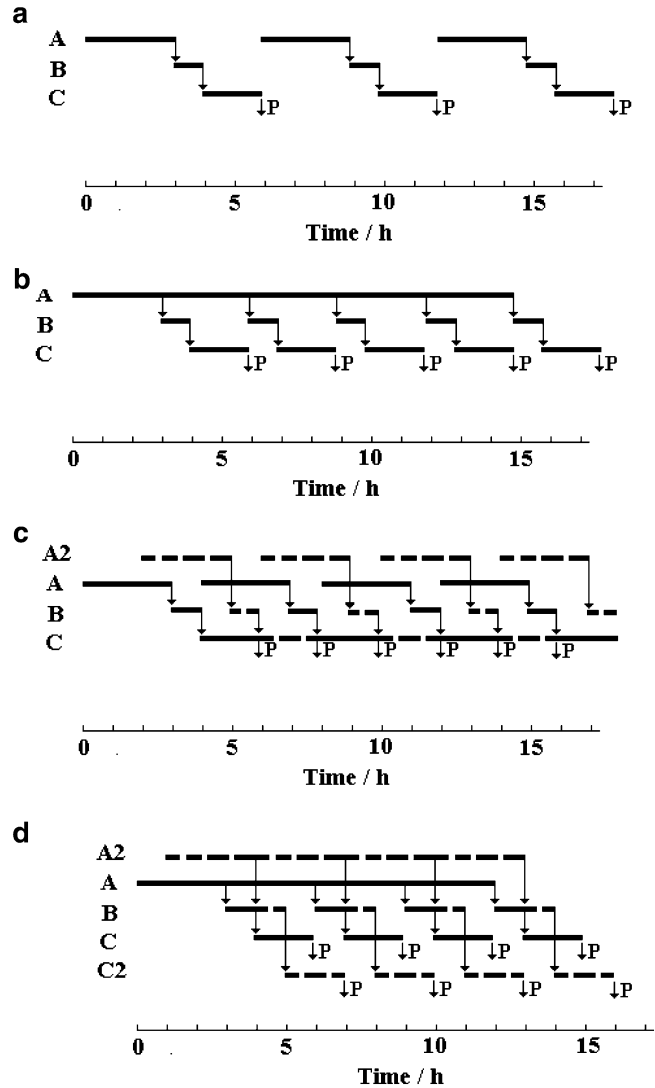


Table 2.2 Effect of schedule on batch equipment size

Strategy	Batch cycle time (h)	No. of batches/24 h	Batch size (kg)
Nonoverlapping (Fig. 2.6a)	6	4	250
Overlapping (Fig. 2.6b)	3	8	125
Duplicate stage A (Fig. 2.6c)	2	12	83
Duplicate stages A and C (Fig. 2.6d)	1.5	16	62.5

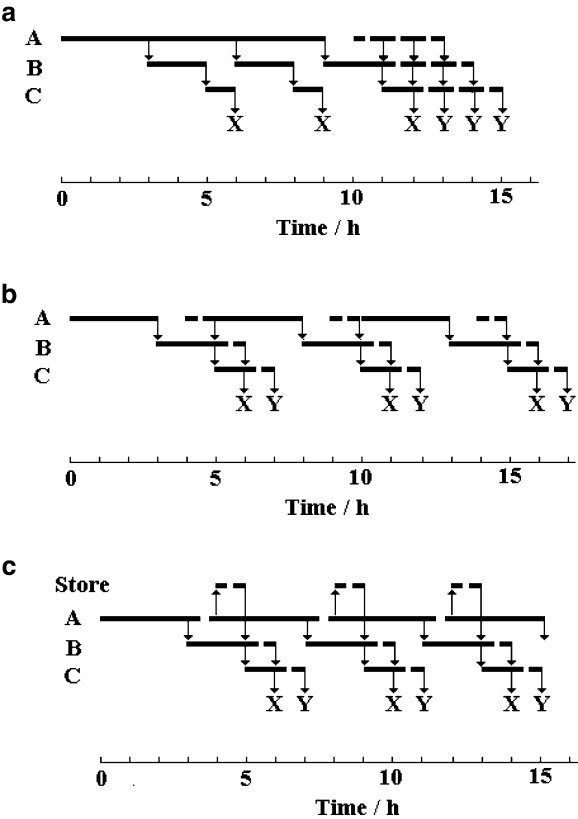
three-stage process manufacturing two products X and Y, using the same equipment but with different processing times, as summarized in Table 2.3.

It is immediately obvious that the choice has to be made between a single-product campaign (i.e., XXX...YYY...) or a multiproduct campaign (e.g., XYXY...) as shown in Fig. 2.7a, b, in which the clean up time between X and Y is neglected. Arbitrarily it is assumed that three batches of each product are to be made. In this example, the mixed-product campaign is less time-efficient than

Table 2.3 A multiproduct process

Product	Batch times (h)		
	Stage A	Stage B	Stage C
X	3	2	1
Y	1	1	1

Fig. 2.7 Manufacture of more than one product: (a) Gantt chart for single product campaign, (b) Gantt chart for multiproduct campaign, (c) Gantt chart for multiproduct campaign with intermediate storage



the single-product campaign. (The total production time for the single product campaign is 15 h; that for the multiproduct campaign is 17 h.) The situation is less favorable still if there is an additional clean up time between products *X* and *Y*. In other situations, however, the mixed-product campaign may be more resource efficient.

Suppose, however, that the mixed-product campaign (Fig. 2.7b) is preferred. What can be done to improve its efficiency? We have already seen the potential advantages of duplicating equipment that acts as a bottleneck. In fact there is no bottleneck in the scheme shown in Fig. 2.7b. At some stage in the production cycle *all* three stages are idle. This suggests that some intermediate storage should be considered (but in food processing, the opportunities will be constrained). For example, Fig. 2.7c shows the effect of introducing an intermediate storage tank for product *Y* after stage *A*. This improves the time efficiency for the multiproduct campaign to a level comparable with that of the single product campaign.

Intermediate storage can also be useful with single-product campaigns, since it can help to decouple the stages so that each can operate with its own cycle time and batch size. This is illustrated in Fig. 2.8. Here the flowsheet corresponding to Fig. 2.6b has been modified by introducing two intermediate storage vessels between Stages *A* and *B* and *B* and *C*. The storage tanks are sized such

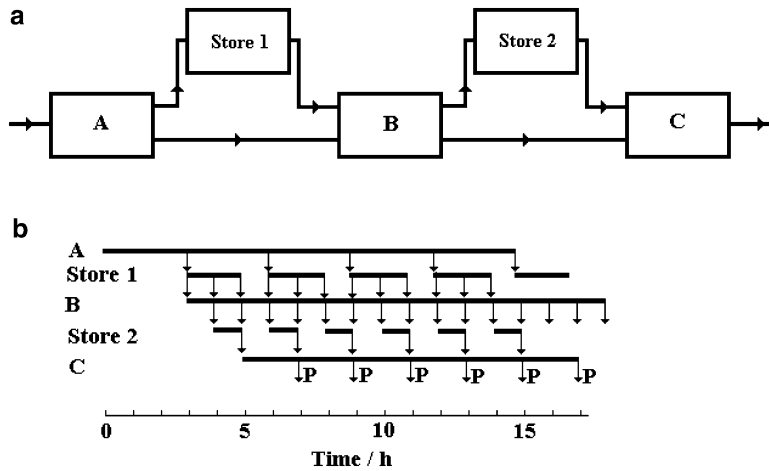


Fig. 2.8 3-Stage process with storage: (a) Flowsheet, (b) Gantt chart

that the first holds two-thirds of the batch discharged from Stage A, and the second holds one “batch” from Stage B. The Gantt chart shows that whilst there is no effect on the overall cycle time, idle time on Stages B and C has been eliminated. These stages can now be smaller than those specified in the original design. The economic and operating consequences should therefore be explored before the plant design is confirmed.

Today, computer-based methods for analyzing production sequencing are available and complex systems can be readily analyzed, although real-time optimization is still a target for the future.

2.3 Estimating Material and Energy Requirements

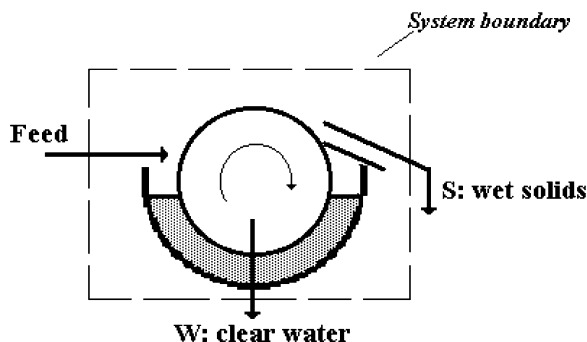
The outline flowsheet, the production targets and the operating conditions, such as the temperature of the cool milk fed to the storage tank or the temperature–time profile required for batch pasteurization, form the basis for preliminary estimation of the flows of materials and energy through the process. These estimates start from the assumption that mass and energy are conserved—i.e., they are neither created nor destroyed. The “form” they take may change profoundly during processing: raw materials are transformed into finished products; the energy used in mixing flour, water and fat helps to convert the mixture into dough with very different physical and biochemical characteristics from the raw materials, and so on. Some of the general principles underpinning material and energy balances are outlined first. The application of material balances and energy balances is then described.

2.3.1 Mass Balancing

2.3.1.1 Law of Conservation of Mass

Ultimately, it is the chemical elements that are conserved. Thus, whether or not reactions occur, one can, in principle, always establish an elemental balance. However, in food processing this is hardly ever convenient or possible. But the fact that the elements are conserved means that mass is also conserved; we shall develop this in more detail below. Moreover, in processes involving only liquid streams, volume conservation can often be assumed as a reasonably adequate first approximation.

Fig. 2.9 Rotary filter



2.3.1.2 Closed and Open Systems

If a mass or energy balance is to be set up, the statement that mass and energy are conserved is meaningless unless the system and its boundaries are defined. A system can be part of an operation, a whole operation (such as a mixer), a production line or a whole factory. Systems may be defined as Closed or Open. A *closed system* is one in which there is no material transfer across the defined boundaries—for example, a sealed can of peas being sterilized in a retort. In contrast, in an *open system*, material may be transferred in and out across the boundaries. A continuous fermenter or a food extruder are examples of open systems.

In the case of batch processes, the definition of the system will also involve time (e.g., the start and end of a process cycle).

Example 1. 100 kg of a waste stream containing 10 wt% suspended solids is separated in a vacuum filter to produce clear water and slightly wet solids containing 2 wt% moisture. How much of each stream is produced?

The total amount of feed to the process, 100 kg, is taken as the basis of the calculation. The choice of basis is important, and should always be stated. The boundaries of the system—the filter itself—are shown as a broken line in Fig. 2.9. This is an open system; the input and output streams (all of which must be accounted for in the balance) cut the system boundary.

Assuming that no material accumulates in the filter, conservation of mass implies that the total quantities of water and solids respectively into and out of the system remain constant. Note that the feed contains 90 kg water and 10 kg dry solids. The wet solids product stream contains 2 % water and 98 % dry solids. Then if S and W are the total masses of the wet solids and clarified water, balances on the total flows, dry solids and water respectively are:

$$100 = W + S. \quad (2.1)$$

$$10 = 0.98S. \quad (2.2)$$

$$90 = W + 0.02S. \quad (2.3)$$

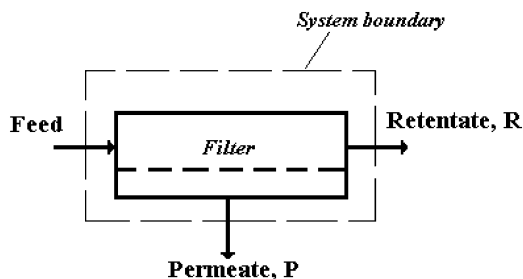
From (2.2), $S = 10.20$ kg, and from (2.1), $W = 89.80$ kg.

Note that although there are three equations and only two unknowns, there is no mathematical problem here, since only two (any two) of the equations are independent. Given balances on the two components (water and dry solids) the overall balance immediately follows. This can be checked by substituting the values for S and W into (2.3). Alternatively, the problem could be solved by solving either (2.2) and (2.3) or (2.1) and (2.3). Choose the simplest!

This method can easily be extended to processes with more unchanging components. The solution, by repeated substitution, is normally straightforward when the number of equations is small.

Table 2.4 Mass balance for Example 1

Component	Stream <i>F</i> /kg	Stream <i>S</i> /kg	Stream <i>W</i> /kg
Water	90	0.2041	89.7959
Solids	10	10	0
Total	100	10.2041	89.7959

Fig. 2.10 Cross-flow membrane filtration

The results can be represented in many ways. It is often convenient to present the data in tabular (e.g., spreadsheet) form, making checking straightforward, as shown in Table 2.4.

In this example, since there is no reaction between the components, the entries in each of the boxes in the second column (the inputs to the system) must balance the sum of the entries on the same row in the third and fourth columns (i.e., the sum of the output streams).

This simple example also provides insights for developing some important concepts and rules. The first concept is that of the basis for a calculation.

2.3.1.3 The Choice of Basis

Few real-world problems have such an obvious basis for the calculation as the example above. A more typical process-engineering design will start with a brief to produce a defined amount of product. For example, a company may want to produce 10,000 pots of yogurt. Even with a well-defined flowsheet, back-calculation (i.e., from the product) is extremely inconvenient. Calculation is normally easiest when the “flows” of information and mass are in the same direction. Thus, it is far more convenient to carry out the calculation on the basis of, say, 100 kg of milk (i.e., an input to the process), and then, finally, to adjust the numbers throughout by the appropriate ratio.

Two useful principles apply to the choice of basis:

- Choose a basis which is convenient for subsequent calculation and checking.
- Don't change the basis during the calculation.

Example 2. A cross-flow membrane (Fig. 2.10) is used to concentrate an aqueous solution containing 5 wt% whey protein. The retentate (product) stream *R* is to contain 30 % protein; the separation is not perfect and the permeate *P* will contain 2 wt% protein. It is desired to produce 80 kg of protein per day in the retentate. Calculate the flows and compositions of all three streams.

Basis: Whilst it would be possible to choose 80 kg of protein, i.e., 80/0.3 or 266.67 kg of retentate, as basis, it is simpler to base the calculation on the feed. So we take the basis as 100 kg feed solution.

As before, we set up an overall balance and balances on the components—protein and water—remembering that only two of these equations are independent:

$$\text{Overall : } 100 = R + P. \quad (2.4)$$

$$\text{Protein : } 5 = 0.3R + 0.02P. \quad (2.5)$$

$$\text{Water : } 95 = 0.7R + 0.98P \quad (2.6)$$

Solving any two of (2.4)–(2.6) gives:

$$P = 89.29 \text{ kg.}$$

$$R = 10.71 \text{ kg.}$$

Since the calculated retentate is not the design value (266.67 kg), all mass flows must be multiplied by 266.67/10.71, to give:

$$\text{Feed} = 2,489.0 \text{ kg.}$$

$$\text{Permeate} = 2,222.3 \text{ kg.}$$

$$\text{Retentate} = 266.7 \text{ kg.}$$

A complete check on both components is easily constructed, as in the previous example.

2.3.1.4 Losses

A key assumption in the above examples was that no material was lost during processing. In the real world, material is lost during cleaning, from spills, etc. The assumption of zero losses is a good starting point for a preliminary analysis, however. Historical data can be used to refine the calculations where necessary. Most important, calculations based on an assumption of “ideal” behavior provide an important reference point for analyzing plant performance in practice and for calculating the losses and wastage: the benefits of benchmarking for process efficiency should not be underestimated.

2.3.1.5 The Concept of Steady State

The examples above could refer either to batch or continuous operation. If the process is carried out continuously, the assumption that there is no change in the material inventory within the system implies that the system is steady—i.e., that it does not change with time.

For a steady non-reacting system, a balance on every component or the sum of the components (i.e., the total inputs and outputs) is:

$$\text{Mass in} = \text{Mass out.} \quad (2.7)$$

The balance can also be formulated in terms of rates of flow (e.g., expressed in kg/s) and, again, for every component and the total:

$$\text{Rate of mass flow in} = \text{Rate of mass flow out.} \quad (2.8)$$

For a reacting system, whilst balances on individual components will not be so simple (unless they are inert), the *overall* material balance must obey these equations.

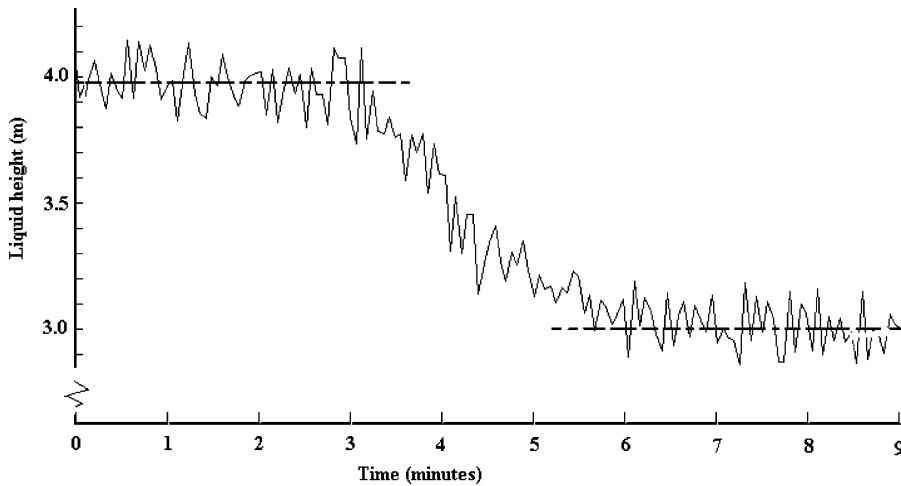


Fig. 2.11 Liquid level in process vessel

In practice, flowrates and compositions are never precisely constant, even though over a longer period the fluctuations average out. Instantaneously, then, the simple steady-state balance may not hold, as the system inventory or holdup changes to accommodate the fluctuations. Checking the mass balance from operating data on a process is never trivial. In a continuous process it is important to look at the time record of inputs and outputs. This is illustrated in Fig. 2.11, which shows the variations in liquid level in a continuously operating vessel. During the period shown, the system is moving from one steady state to another. In this example the measurements are “corrupted” (as in the real world) by noise. Figure 2.11 has three regions: over the first 3 min, the system fluctuates around a steady state, with an average liquid level just below 4 m. At any instant the system is not strictly steady, but over a period of a few minutes the system can be assumed, on average, to be steady. Between approximately 3 and 6 min, the plant is adjusting to a new set of conditions and is unsteady whether the timescale is of order seconds or minutes. After around 6 min, the level fluctuates around a new steady state value of 3 m. It is important, therefore, to be clear about the time scale over which the plant is analyzed and to differentiate between steady and unsteady operation.

If the system is such that changes in mass holdup are impossible (for example, with a constant-volume liquid mixer) then, even if one or more of the inputs changes with time, the system can be treated as if it were in steady state since the instantaneous flowrates in and out must balance. The simplest example is the flow of an incompressible fluid through a constant volume tank.

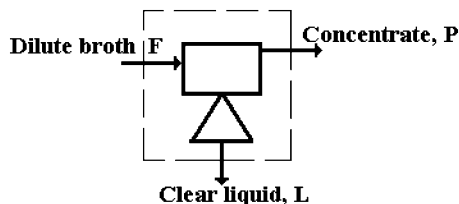
2.3.1.6 Batch Processes

Batch processes, such as a batch retort or a dough mixer, are, by definition, ones in which something changes with time—i.e., they are unsteady. However, in terms of overall process calculations, it can often be safely assumed that there is no net accumulation or loss within the system, provided the start and end of the batch cycle are properly specified.

2.3.1.7 Unsteady Systems

An unsteady system is one in which conditions within the system change with time. This might be a storage silo, a batch process, a warehouse, or, as in Fig. 2.11, during a change of operating conditions

Fig. 2.12 Centrifuge



in a continuous process. In this general case, the overall balance must account for the possibility of accumulation (or depletion). Taking as basis a fixed period of time or a defined quantity of a feed stream, the balance on any component, or their sum is, for a non-reacting system:

$$\text{Quantity accumulated} = \text{Quantity in} - \text{Quantity out} \quad (2.9)$$

or, instantaneously:

$$\text{Rate of accumulation} = \text{Rate of flow in} - \text{Rate of flow out.} \quad (2.10)$$

Example 3. A continuous cornflake production process includes a surge tank between the dryer and the toaster. The process is designed to operate steadily at a flowrate of 10 t/h. However, the feed flowrate may change by up to 2 t/h for periods of up to 30 min. What size of surge tank is necessary if the flowrate to the toaster is not to be interrupted during the operation?

Basis: Thirty minutes operation.

During a surge in inlet flow, the system is unsteady and the maximum total change in inventory during half an hour of increased or decreased inlet flow is ± 1 t. Thus, the surge tank would need to have a capacity of at least 2 t to cope with the foreseen surges, assuming that under normal steady operation the tank was run half full. This assumes that the probability of two successive surges in the same direction is very small. Is the answer realistic?

2.3.1.8 Inert Materials

Many processes involve inerts—i.e., components, which pass through the process unchanged. For example, atmospheric nitrogen can be treated as an inert in aerobic fermentations.

Sometimes an inert component, which enters in one stream and leaves in another, can be a convenient “hanger” on which to base—and simplify—a calculation. This is known as a “tie” substance. The solids in Example 1 are, in effect, a tie substance.

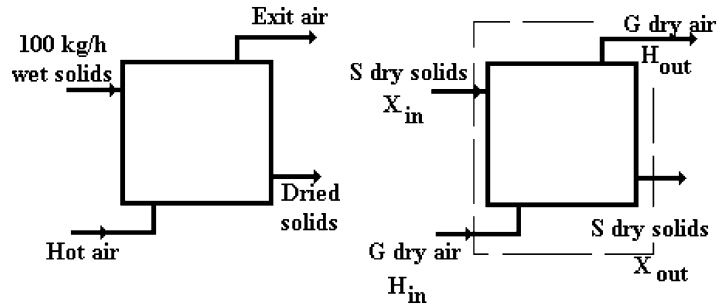
Example 4. A product stream F from a baker’s yeast fermenter contains 5 wt% yeast cells as suspended solids. The plant produces 100 kg of dry yeast/day. The broth is concentrated in a continuous centrifuge into a stream P containing 15 wt% suspended solids and clear liquid L , some or all of which could potentially be recycled. What is the maximum amount of clear liquid, which could be recycled?

Basis: 100 kg/day dry yeast (i.e., 1 day’s production).

The process is illustrated in Fig. 2.12. In this case the 100 kg of yeast can be treated as a tie substance. Assume that the broth comprises cells and (clear) liquid. An overall balance (in kg/day) gives:

$$F = P + L$$

Fig. 2.13 Continuous dryer



and a balance on the tie substance gives

$$100 = 0.05F = 0.15P,$$

whence:

$$F = 2,000 \text{ kg.}$$

$$P = 666.7 \text{ kg}$$

$$L = 1,333.3 \text{ kg.}$$

2.3.1.9 Wet and Dry Basis

The concept of a tie substance is also useful with processes such as humidification and air drying. The mass concentration of water vapor in air is usually defined in terms of humidity, i.e., the mass of water vapor per unit mass of dry air (*not* the mass of water vapor per kg of moist air). In a similar way, the moisture content of a wet solid can be defined on a *wet basis* (i.e., X_w kg water/kg wet material) or on a *dry basis* (i.e., X_d kg water/kg dry solids). It is easily shown that:

$$X_w = X_d / (1 + X_d) \quad (2.11)$$

and

$$X_d = X_w / (1 - X_w). \quad (2.12)$$

Suppose we have F kg moist solids with moisture content X_w (wet basis). The quantity of dry solids is $F(1 - X_w) = S$, say. The quantity of water is SX_d or FX_w .

Example 5. 100 kg/h of moist milk solids containing 10 wt% moisture (wet basis) is dried continuously to a final moisture content of 0.5 wt% using warm air of humidity 0.01 kg/kg (Fig. 2.13). The exit air humidity must not exceed 0.02 kg/kg dry air. Calculate the minimum air flowrate (dry basis) and the production rate of dried solids.

Basis: 100 kg/h moist solids.

The solution to this problem is simplified if it is recognized that the flows of dry solids and dry air (S and G kg/h respectively) do not change between the inlet and outlet of the dryer. We denote X_{in}

and X_{out} as the moisture contents of the inlet and outlet solids expressed in kg water/kg dry solids, and Y_{in} and Y_{out} as the humidities of the inlet and outlet air streams.

Now, since S and G are constant, a mass balance on water over the system gives:

$$S(X_{\text{in}} - X_{\text{out}}) = G(Y_{\text{out}} - Y_{\text{in}}) = M, \quad (2.13)$$

where M is the rate of transfer of water between the solids and the air.

From (2.12), the dry-basis moisture contents of the feed and product are:

$$X_{\text{in}} = 0.1/0.9 = 0.1111$$

and

$$X_{\text{out}} = 0.005/0.995 = 0.0050.$$

The dry solids flowrate is

$$S = 0.9 \times 100 = 90 \text{ kg/h.}$$

Therefore, the outlet flowrate of moist solids = $S(1 + X_{\text{out}}) = 90 \times (1.005025) = 90.45 \text{ kg/h}$

The moisture balance, (2.13), is thus:

$$90 \times (0.1111 - 0.0050) = G(0.02 - 0.01).$$

Hence

$$G = 954.8 \text{ kg dry air/h.}$$

2.3.1.10 Multistage Processes

Real processes rarely consist of a single piece of equipment—most production lines involve several interconnected units. Provided that all the input and output streams are identified correctly, the system boundary—and the mass and energy balances—need not be confined to a single operation. The system may be the whole process, or part of it. The system definition will depend on the objectives of the calculation and on the information available. To set up a mass balance over a complete process, it is often necessary to carry out detailed calculations over the individual units in the flowsheet.

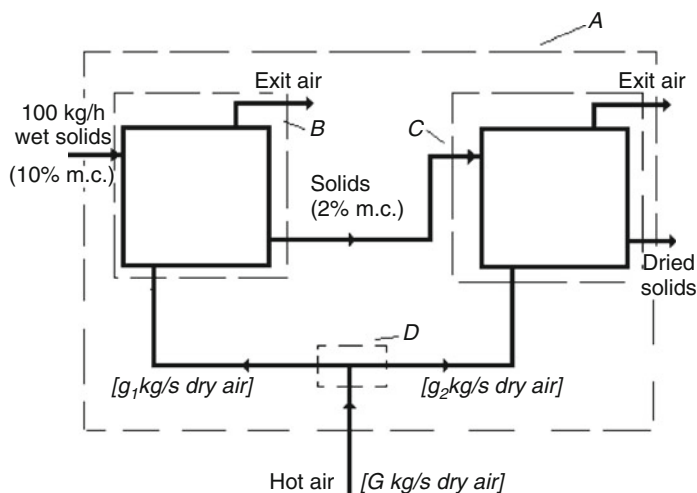
To illustrate some of the methods involved in handling multistage processes, consider a simple extension of the previous example.

Example 6. Suppose that instead of carrying out the drying operation in a single dryer, the operation is carried out instead in two linked dryers, see Fig. 2.14. The intermediate solids moisture content is 2 wt% on a wet basis. The final solids moisture content is 0.5 wt%, as before. Calculate the air flows, if the exit air humidity from each unit is again 0.02 kg/kg.

Basis: 100 kg wet solids (as before) (i.e., 90 kg dry solids).

It is possible to define four system boundaries, as shown in Fig. 2.14:

- (A) Around the whole process.
- (B) Around the first unit.
- (C) Around the second unit.
- (D) Around the air stream feed split.

Fig. 2.14 Two-stage dryer

B, C and D can be thought of as subsystems of A.

Note that the overall system is exactly the same as in the previous example. The input and output flows of solids and gas are therefore the same as above. Since the moisture balance is unchanged we can immediately conclude from a balance over the stream split (D) that

$$g_1 + g_2 = G = 954.8 \text{ kg dry air/h.} \quad (2.14)$$

In order to calculate g_1 and g_2 , it is now necessary to carry out balances over the sub-systems. Recall that, on a dry basis, the moisture content of the solids feed is 0.1111; the dry-basis moisture content of the intermediate solids stream is, from (2.12), 0.0204 on a fractional basis. A moisture balance over the first dryer, i.e., system boundary B, gives:

$$90 \times (0.1111 - 0.0204) = g_1(0.02 - 0.01)$$

so that

$$g_1 = 816.3 \text{ kg dry air/h.}$$

The air flow to the second dryer, g_2 , can be obtained directly from the overall (air) balance around the split (D), (2.14), giving $g_2 = 138.4 \text{ kg dry air/h.}$

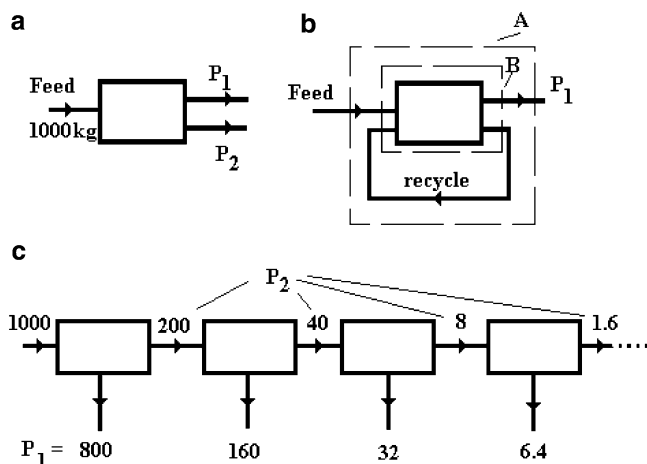
A moisture balance on the second drier is strictly redundant, but serves to check on the calculation:

$$g_2 = 90 \times (0.0204 - 0.0050)/0.01 = 138.4 \text{ kg dry air/h.}$$

The example confirms the obvious: provided the systems boundaries are correctly defined, the sum of the balances over the sub-systems must equal the balance over the whole system. This result applies equally to material and energy balances. Moreover, for a defined set of input/output conditions, the number (or indeed the type) of operations within the system boundary does not affect the overall balance. In the example above, as far as the overall balance is concerned, it is immaterial whether there are one, two, or more dryers.

In this example, the calculations were quite straightforward, since it was possible to complete the solution by marching forward from the inlet of the system. This may not always be possible when the plant is more complex or information is less well defined. As an example, we now consider systems with recycle.

Fig. 2.15 Recycle systems



2.3.1.11 Systems with Recycle

The ability to recycle material streams is very important in many processes. In general, the intention is to improve the process efficiency. For example, recycling may be used to recover and reuse raw materials that are not completely converted or to make better use of other process streams. Recycling of process streams and utilities is also a key element in strategies for waste minimization.

The principles can be explained by considering the hypothetical *continuous* food process shown in Fig. 2.15a where raw materials are transformed into the finished product P_1 and partially finished material P_2 . The process efficiency is 80 %, i.e., 80 % of the feed is converted into product. Thus, for 1,000 kg feed the process gives 800 kg of product P_1 and 200 kg P_2 .

Now suppose that all P_2 can be reworked into the product in the same machine, also with an efficiency of 80 %. Then a system with total recycle could be employed as shown in Fig. 2.15b.

The mathematics (i.e., mass balance) is deceptively simple. First we take a balance over the whole system, including the recycle loop, i.e., system A. Then

$$1,000 = P_1.$$

In other words, all the raw material is converted to product.

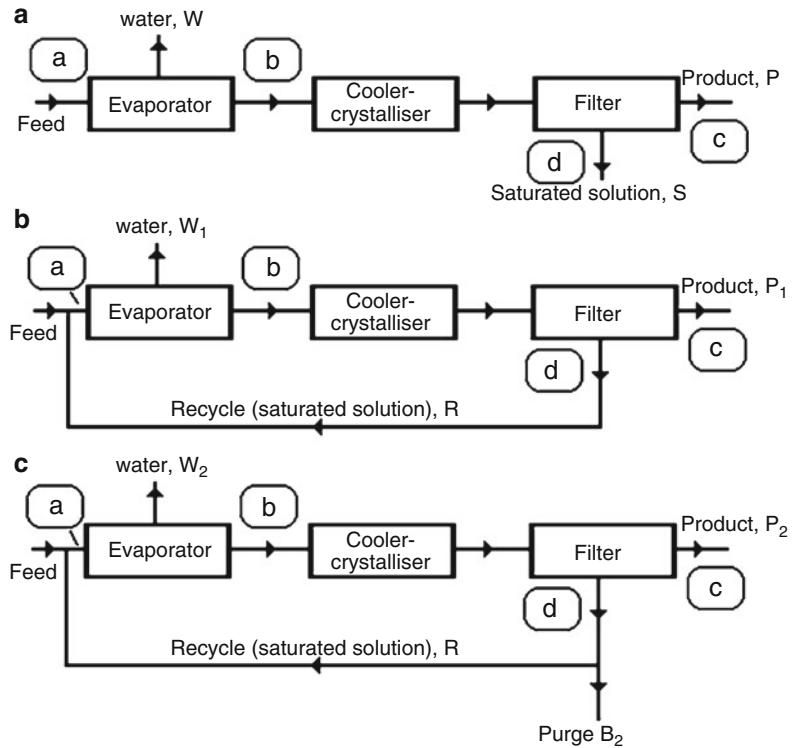
The question is then: how much material must be recycled per 1,000 kg feed? Call the recycle quantity R . Now assuming the same processing efficiency—i.e., 80 % of the total feed to the operation is converted to “useful” product—then a balance over system B gives:

$$P_1 = 0.8(1,000 + R) = 1,000. \quad (2.15)$$

So the recycle stream, $R = 250$ kg.

Note that the recycle stream is *not* the same as the “waste” stream in the system without recycle (i.e., 200 kg). This is because the system with recycle is *different* from the original system. In this analysis, it is assumed that the recycle is operating at steady state—it does not include the transient period during which steady state is reached.

One way of interpreting this particular result is to think of the process as one in which the by-product P_2 is successfully reworked into the product through an infinite number of reworking stages, each with an efficiency of 80 %, as shown in Fig. 2.15c.

Fig. 2.16 Citric acid production

$$\text{The total amount of reworked product } (P_2) = 200 + 40 + 8 + 1.6 + \dots \quad (2.16a)$$

and

$$\text{The total amount of product } (P_1) = 800 + 160 + 32 + 6.4 + \dots \quad (2.16b)$$

In fact, both these series converge: the total reworked product

$$= 200(1 + 0.2 + 0.22 + 0.24 + \dots) = \frac{200}{1 - 0.2} = 250 \text{ kg} \quad (2.16c)$$

and, from (2.16b), the total product $P_1 = 1,000$ kg.

There are limits to the amount of reworking that can be achieved in practice. One reason is that the efficiency of conversion falls with increased reworking; another is the possible buildup of unwanted or even dangerous components. This is why recycling in agricultural and food systems must always be pursued with a good deal of caution.

Example 7. To recover citric acid from a fermentation broth, the acid is evaporated from a concentration of around 10–40 wt%, followed by crystallization at a lower temperature where the saturation concentration is around 2 wt% citric acid, and subsequent filtration. This is shown in Fig. 2.16a. What recycle options are available to increase the yield?

The slurry of citric acid crystals and saturated solution is separated by filtration. Since the latter stream still contains some citric acid, further recovery by recycling is clearly a possibility. Indeed, if there were no components other than citric acid and water in this stream, total recycle could be employed (Fig. 2.16b). However, if the stream contained another nonvolatile component, total

recycle could not be employed since this component would build up in the system until it contaminated the product itself. In this case, either some means of separating the contaminant or of controlling its buildup in the system must be found. The easiest way of limiting the buildup is by adding a controlled purge or bleed to the recycle, as shown in Fig. 2.16c.

We now establish mass balances over these flowsheets and compare the consequences for process efficiency.

Consider a plant where the feed rate to the process is 5,000 kg/day. The mass fraction of citric acid in the aqueous feed is 0.1; the mass fraction in the stream leaving the evaporator (i.e., at “b”) = 0.4 and the mass fraction in the saturated solution leaving the filter is 0.02. We also assume that the citric acid crystals in the product stream are pure and bone dry.

Take as basis one day’s operation, i.e., $F = 5,000$ kg.

1. System Without Recycle (Fig. 2.16a):

Since the compositions of the streams at “a” and “b” are known, we first set up a mass balance over the evaporator. Let B = total flow of the stream at “b”:

Given the data on flows and compositions, we can write three balance equations over the evaporator. Since there are two components (citric acid and water), only two are independent:

$$\text{Water : } 4,500 = W + 0.6B. \quad (2.17)$$

$$\text{Citric acid : } 500 = 0.4B. \quad (2.18)$$

$$\text{Total } 5,000 = W + B. \quad (2.19)$$

Solving any two from these gives

$$B(\text{stream to crystallizer}) = 1,250 \text{ kg.}$$

$$W(\text{water evaporated}) = 3,750 \text{ kg.}$$

Having solved for stream B , a mass balance over the cooler-crystallizer and the filter together can be established. Here we use the fact that stream S contains 2 % citric acid and 98 % water. These are, for each component and the total flows, respectively:

$$\text{Citric acid : } 0.4B = 500 = P + 0.02S. \quad (2.20)$$

$$\text{Water : } 0.6B = 750 = 0.98S. \quad (2.21)$$

$$\text{Total } B = 1,250 = P + S. \quad (2.22)$$

From which:

$$S = 765.3 \text{ kg.}$$

$$P = 484.7 \text{ kg.}$$

The overall yield of citric acid is therefore $100 \times 484.694/500 = 96.94 \%$.

2. System With Total Recycle (Fig. 2.16b):

The problem here is that the recycle flow, R , is not known. The values of W_1 and P_1 are also different from their values in the first example. However, they can be obtained directly from an overall mass balance.

Overall balances on the two components are, respectively:

$$\text{Citric acid : } 500 = P_1. \quad (2.23)$$

$$\text{Water : } 4,500 = W_1. \quad (2.24)$$

These clearly also satisfy the total balance.

We can immediately conclude that it is possible, using total recycle, to recover 100 % of the citric acid. Note that the price paid for this includes larger and more complex equipment, and increased energy use, since all the water in the feed is now evaporated, compared with the 3,750 kg evaporated in the system without recycle.

In order to estimate the recycle flowrate, it is necessary to carry out one or more internal balances. In general there are two ways of proceeding. The first, which is employed in many computer-aided design packages, is to use a trial and error or “tearing” algorithm. The second, which is sometimes possible with relatively simple problems, is to find an algebraic solution. Both methods are illustrated here.

Trial and Error: “Tearing”

In this example, although the composition of the recycle stream is known (it is saturated with citric acid, and therefore contains 2 % acid and 98 % water), its magnitude is not known. A process of trial and error, based on an initial guess for the value of R is used to converge on the true value. It will clearly be different from the value of the waste stream S in the first example, but presumably it should be of a similar order of magnitude.

We therefore assume a first estimate for $R = 750$ kg. Given this estimate, and knowing its composition we can calculate the flow and composition of the stream at “a,” the inlet to the evaporator. Balances on citric acid and water give the composition of stream “a” shown in Table 2.5a.

Since 4,500 kg of water is evaporated, the composition of stream “b” is as given in Table 2.5b. However, stream “b” should contain 40 % citric acid, rather than 41.2 %. This means that our estimate of R was wrong. As a second estimate try $R = 800$ kg. Proceeding as before, we obtain the corresponding results shown in Table 2.5c, d; these indicate that the second estimate for R is slightly too high. Linear interpolation suggests that the next estimate should be

$$R = 800 - (40 - 39.69) \times 50 / (41.2 - 39.69) = 789.7 \text{ kg.}$$

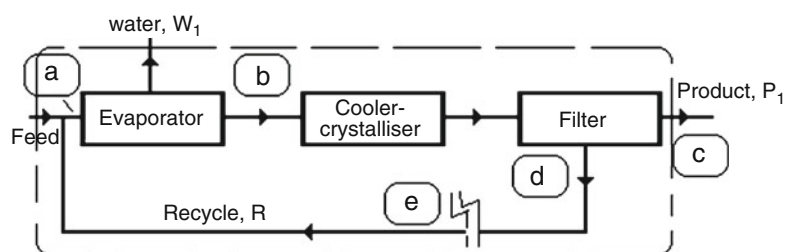
With this value we obtain the results shown in Table 2.5e, f. The calculated value of the citric acid content of stream “b” is now acceptably close to the target value.

This procedure is called tearing, since it can be interpreted as “tearing” one of the streams, estimating its value and then computing around the flowsheet to see if the values correspond. Here, as shown in Fig. 2.17, the calculation starts at point “e” and is carried through to “d,” at which point in the flowsheet the values should coincide.

In this case, it was not necessary to complete the calculation in order to arrive at a good estimate for R . However, for good measure, the whole procedure with $R = 789.7$ kg/h is summarized in Table 2.5g. This shows that the contents of streams “e” and “d,” on either side of the torn stream, do coincide. Note that this type of calculation is easily performed using for example the Solver routine in MS Excel.

Table 2.5 Mass balances for Example 7(a)

Component	kg	wt%		
(a) ($R = 750$ kg)				
Citric acid	$500 + 0.02 \times 750 = 515$	8.957		
Water	$4,500 + 0.98 \times 750 = 5,235$	91.043		
Total	$F + R = 5,750$	100		
(b)				
Citric acid	515	41.2		
Water	735	58.8		
Total	1,250	100		
(c) ($R = 800$ kg)				
Citric acid	$500 + 0.02 \times 800 = 516$	8.897		
Water	$4,500 + 0.98 \times 800 = 5,284$	91.103		
Total	$F + R = 5,800$	100		
(d)				
Citric acid	516	39.69		
Water	784	60.31		
Total	1,300	100		
(e) ($R = 789.7$ kg)				
Citric acid	$500 + 0.02 \times 789.7 = 515.79$	8.909		
Water	$4,500 + 0.98 \times 789.7 = 5,273.91$	91.091		
Total	$F + R = 5,789.7$	100		
(f)				
Citric acid	515.79	39.999		
Water	773.91	60.001		
Total	1,289.7	100		
(g) (Summary)				
Component	Stream at “e”	Stream at “a”	Stream at “b”	Stream at “d”
Citric acid	15.79	515.79	515.79	15.79
Water	773.91	5,273.91	773.91	773.91
Total	789.7	5,789.7	1,289.7	789.7

Fig. 2.17 Citric acid production: "tearing"

Analytical Solutions

In some relatively simple problems with recycle, particularly when the number of components is small, it is possible to solve problems like the one above directly. Instead of guessing the value of the recycle flowrate, R , and iterating until the solution converges, the problem is converted into algebraic form in order to solve for R . As above, the solution starts at point "e". Since the composition is known at this point (2 % citric acid and 98 % water) the second column in Table 2.6 can be

Table 2.6 Mass balances for Example 7(b)

Component	Stream at “e”	Stream at “a”	Stream at “b”
Citric acid	$0.02R$	$500 + 0.02R$	$500 + 0.02R$
Water	$0.98R$	$4,500 + 0.98R$	$0.98R$
Total	R	$5,000 + R$	$500 + R$

completed directly. The entries in the third column (at “a”) then follow directly from a mass balance over the mixing point, and the entries in the fourth column—the stream leaving the evaporator—follow from a mass balance over this unit. Since the citric acid must comprise 40 % of stream “b,” it follows that

$$500 + 0.02R = 0.4(500 + R) = 200 + 0.4R. \quad (2.25)$$

This yields:

$$R = 300/0.38 = 789.5 \text{ kg.}$$

In this example, the feed stream is unlikely to comprise only citric acid and water. Suppose this stream contains 1 % citric acid, 0.5 % contaminants and 98.5 % water. An immediate consequence is that operation with total recycle is no longer possible, since the contaminant would build up continuously in the system. In practice there will usually be some constraint on the permitted buildup of contaminant. Let us suppose that it should not exceed 5 % in the recycle stream. The strategy to control this level will be to introduce a controlled purge or bleed stream from the system.

We assume that no contaminant leaves with evaporated water or with the citric acid product. The obvious stream to purge is the saturated citric acid stream leaving the filter, i.e., stream “d”. The purge system is shown in Fig. 2.16c.

The introduction of the purge stream changes the overall balances to

$$\text{Citric acid : } 500 = P_2 + 0.02B_2. \quad (2.26)$$

$$\text{Contaminant : } 25 = 0.05B_2. \quad (2.27)$$

$$\text{Water : } 4,475 = W_2 + 0.93B_2. \quad (2.28)$$

$$\text{Total } 5,000 = P_2 + B_2 + W_2. \quad (2.29)$$

Note that, having fixed the level of the contaminant in the recycle and purge, the purge rate is fixed since, from (2.27), $B_2 = 500$ kg. The complete solution to the equation set above is:

$$B_2 = 500 \text{ kg.}$$

$$P_2 = 490 \text{ kg.}$$

$$W_2 = 4,010 \text{ kg.}$$

The purge stream and the recycle have the same composition, viz. 2 % citric acid, 5 % contaminant, 93 % water.

Proceeding as before, mass balances on the mix between the fresh feed and the recycle and then over the evaporator (where 4,010 kg of water are evaporated) give the results shown in Table 2.5b. The citric acid concentration in the exit stream from the evaporator is 40 % so that

$$500 + 0.02R = 0.4(990 + R)$$

or

$$R = 273.7 \text{ kg.}$$

The complete mass balance around the flowsheet is summarized in Table 2.6 (all numbers in kg), where the final column serves as a check on the second.

It should be clear from this example that, since the level of contaminant directly controls the purge rate, this also therefore influences the recovery of pure product (in stream P). There is always a trade-off between the level of recycle and the overall process efficiency.

2.3.2 Energy Balancing

The economic and environmental costs of energy use in the food processing industry are extremely important. The carbon tax levy puts even greater pressure on the industry to minimize energy use and to optimize its use over time across the whole operation. To begin to address this problem it is necessary to be able to estimate the heating or cooling requirements for typical operations such as thermal processing, cooling and freezing, mixing, etc. operations. This, together with information on the scheduling of operations, provides the basis for addressing problems such as: “Is a particular method of heating or cooling the best option?” or “How can the overall energy efficiency of the process be maximized?” To answer the first sort of question, we rely principally on the first law of thermodynamics: the statement that energy is conserved. To answer the second, we often need to draw on the second law. Here we will concentrate mainly on the first type of question, whilst introducing some approaches to the second.

First we must clarify what we mean by energy and how it is measured. In the SI system, the unit of force (dimensions: mass times acceleration) is the Newton (N) defined as 1 kg m/s^2 . The most familiar definition of energy is that in a mechanical system: it is the force applied times the distance moved. Energy or work thus has units of Newton meters or joules (J) where $1 \text{ J} = 1 \text{ N m}$. This applies to any form of energy. For example, an apple of mass 100 g held stationary 10 m above the ground has a potential energy of 9.81 J. Note that it is necessary to define a datum level—the ground. If some other reference or datum level had been chosen, the energy of the object would of course be different. Power, which is defined as the *rate of doing work*, is measured in watts where $1 \text{ W} = 1 \text{ J/s}$.

Of course, materials have energy because of other attributes: their velocity, temperature, pressure, physical state (solid, liquid, or gas), and chemical composition all contribute to their energy (or ability to do work). It is always important to define datum levels unambiguously; energy is always a function of state and calculations should always use the same reference level (i.e., where the energy level is arbitrarily taken to be zero). It is also important to work in consistent units: then the question of the energy source is irrelevant in the energy balance (but not its efficiency or cost).

Whilst energy is conserved, its value changes because it is transferred or converted from one form to another. In mixing, for example, the shaft power is converted into motion and, ultimately, through friction and viscosity, into heat. The second law states effectively that heat can only be transferred down a temperature gradient. Thus, the heat carried away from a process vessel by cooling water is always at a lower temperature—and therefore is less valuable in its capacity to do useful work—than the vessel contents. In other words, the quality of heat energy depends on its temperature and this must be accounted for in any analysis of energy efficiency. It requires ingenuity and money to upgrade its quality. Also, keep in mind that in the real world many energy

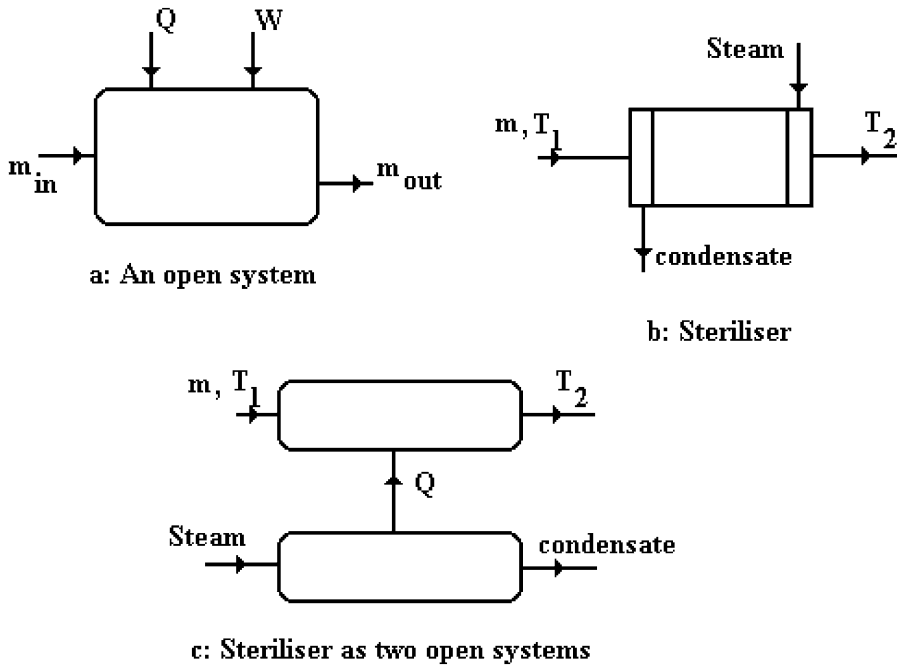


Fig. 2.18 Open systems

transformation processes are irreversible: heat released from the mixer cannot be used to unmix the dough.

In order to apply the first (and second) law to any real situation, it is important to have a clear definition of the system over which the balance is made. We must also differentiate between *closed* systems and *open* systems. A batch dough mixer is an example of a *closed* system: the only transfers of energy across its boundary during operation are in the nonmaterial forms of heat and shaft power. On the other hand, a continuous steam sterilizer is an open system since energy is transported into and out of the process, and used at the same time, by the process liquid and the steam and condensate (Fig. 2.18).

For a closed system, application of the first law is very simple. Consider a process operated over a given time interval in which the total *net* flows of heat and work are Q and W . (If a process picks up 5,000 kJ of heat from a steam supply and at the same time loses 500 kJ to the environment, the net flow of heat is 4,500 kJ). We define both Q and W to be positive for flows *into* the system.

Then the increase (decrease) in energy stored in the system must, because of conservation, equal the net transfer of energy into (out of) the system, or:

$$\Delta E = Q + W. \quad (2.30)$$

This equation will be in joules or kilojoules.

If ΔE is not zero then the final system contents must differ from the starting contents in some way, e.g., by changes in composition, temperature, physical state or internal surface area. For example, if an operation simply involved stirring and heating a liquid, the increase in energy would be entirely accounted for by an increase in temperature. In the case of a dough maker, however, the energy transferred through mixing would be used to drive chemical and physical changes as well as increasing the temperature. Clearly, (2.30) is only really useful when ΔE can be related to these physical and chemical changes.

Equation (2.30) can also be written in terms of the *rates* of energy change and transfer so that instantaneously

$$\frac{dE}{dt} = q + W, \quad (2.31)$$

where q and W are the net *rates* of energy and work input into the system. Equation (2.31) is thus a power balance, normally having the units of watts or kilowatts.

In the more general case of an open system we must allow for the energy transported in to and out of the system by the material flows, together with the “flow” work they generate. To set up equations similar to those for the closed system, we use the symbol M to represent the mass transferred by a particular stream of material; Q and W represent the total (net) quantities of heat and work added to the system over a period of time; alternatively, m represents the mass flowrate; q and w are the net rates of heat and work input (i.e., power as heat and mechanical energy).

Since energy cannot disappear, the *net* change in stored energy within the system boundary over a given period of operation must exactly balance the net energy input by heat and work and the net energy difference between all the inflowing and outflowing streams. For a process involving only one stream with energy/kg E_{in} and E_{out} respectively at the entry and exit from the system (where the energy levels are defined with reference to consistent datum levels), the energy balance is:

$$\Delta E = Q + W + (ME)_{\text{in}} - (ME)_{\text{out}}, \quad (2.32)$$

where ΔE is the *change* in stored system energy over the operating period in question. Alternatively, in terms of rates:

$$\frac{dE}{dt} = q + w + (mE)_{\text{in}} - (mE)_{\text{out}}. \quad (2.33)$$

These equations are easily generalized to systems with many input and output streams by adding the appropriate “ mE ” or “ ME ” terms.

These equations represent the balances over an unsteady open system—unsteady because they include the possibility of changes in the stored energy of the system with time. However, once the system *has achieved a steady state* in which all flows and temperatures are independent of time, the stored energy is then constant and the energy balance equations become respectively:

$$Q + W + (ME)_{\text{in}} - (ME)_{\text{out}} = 0 \quad (2.34)$$

and,

$$q + w + (mE)_{\text{in}} - (mE)_{\text{out}} = 0, \quad (2.35)$$

which are the general equations for a *steady open* system.

2.3.3 Stored and Internal Energy: Enthalpy

2.3.3.1 Single Components

As noted earlier, the energy of a process stream includes contributions from a variety of sources such as kinetic and potential energies, surface energy, the temperature, composition, physical conditions and state of the stream, etc. In any particular situation, the changes in many of these terms are

negligible: for example in situations involving only a single phase, interfacial forces can be forgotten. In heat exchangers, the changes in kinetic and potential energies are negligible, and so on. It is important to remember that all the contributions must be measured with respect to a defined but usually arbitrary datum.

In many applications, we use the *enthalpy*, or more correctly specific enthalpy, of a component or mixture of components as the measure of its energy content (the term *specific* is used to denote “per unit mass”). Enthalpy is the sum of:

- Internal energy U (reflecting temperature, physical state, etc., in relation to a defined datum and measured in J/kg or kJ/kg).

and

- A “flow work” term, pV (where p is pressure and V specific volume, which is the reciprocal of the density ρ), also in J/kg or kJ/kg.

Thus

$$h = U + pV = U + p/\rho. \quad (2.36)$$

The heat content of m kg of a material with specific enthalpy h kJ/kg is simply $m \cdot h$ kJ. Values of h for pure components are easily calculated or can be found from tables. For example, the steam tables tabulate the enthalpies of water (liquid and vapor) over a wide range of conditions relative to water at its triple point (essentially 0 °C). Such tables can be found in for example Green and Perry (2008) and also in most thermodynamics texts. The thermodynamic properties of food materials are less complete, but there are several good sources (see, e.g., Sahin and Sumnu 2006; Rahman 2009). The rules for mixtures (see Sect. 2.3.3.2) are often useful in food calculations.

Note also that for liquids and solids under “normal” processing conditions, the enthalpy and internal energy values are almost the same, and enthalpy values can be used in rough calculations without serious error. This is *not* true for gases and vapors where compressibility implies that the enthalpy is always measurably greater than the internal energy.

For a steady *open* system, the integrated form of the first law becomes

$$Q + W + (Mh)_{\text{in}} - (Mh)_{\text{out}} = 0 \quad (2.37)$$

and for a *closed* system:

$$Q + W = M[U_{\text{final}} - U_{\text{start}}], \quad (2.38)$$

where M is the total mass inventory of the system. Thus, for solids and liquids we can write:

$$Q + W = M[h_{\text{final}} - h_{\text{start}}]. \quad (2.39)$$

Often, changes in enthalpy are mainly due to changes in heat content, in particular, from a change in temperature (so called “sensible” heat) and/or from a phase change which, although isothermal, involves a “latent” heat. Thus, the enthalpy of a pure component at T_2 with reference to (itself in the same physical state at) T_1 is:

$$h_{21} = \int_{T_1}^{T_2} C_p dT = C_{p_m}(T_2 - T_1), \quad (2.40)$$

where C_{p_m} is the mean heat capacity between T_1 and T_2 .

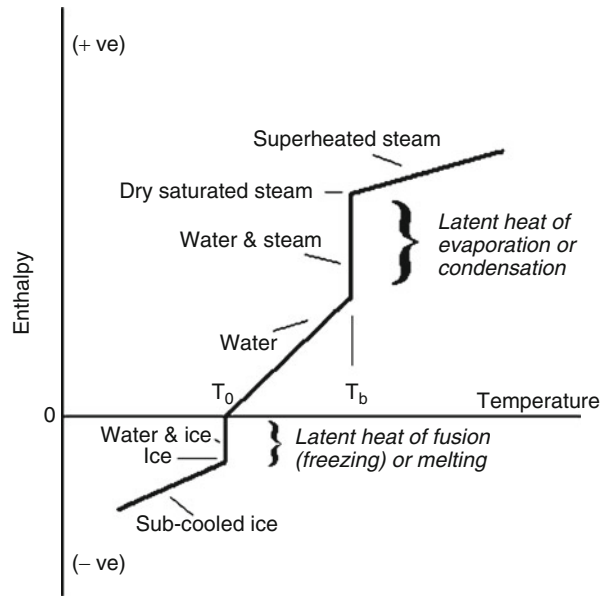


Fig. 2.19 Enthalpy–temperature diagram

For example, liquid water has a mean heat capacity of 4.19 kJ/kg K over the range 0–100 °C; taking its enthalpy to be zero at 0 °C, the enthalpy of water at 80 °C is thus approximately $80 \times 4.19 = 335.2$ kJ/kg (this is within 1 % of the accurate value given in steam tables). The energy to heat 100 kg of water from 25 to 80 °C is, from the first law (2.37):

$$Q = 100 \times 4.19 \times (80 - 25) = 23,045 \text{ kJ.}$$

(Note that this assumes that the mean heat capacity is constant; it isn't but the errors involved here are small).

A phase change at a given temperature and pressure requires or liberates a quantum of energy—the (specific) latent heat. For example, 334 kJ heat must be removed to freeze 1 kg water at 0 °C; to boil water at 100 °C requires 2,257 kJ/kg of water evaporated. Thus, the enthalpy of ice at 0 °C is lower than the enthalpy of water at the same temperature by the latent heat of fusion (i.e., freezing); the enthalpy of steam is greater than the enthalpy of liquid water at its boiling point by the latent heat of evaporation. It is interesting to note that it should require less energy to concentrate a liquid such as fruit juice by freezing rather than evaporation. Whilst true, practicalities intervene—it is more complex to design efficient systems operating below ambient temperatures and carryover of fruit into the ice phase is also a problem.

For a single, pure component the above discussion can be generalized. We take the enthalpy datum to be T_1 , liquid phase. Suppose the component evaporates at T_2 and is then superheated to T_3 . We also assume the mean heat capacities of the liquid and vapor over the temperature ranges in question to be C_f and C_g respectively; then the enthalpy at T_3 , h_{31} , is:

$$h_{31} = C_f(T_2 - T_1) + h_{fg,2} + C_g(T_3 - T_2). \quad (2.41)$$

The steam tables record values of h_f , h_{fg} , and h_g for dry, saturated and superheated steam. Note that h_g is defined relative to the triple point. Figure 2.19, a qualitative enthalpy–temperature diagram for water (at constant pressure), illustrates the principles involved in determining the enthalpy of a

pure substance as it goes through phase and temperature changes. The datum level is taken as the freezing point T_o . The boiling point is T_b .

Because energy and enthalpy are state functions, the final enthalpy in (2.41) is independent of the path between the reference at T_1 and the final state at T_3 ; in other words the temperature T_2 need not correspond to the actual temperature at which the phase change occurs.

To a very rough approximation, the enthalpy of many dilute liquid streams can be assumed to be close to that of water at the same temperature. This assumption is reasonable during the early stages of design; it should not be made during detailed calculations.

2.3.3.2 Mixtures

In practice most process streams comprise mixtures of components, whether dissolved or suspended. Sometimes these mixtures are far from ideal, so that enthalpy and other data must be gathered from the literature or measured directly. Often, a reasonable first approximation in food and bioprocessing is to assume ideal behavior, i.e., that *the enthalpy of a mixture is the weighted sum of the specific enthalpies of the various components*. This assumption breaks down where there are strong solutions, etc. Three effects in particular can invalidate the assumption of ideality: heat effects due to dissolution, mixing and dissociation. In many cases it is reasonable and justifiable for the purposes of establishing an energy balance to treat a complex stream as if were a single component with empirically determined physical and thermodynamic properties.

2.3.3.3 Biochemical Reactions

So far this discussion has been concerned with essentially physical changes. There are many situations where chemical or biochemical changes occur. We consider briefly how the treatment may be extended to cover such eventualities.

Consider an energy balance about a bread-baking line. We start with the dough mixer. Suppose the datum level is taken as the raw material ingredients (flour, fat, water, etc.) at 0 °C. The materials fed to the dough mixer will then have positive enthalpy values because their temperatures are generally greater than the datum. At the end of the dough-making cycle, the dough also has a positive “sensible” enthalpy by virtue of its temperature, but its enthalpy must also reflect the fact the dough is not simply an ideal mixture of the raw ingredients. In practice some of the energy input into the mixer goes into the creation of chemical bonds, and this is reflected in the enthalpy of the dough.

In principle, food and biological operations can be handled in just the same way as chemical reactions for which the necessary thermodynamics is well established. In practice, just as with mixtures, many unknown transformations may occur, in which case the process can be handled as a pseudo-reaction (which is useful in fermentation processes); alternatively it may be possible to identify the most important or limiting processes to help quantification. Many food operations are dominated by work (mechanical power) and thermal effects: the thermodynamics of the “reactions” involved in bread making, for example, are totally swamped by heat effects.

To set up the energy balance to include the effects of reactions and mixing (where these generate or require heat), it is best to choose the input raw materials as the datum level. Thus, to continue the qualitative discussion of dough mixing:

- The enthalpy of the input stream(s) is the sum of the sensible enthalpies of all the feed materials—i.e., allowing for temperature and possibly phase changes from the standard state taken as datum.
- The enthalpy of the output stream(s) has two additive components: the sensible enthalpy, which measures the enthalpy of the output materials with respect to themselves at 25 °C, and the

enthalpy of the output compounds at 25 °C with respect to the feed materials also at 25 °C. This second term would include the standard heats of reaction and mixing if these are significant.

For a single reaction in which x kg of product are formed and ΔH is the heat of reaction and/or mixing per kg product, the overall energy balance has the form:

$$\sum (mh)_{\text{inputs}} + q + w = x\Delta H + \sum (mh)_{\text{outputs}}, \quad (2.42)$$

where the enthalpies are measured/calculated from the same compound at 25 °C as the datum (i.e., the h terms are the “sensible” enthalpies).

Since this equation is written in terms of flowrates, q and w will be in kW; the balance can of course be written in terms of quantities in which case Q and W will be in energy units.

Note that the equation reduces to the earlier balances for non-reacting systems when the “ $x\Delta H$ ” term is negligible. Note, too, that in many food processes—dough mixing, food extrusion, and fermentation, for example, the work input term (w) in the energy balance is highly significant.

2.3.4 Examples of Energy Balance Calculations

2.3.4.1 Continuous Heat Exchangers

Consider the continuous heat exchanger shown in Fig. 2.20. Assuming that the heat losses and work done on the systems are negligible, writing the energy balance (2.35) in terms of rates gives:

$$\sum (mh)_{\text{in}} - \sum (mh)_{\text{out}} = 0 \quad (2.43)$$

where “ Σ ” denotes “the sum of”. Since there are only two streams (labeled “1” and “2”), this becomes:

$$m_1(h_{1,\text{in}} - h_{1,\text{out}}) = m_2(h_{2,\text{out}} - h_{2,\text{in}}) = Q, \quad (2.44)$$

i.e., the *change* in enthalpy of one stream exactly balances the other. Note that Q is the rate of heat transfer between the two sides of the exchanger. It is also known as the heat load.

Assuming no phase changes and that the streams have constant heat capacities C_{p1} and C_{p2} , (2.44) becomes:

$$m_1 C_{p1} (T_{\text{in}} - T_{\text{out}}) = m_2 C_{p2} (\Theta_{\text{out}} - \Theta_{\text{in}}) = Q. \quad (2.45)$$

2.3.4.2 Evaporators

Figure 2.20 also illustrates a simple continuous evaporator. Process liquid entering at T_i is heated to its boiling point T at the operating pressure P ; vapor and concentrated liquid also leave at T . Heat is supplied by condensing saturated steam at flowrate S and temperature T_s . Again assuming no sub-cooling, an energy balance over the evaporator implies that the loss of enthalpy on the steam side must be balanced by the increase in enthalpy on the process side, viz.:

$$Sh_{fg,T_s} = Vh_{g,T} + Lh_{f,T} - Fh_{f,T_i}. \quad (2.46)$$

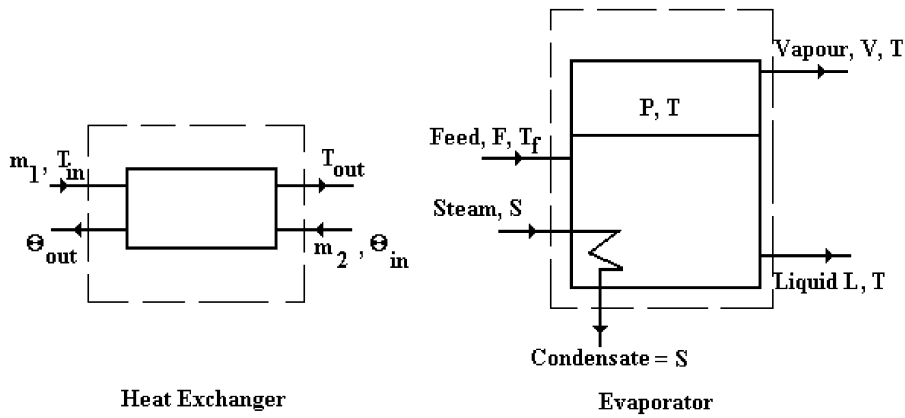


Fig. 2.20 Heat exchange and evaporation

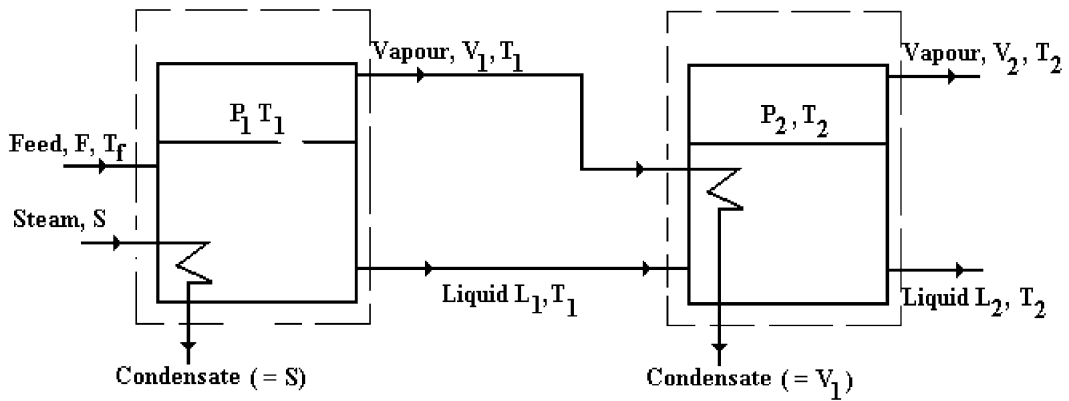


Fig. 2.21 Two-stage evaporator

For a liquid with constant mean specific heat capacity C_p this reduces to

$$Sh_{fg,T_s} = VC_pT + Vh_{fg,T} + LC_pT - FC_pT_i. \quad (2.47)$$

where $h_{fg,T}$ and h_{fg,T_s} denote the latent heat at T and T_s , respectively.

The process flows are also linked through the overall mass balance:

$$F = V + L. \quad (2.48)$$

Since the latent heat of evaporation of water is high the “ Vh_{fg} ” term always dominates the right hand side of the energy balance, (2.47); V is thus similar in magnitude to, but always smaller than, the steam flow S . With an idealized evaporator, 1 kg of steam produces around 1 kg of vapor. In practice, the ratio is more likely to be around 0.6–0.8.

This is the motivation for multi-effect evaporation. Figure 2.21 shows a two-stage forward feed evaporator: vapor from the first “effect” is used as the heat supply to the second. The temperature (and therefore the pressure) in the first effect must, of course, be higher than the temperature in the second effect if heat is to be transferred to the second stage. Here 1 kg of steam fed to the first effect

produces almost twice as much vapor as in the single effect system. Each additional effect improves the energy efficiency whilst, of course, adding to the investment costs. The optimum is where the benefits and costs balance.

Example 8. Consider an evaporator designed to raise the solids content of a milk stream from 5 to 10 %. Assume that the milk is fed at 12 °C, that evaporation takes place at 50 kPa (approximately 0.5 atm), and that the process streams have the same thermodynamic properties as water. From steam tables, the milk boils at 81.3 °C. The energy supply is dry saturated steam at 800 kPa (170 °C), with latent heat $h_{fg} = 2,048$ kJ/kg.

Take the basis of the calculations as 100 kg milk feed. The flowrate of liquid leaving the evaporator is $L = 100 \times 5/10 = 50$ kg and, from the overall mass balance, $V = 50$ kg.

Taking enthalpy values from steam tables, the energy balance (2.47) becomes:

$$S \times 2,048 = 50 \times 341.3 + 50 \times 2,304.1 + 50 \times 341.3 - 100 \times 48.8$$

or

$$S = 70.5 \text{ kg.}$$

Example 9. We consider the milk processing line shown in Fig. 2.2 and aim to calculate the cooling and heating requirements for the cooler, the storage tank and the pasteurizer. In addition to the plant items shown, a cooler operates on the pasteurizer discharge.

The following conditions are assumed:

Tanker capacity: 10 t milk.

Milk delivered at 10 °C.

Time to discharge milk to storage = 20 min.

Milk to be cooled to 4 °C.

Coolant: glycol solution at -10 °C; maximum temperature to be -5 °C.

Storage conditions: milk at 4 °C. Heat losses to environment average 5 kW.

Time to process batch through pasteurizer: 30 min.

Pasteurization temperature = 70 °C.

Pasteurizer heated (indirectly) by dry saturated steam at 180 °C.

Final delivery temperature of milk from pasteurizer = 5 °C.

In addition:

Assume mean heat capacity of milk = 4.2 kJ/(kg K).

Assume mean heat capacity of glycol solution = 3.5 kJ/(kg K).

Neglect heat losses from the cooler and pasteurizer.

Neglect subcooling of steam (i.e., condensate leaves at 180 °C).

From steam tables the properties of steam at 180 °C are:

$$h_f = 763.1 \text{ kJ/kg.}$$

$$h_{fg} = 2,014.9 \text{ kJ/kg.}$$

$$h_g = 2,778 \text{ kJ/kg.}$$

In practice, some preliminary cooling (in the first stage) and heating (in the pasteurizer) are needed to reach steady state. We neglect these effects and assume steady-state behavior throughout. The material balance is trivial: neglecting losses in the system, the total quantity of milk processed is 10 t.

The datum level for the energy balance calculations is liquid water and milk at 0 °C. We can establish the energy balance over each plant item in turn.

1. Cooler

All the heat lost by the milk is transferred to the glycol; there are no additional heat or work inputs to the system. Let G = glycol flowrate. Then the average energy (power) balance (2.45) becomes:

$$(10,000 \times 3/3,600) \times 4.2 \times (10 - 4) = G \times 3.5 \times 5 = \text{Heat load} = 35 \text{ kW},$$

and $G = 2 \text{ kg/s}$, i.e., 7.2 t/h. Assuming a batch discharge time of 20 min, this corresponds to a total glycol flow of 2.4 t. The refrigeration system required to maintain the glycol temperature at -10°C would therefore need a capacity of at least 40 kW for steady operation. In practice the maximum rating would be considerably greater than this to permit rapid start up, etc.

2. Storage tank

A further 5 kW cooling capacity would be needed (on average) to maintain the milk temperature during storage.

3. Pasteurizer

Let S = steam flowrate (kg/s).

The energy balance is given by (2.46). The heat load on the system is:

$$10,000/(0.5 \times 3,600) \times 4.2 \times (70 - 4) = S(h_g - h_f) = Sh_{fg} = S \times 2,014.9.$$

$$= 1,540 \text{ kW (for 30 min)}$$

$$\text{and } S = 0.764 \text{ kg/s or } 2.7 \text{ t/h.}$$

The total steam requirement per batch is thus 1.376 t, corresponding to a total heat load of 2.77 MJ.

4. Cooler

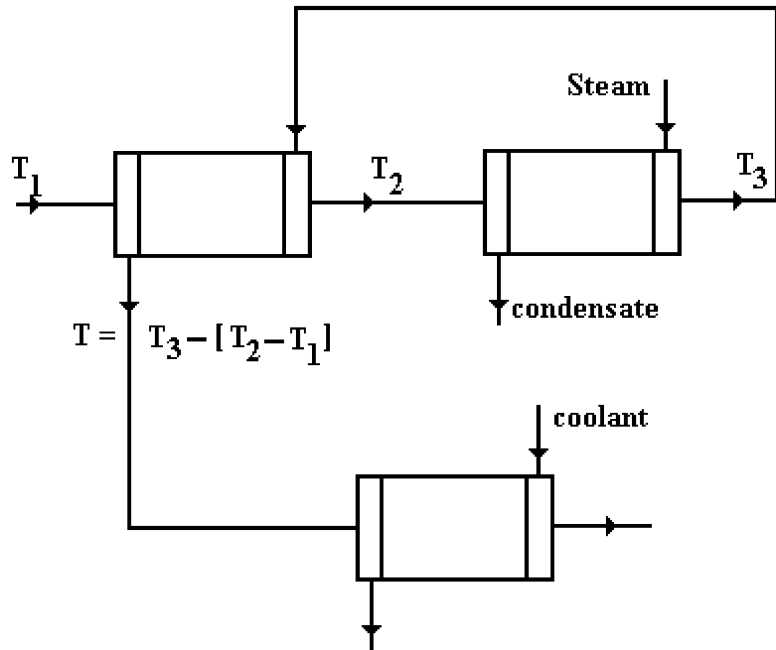
There are two possibilities. One is to use glycol from the refrigeration circuit. The other is to use cooling water (typically at $15\text{--}18^\circ\text{C}$) to reduce the temperature to around 25°C and then to further cool the stream using glycol to the desired temperature. The total heating load will be the same in both cases: the problem is essentially economic (cooling water being cheaper than refrigerant).

In either case the total heat to be rejected by the cooling system $= 10,000 \times 4.2 \times (70 - 5) = 2.6 \text{ MJ}$. (This, of course, is almost the same as the heat added to the system by the steam.)

Even if there are no other process units involved in the factory, the flowsheet used in this problem is very inefficient from an energy conservation point of view. This refers particularly to the pasteurization/cooling system since the energy transferred from the steam is rejected into the cooling water and/or glycol at a low temperature with no attempt at recovery. A more efficient scheme might be to use the warm milk stream leaving the pasteurizer to heat up the incoming feed. In practice it would be uneconomic to try to raise the temperature to more than about 55 or 60°C since this would imply a very large heat exchanger. The flowsheet in Fig. 2.22 shows how this might be achieved. Assume that the milk feed is preheated to 60°C in the first heat exchanger. The heat load on this exchanger is thus $(10,000 \times 3/3,600) \times (60 - 4) \times 4.2 = 1,960 \text{ kW}$, which is supplied by the warm milk. Since the same process liquid is being used to heat itself (i.e., the mass flowrate and thermal properties are the same on both sides of the heat exchanger), the warm milk must cool down by exactly the same amount as the cold stream warms up—i.e., to 14°C .

The steam requirements in the pasteurizer are now greatly reduced, since the heat load is reduced in the ratio 10/66, giving a steam requirement of 0.417 t/h and a total requirement of 0.21 t. The cooling requirements are also dramatically reduced.

Fig. 2.22 Pasteurizer with heat recovery



In practice, the limitation on efficiency is economic, stemming from the second law of thermodynamics. For heat transfer there must always be a finite driving force. As the exit and inlet temperatures on the two sides of the exchanger approach each other, the size of the heat exchanger also increases. Approach temperatures of around 10°C are typical.

From a rather broader perspective, we can note that a design study must incorporate aspects of the plant scheduling, particularly if the plant is a typical multiproduct, multipurpose unit. The operation of the heating, cooling and refrigeration systems can also be shown on the Gantt chart. This allows any bottlenecks and constraints to be identified and provides a basis for optimizing resource use across the whole set of operations.

2.3.4.3 Coupled Heat and Mass Balances

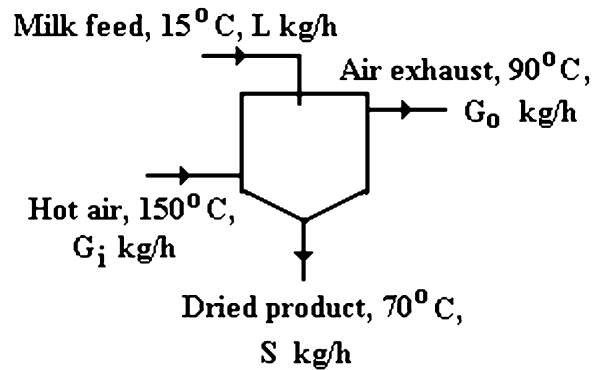
Sometimes it is not possible to solve the mass and energy balances independently. The solution depends on solving both balances simultaneously. This is illustrated in the final example of a spray dryer.

Example 10. 100 kg/h of milk powder containing 2 % moisture are produced in a continuous spray dryer. The feed solution contains 40 wt% milk solids and enters at 15°C . Atmospheric air with humidity $Y = 0.005$ kg water/kg dry air is heated to 150°C before entering the dryer. The air stream leaves the dryer at 90°C , and the solids product leaves at 70°C . The process is outlined in Fig. 2.23. Neglecting any heat losses, calculate the inlet liquid flow, the air flow, and the exit humidity of the air stream. Set up an overall summary mass and energy balance.

Data:

- Mean heat capacity of dry air = 1 kJ/kg K.
- Mean heat capacity of water vapor = 1.67 kJ/kg K.
- Mean heat capacity of dry solids = 1.6 kJ/kg K.
- Mean heat capacity of liquid water = 4.2 kJ/kg K.

Fig. 2.23 Spray dryer



Latent heat of evaporation of water at $0^\circ\text{C} = 2,500 \text{ kJ/kg}$.

Nomenclature:

L = liquid feed rate (kg/h).

g = dry air rate (kg/h).

G_i = total air flow in (kg/h).

G_o = total air flow out (kg/h).

S = product rate = 100 kg/h.

Solution:

Basis: 100 kg spray dried product.

Datum: 0°C ; liquid water.

1. *Balance on dry solids*

Solids in = Solids out.

$$0.4L = 98.$$

Therefore, $L = 245 \text{ kg/h}$.

Water in feed = $0.6L = 147 \text{ kg/h}$.

2. *Dry air*

The dry air in ($=g$) is the same as the dry air out in the wet stream.

3. *Water balance*

Water in hot air + Water in the liquid feed = Water out in exit air + Water in dried solids

$$0.005g + 147 = Yg + 2,$$

where, Y = humidity of exit stream. Thus,

$$g(Y - 0.005) = 145. \quad (2.49)$$

4. *Enthalpy balance*

The energy balance is

$$\text{Enthalpy of } L + \text{Enthalpy of } G_i = \text{Enthalpy of } G_o + \text{Enthalpy of } S, \quad (2.50)$$

where the enthalpies are respectively:

$$L: (147 \times 4.2 + 98 \times 1.6) \times 15 = 11,613 \text{ kJ}.$$

Table 2.7 Summary of overall mass and energy balances for Example 10

Component	In		Out	
	kg/h	kW	kg/h	kW
Milk solids	98	0.653	98	3.049
Water	147	2.572	2	0.163
Total (liquid streams)	245	3.225	100	3.212
Dry air	6,351.16	264.63	6,351.16	158.78
Moisture	31.76	24.27	176.75	130.12
Total (air streams)	6,382.92	288.9	6,527.91	288.9
Total	6,627.92	292.13	6,627.91	292.11

$$G_i: 150 \text{ g} + 0.005 \times g (1.67 \times 150 + 2,500) = 163.75 \text{ g kJ.}$$

$$G_o: 90 \text{ g} + gH(1.67 \times 90 + 2,500) = (90 + 2,650.3H) \text{ g kJ.}$$

$$S: (98 \times 1.6 + 2 \times 4.2) \times 70 = 11,564 \text{ kJ.}$$

Thus:

$$11,613 + 163.75 \text{ g} = 11,564 + (90 + 2,650.3Y)g$$

or,

$$g(2,650.3Y - 73.75) = 49. \quad (2.51)$$

5. Solution for g and Y

Equations (2.49) and (2.51) can now be solved simultaneously for g and Y , to give:

$$g = 6,351.2 \text{ kg dry air/h,}$$

$$Y = 0.028 \text{ kg/kg dry air,}$$

so that

$$G_i = 6,351.16 \times (1 + 0.005) = 6,382.9 \text{ kg/h.}$$

$$G_o = 6,351.16 \times (1 + 0.02783) = 6,527.9 \text{ kg/h.}$$

The heat and mass balance is summarized in Table 2.7. The values in kW are obtained by dividing the enthalpy values above by 3,600.

2.3.5 Energy Integration

The discussion of pasteurization processes illustrated that, as well as considering the design of each individual piece of equipment, it is also important to consider the design of the whole system. This can have great significance for the way in which the individual operations interact with each other—and for the way in which the plant is to be controlled and operated. In particular, the rationalization of energy use across the whole plant must be considered. The pasteurization example showed how a

simple recycle loop within the processing line can have a profound effect on energy use. Preheating the feed by the same, but hotter, stream, reduced the steam and cooling requirements dramatically. In a real plant there are potentially many hot and cold streams. The question thus arises: how can the hot streams be used to heat the cold streams to their target temperatures whilst minimizing the cost of equipment, steam, cooling water, and refrigerant?

The problem is readily grasped, but not so easily solved; however, considerable progress has been made in developing the theoretical basis (which is essentially the second law of thermodynamics) and computational methods for solving this class of problem. The approach employed is commonly known as Process Integration or Pinch Technology. Although most of the methods now available were originally developed for continuous processes, techniques have also been formulated for intermittent batch and semi-batch processes. Process integration is an important weapon in the designer's armory, and should certainly be used in design, just as upgrading and resource recovery from "waste" streams must be considered very seriously. Kemp (2006) provides a useful introduction to the subject.

2.4 Conclusions

In this chapter, we have outlined some of the issues and the tools for undertaking the preliminary process design of a food-processing facility. The principal design tools for estimating the material and energy flows and requirements—which are fundamental to the whole process—are material and energy balancing. Some of the principles of these methods and their application to food operations have been outlined. It is stressed throughout that an integral part of any design must be to consider how the various parts of the plant are likely to be used—i.e., scheduling. Bottlenecks that might arise from a lack of equipment or constraints on services, such as energy or cooling supplies, must be identified at the design stage. Equipment (including storage) sizing is profoundly influenced by the scheduling strategy. In any event, an over-riding concern must be to design a plant with a view to flexibility, since the rate of innovation of new products is high in the food sector. The importance of linking together aspects of hygiene (especially CIP), HACCP, product quality, and process control at an early stage in the design process has also been stressed.

References

- Blanc, D. (2006) "ISO 22000 – From intent to implementation", *Management Systems*, May–Jun, 7–11.
- Green, D.W. and Perry, R.H. (2008) (eds) "Perry's chemical engineers' handbook", 8th edit., McGraw Hill, New York, NY, Section 2.
- Kemp, I.C. (2006) "Pinch analysis and process integration", Butterworth-Heinemann, Oxford, UK, 416 pp.
- Rahman, M.S. (2009) (ed.) "Food properties handbook", 2nd edit., CRC Press, Boca Raton, FL, 859 pp.
- Sahin, S. and Sumnu, S. G. (2006) "Physical properties of foods", Springer, New York, NY, 258 pp.

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