9 Thermal treatment of foods

P.J. FRYER

Introduction

Many processes in the food industry use heat transfer. Rather than describing these processes in detail, and thus becoming rapidly out of date whenever a new process is introduced or a modification developed, this book seeks to explain the principles underlying them. The concepts introduced in previous chapters, such as the ideas of heat transfer as a set of sequential steps, of residence time distributions, and of fluid flows, will be applied here to the study of real systems.

Thermal sterilization is taken as an example where a process engineering approach has led to a redesign of real plant. Although the canning process is efficient in a mechanical sense, it does not produce high quality product. The type of analysis developed in the previous chapters can be used to explain why this happens, along with ways of designing processes to enhance product quality. Higher quality product can be produced in continuous sterilizers which use the principles developed in Chapter 3; for example it is necessary to know the range of product residence times to ensure that the process is safe. To design plant that processes food in continuous flow requires an understanding of the flow properties of the fluid and of the rates of heat and mass transfer which will result from that processing.

It is difficult to heat or cool solid foods quickly because of the slowness of thermal conduction through solid materials. Alternative methods, in which heat is generated within a food by the action either of electric current or microwaves, allow rapid heating of the food. These processes require different skills of the food process engineer, and the rates of heating are often difficult to predict. Although the equations which govern electrical and microwave heating are difficult to solve in practice, they are introduced here to show how processes could be designed and controlled. The problems of continuous sterilization clearly demonstrate the importance of the ideas developed elsewhere in the book.

Finally, this chapter considers one of the key processing problems, that of the fouling of heat exchanger plant and the consequent need to clean the equipment. Fouling problems are very rapid and severe in food processing;

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frequently, daily cleaning of process plant is required. The operation of food plant undergoing fouling is very difficult to predict. However, simple models for fouling can be developed, and guidelines for minimizing fouling in practice produced. This book has shown frequently that processes must be designed as a unit rather than as a series of individual steps; the length of operating cycles must be set to ensure that the plant is not difficult to clean when processing is stopped.

Engineering in food preservation

Historically, food preservation techniques were developed to enable locally grown food to be eaten all year round. The increased separation of the producer and consumer in an industrial society added another reason: food must be safe and palatable as it reaches the consumer. Many ways of preserving foods are used industrially, among which are:

- **drying**, which aims to reduce the water activity within the food to prevent bacterial and enzyme action, typically to a water content less than c.5%;
- adding **preservatives**, such as salt, which retard bacterial action;
- **low-temperature storage** – freezing, which effectively prevents microbial growth, and refrigeration, which slows it to acceptable levels;
- **high temperatures**, in which microbes and microbial spores are killed by heat;
- **asepsis**, in which food is kept sterile throughout processing;
- the **removal of organisms**, such as by filtration;
- **irradiation**, in which high-energy radiation is used to destroy contaminants.

The demands of the consumer determine which products are successful. Making food processes produce safe and acceptable products is extremely difficult. A variety of engineering skills are required. Equipment and materials for the process plant are designed by mechanical engineers, while electrical engineers design and install the systems that enable the plant to be operated and controlled. The role of the chemical engineer is less immediately obvious; the chemical engineer is concerned within the design and specification of the process itself. The aim of this book is to demonstrate the ways in which chemical engineering techniques can be applied to food processing. This chapter uses the principles outlined elsewhere in this book and applies them to the problems of thermally processing foods. This is an area of active research in which process engineering techniques are being increasingly applied.

Chemical engineering evolved to meet the challenges of the oil industry. In petrochemical and plastics processing, heat transfer and chemical reactions generally occur separately; reactants are heated to the required tem-

perature, the reaction occurs, and then heat is removed from the products. Many petrochemical fluids can also be heated up and cooled down without the properties of the fluid changing; if reactions do occur, they may well be reversible. This rarely, if ever, happens in the food industry. In food processing, the rates of reaction are commonly so swift that heat transfer and reaction cannot be separated; whenever a food material is heated, changes occur within the food that cannot be reversed by subsequent cooling. Food-heating processes must thus be considered in terms of both heat transfer and reaction.

Heating and cooling in the food industry

Heating is carried out both to preserve the food product and to add palatability – and digestibility – to the material. Foods are so thermally unstable that reactions occur very rapidly at low temperatures and pressures, below 150 °C and at 1 bar, whereas in the petrochemical industry much higher temperatures and pressures are common. The chemical processes that occur on heating foods are complex, but have been thoroughly studied by food scientists. When a food is heated, reactions occur that result in increased sterility and in the development of the taste and texture of the food material. Thermal processing results in protein denaturation and aggregation reactions, which cause enzyme deactivation and consequent bacterial death.

A number of types of heating process are found in the food industry. They include:

- **blanching** or **pasteurization**, which are designed to extend the shelf-life of a product by destroying enzymes that might subsequently reduce product quality, by perhaps allowing volatile components to escape and reducing the content of bacteria (these do not result in substantial changes to the structure or properties of the material);
- **baking**, **roasting** and **frying** – 'cooking' processes in which significant changes are made to the food, which can then be eaten directly;
- **sterilization** processes, in which the amount of bacterial contamination in a food is reduced to a statistically insignificant level, enabling the food to be stored for a considerable length of time. Once a product has been made sterile, it must be kept sterile. Aseptic packaging processes, which allow sterile material to be packed and sealed without contamination, have been developed for liquids and then solid–liquid mixtures, making the commercial development of continuous sterilization processes possible.

A number of cooling processes are also common:

- **refrigeration** to slow bacterial action – for example, 'cook-chill' meals, which have been cooked as in the home and which are then chilled prior

to distribution, rely on this slowed bacterial growth to prevent damage to the food;

- **freezing** to prevent bacterial and enzyme-induced decay, allowing food to be stored for a considerable time. Chapter 3 has discussed simple models for freezing which allow an approximate estimate of the freezing time to be made. However, the structural damage caused by the expansion of water on freezing can impair product quality. The selection of optimal freezing conditions requires an understanding of material properties as well as information on the design of the freezer.

Sociological factors, such as the increase in the number of women employed outside the home, have increased the market for food that requires minimal home cooking, and thus for 'ready-meal' products. Neither canning nor bottling, which contain material that has been extensively processed, are well suited to this type of product. The range of storable food products available has grown enormously over the last 20 years, beginning with dried foods, which involve significant home preparation before they can be eaten, and then frozen foods, which, although convenient, require a freezer. More recently, cook-chill meals, bought from the cold cabinet and then kept in the refrigerator, and ambient shelf-stable products, which can be stored at room temperature, have been introduced.

The production of sterile food of high quality is an area that still presents many engineering challenges. This chapter will concentrate on sterilization as an example of the ways in which chemical and process engineering techniques can be applied to a food industry problem. Effectively, two sorts of reaction are taking place when food is cooked: those that lead to a sterile product, and those that result in a loss in product quality. The basic problem of sterilization is in many cases to maximize the first set of reactions whilst minimizing the second; this sort of optimization is common in the petrochemical industry, and can be examined with chemical engineering principles.

The kinetics of cooking

Thermal processes involve chemical reactions that can be quantified if they are to be optimized. As discussed in Chapter 8, the rates of chemical reactions can be expressed as a chemical rate law, a function of concentration, with a temperature dependence characterized by an activation energy:

$$\text{Rate} = k_r c^n = A \exp\left[\frac{-E_a}{R_g T}\right] c^n \tag{9.1}$$

but the rates of food processes are not generally described in terms of conventional chemical kinetics. The F-value (in minutes) is used to define the amount of sterilization that the food has received. If the temperature T

of the food is known as a function of time t the **integrated lethality**, defined as

$$F = \int_{t_{START}}^{t_{END}} 10\left(\frac{T - T_{ref}^{F}}{z_F}\right)dt \tag{9.2}$$

can be calculated. As described in Chapter 8, this is widely used within the industry.

It is much more difficult to quantify quality, which is a more non-mathematical concept, than to measure sterility. Some attempts have been made to relate quality in similar terms to the F-value, using data for enzyme loss. Thus it is possible to define the **C-value** of a process:

$$C = \int_{t_{START}}^{t_{END}} 10\left(\frac{T - T_{ref}^{C}}{z_C}\right)dt \tag{9.3}$$

where T_{Ref}^{C} is the reference temperature (now generally 100 °C) of the process at which the slope of the 'cooking' rate curve is z_c. This approximates the true kinetics in the same way as does the F-value, but has the added complication that it is attempting to measure a much more diffuse quantity. Some work has been done relating the C-value to quality; a value of 100 min is generally taken as corresponding to 'satisfactory' for z_c in the range 20–35 °C.

To see how to design process plant, activation energies are more immediately useful than z values. A table of activation energies has already been given (Table 8.1 in Chapter 8), and can be used with equation (9.1) to investigate the effects of temperature on the two rates.

EXAMPLE 9.1

Compare the ratios of the rates of the sterilization and quality loss at 120 °C and 140 °C, assuming: (a) the activation energy for C. botulinum *death is 300 kJ mol⁻¹; (b) the activation energy for the loss of quality in foods is 125 kJ mol⁻¹.*

If sterilization is an nth-order reaction while quality loss is an mth-order reaction, the ratio of the rates of the two at any temperature T is

$$\frac{\text{Rate of sterilization}}{\text{Rate of quality loss}} = \frac{A_s \exp\left[\dfrac{-E_{a(s)}}{R_g T}\right] c^n}{A_q \exp\left[\dfrac{-E_{a(q)}}{R_g T}\right] c^m} \tag{9.4}$$

When the ratios are compared for the same concentration but different

temperatures, the pre-exponential factors and concentration dependences cancel. R_g is 8.314 J mol^{-1} K^{-1}, so $E_{a(s)}/R_g = 300\,000/8.314 = 36\,083$ K. The ratio is thus

$$\frac{\text{Rate at 140°C (313K)}}{\text{Rate at 120°C (293K)}} = \frac{\exp\left[\dfrac{-E_{a(s)}}{R_g 313}\right]\exp\left[\dfrac{-E_{a(q)}}{R_g 293}\right]}{\exp\left[\dfrac{-E_{a(q)}}{R_g 313}\right]\exp\left[\dfrac{-E_{a(s)}}{R_g 293}\right]}$$

$$= \frac{\exp[-115.28]}{\exp[-48.04]}\,\frac{\exp[-51.313]}{\exp[-123.152]} = 98.7 \quad (9.5)$$

The calculation in Example 9.1 shows that sterilization reactions proceed about 100 times faster than loss-of-quality reactions at the higher temperature. However, it says nothing about the absolute rates of the reactions. As a result of the temperature change, the time required for a given process will also change, as shown in the next example.

EXAMPLE 9.2

Compare the time required to carry out a sterilization at 120 and 140 °C, for the same data as Example 9.1.

Here only a single ratio is required since the ratio of the rates is the inverse of the times:

$$\frac{\text{Rate at 140°C (313K)}}{\text{Rate at 120°C (293K)}} = \frac{\exp\left[\dfrac{-E_{a(s)}}{R_g 313}\right]}{\exp\left[\dfrac{-E_{a(s)}}{R_g 293}\right]} = \frac{\exp[-115.28]}{\exp[-123.15]} = 2600 \quad (9.6)$$

Example 9.2 shows that, if sterilization is carried out at high temperature, both the time needed for sterilization and the amount of quality loss are reduced. There are thus good reasons for sterilizing at high temperatures and for short times. Before studying this further, it is important to understand why classical sterilization processes, such as canning, may be unsatisfactory.

9.1 Engineering principles

9.1.1 Heat transfer and sterilization: convection and conduction

Canning is the classical sterilization process, invented by Appert in the 19th century, and still the basis of a very large industry. A typical process, based

on a series of pressurized steam cookers containing many cans, is shown schematically in Fig. 9.1. The key step in ensuring sterility is the mechanical engineering process of putting the lid on the can. The lidding process is in general not aseptic; usually, cans must be sterilized after the lid has been applied and sealed.

The two governing processes of heat transfer to the sealed can are convection and conduction, discussed in Chapter 3. Enough heat must be given to the system to sterilize the centre of the can. Heat is generally applied by condensing steam. The heat transfer coefficient from the steam to the outside of the can is very high. It might be thought that this would lead to rapid heating. However, Chapter 3 has shown that an overall heat transfer process is governed by the rate of the slowest, 'rate-controlling' step. This can be seen by the construction of the overall heat transfer coefficient in equation (3.21). The Biot number of the can defines the relative roles of internal and external heat transfer in a system.

EXAMPLE 9.3. ESTIMATE THE BIOT NUMBER OF A CAN

The standard size of a UK can is a cylinder about 100 mm high and 66 mm in diameter. An appropriate Biot number would be hr/λ, where h is the external heat transfer coefficient, r the radius of the cylinder and λ the thermal conductivity. Here $r = 0.033$ m, and assuming the filling has the properties of water at 20 °C the thermal conductivity is $0.7\,W\,mK^{-1}$. A lower estimate for the heat transfer coefficient for condensing steam is $1000\,W\,m^{-2}K^{-1}$, so the Biot number will be $1000 \times 0.033/0.7 = 47$. This is a very high Biot number.

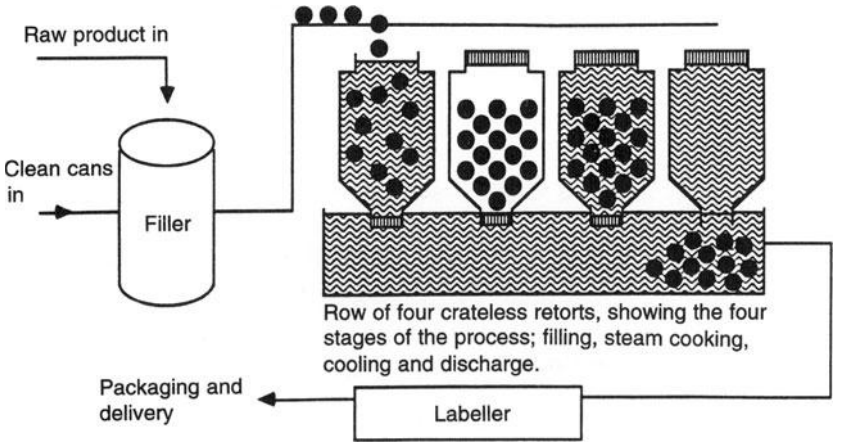


Fig. 9.1 Schematic diagram of a semi-continuous canning process, showing a row of four crateless retorts. Use of the four retorts enables the process to run continuously.

The Biot number in Example 9.3 is so large that heat transfer within the can will be the slowest step. The temperature at the centre of a can will lag behind the outside by an amount dependent on (a) the size of the container and (b) the mode of heat transfer. If the food is a liquid, it will heat by natural convection, setting up a circulating flow within the can that acts to stir it; this ensures that it heats quite quickly. If the food behaves as a solid (in practice, is a solid or a very viscous liquid) however, it will only heat by conduction: a slower process as shown in section 3.3.5. A lower limit can be placed on the possible sterilization time if it is assumed that the liquid within the can is fully stirred: that is, at a uniform temperature.

EXAMPLE 9.4

Assuming (a) that the fluid in the can is fully stirred and (b) that the heat transfer coefficient between the outside of the can and the condensing steam is infinite, find an equation from which the required heating time can be calculated.

Let the volume of the can be V and its surface area A_c, the temperature of the steam and the can be T_s and T_c and the density and specific heat of the food be ρ and c_P respectively. Then a heat balance can be written to describe the change in temperature with time. The heat transferred to the food increases its temperature according to

$$UA_c(T_s - T_c) = V\rho c_P \frac{dT_c}{dt} \tag{9.7}$$

where U is the overall heat transfer coefficient. If the steam temperature is constant, this equation can be integrated as

$$\frac{T_s - T_c(t)}{T_s - T_c(0)} = \exp\left(-\frac{A_c U t}{V\rho c_P}\right) \tag{9.8}$$

That is,

$$t = -\frac{V\rho c_P}{A_c U} \ln\left(\frac{T_s - T_c(t)}{T_s - T_c(0)}\right) \tag{9.9}$$

The heating time thus depends on both the ratio of the can volume to the surface area and the heat transfer coefficient U between the steam and the food. This will be governed by natural convection between the fluid and the inside and the can.

In Example 9.4, if the steam temperature $T_s = 125\,°C$, the initial and final can temperatures are 60 °C and 123 °C, ρ and c_P are for water, and the can

dimensions are 66 mm in diameter and 102 mm deep, then substituting into equation (9.9) gives $t = 180000/U$ s. The internal heat transfer coefficient could well be low, say $U = 300\,\mathrm{W\,m^{-2}K^{-1}}$, giving a heating time of 600 s: about 10 min. The reader may wish to calculate the process time for a catering product in a can where the linear dimensions are doubled.

In practice, longer cook times are needed, because it is not economically sensible to sterilize single cans. Cans are processed either in batch retorts (pressure cookers in which several thousand cans are sterilized simultaneously) or in continuous systems, in which cans are conveyed through regions of varying temperature. Not all the cans will be exposed to the steam at the same time. This creates a problem: the process time must be set to ensure that the centre of the coldest can in the retort is sterilized.

The net result is to ensure that the vast majority of food in cans has a higher F-value that that specified for the process. In many cases this heat is so great that the food is significantly overprocessed. C-values of 400–500 min can be found in canning or retort pack foods, giving tinned vegetables and fruit that may bear little relation to the fresh product. In some cases canned food may be acceptable, such as in baked beans and mushy peas, but for some foods, such as strawberries, the reduced quality of the canned product is less acceptable.

One approach would be to sterilize using higher steam temperatures, in an attempt to process for a shorter time. However, the size of the can means that thermal lag will always be present, so the outside of the can will always be overcooked.

EXAMPLE 9.5

Estimate the increase in quality damage if the process of Example 9.4 is carried out using steam at 140 °C rather than 125 °C.

Equation (9.9) can be used to estimate the change in the process time:

$$\frac{\text{Time at } 140\,^{\circ}\text{C}}{\text{Time at } 125\,^{\circ}\text{C}} = \frac{\ln\left(\dfrac{140-123}{140-60}\right)}{\ln\left(\dfrac{125-123}{125-60}\right)} = 0.44 \qquad (9.10)$$

but the ratio of the accumulation of C-value will also change. For $T^{C}_{ref} = 100\,^{\circ}\text{C}$ and $z_c = 20\,^{\circ}\text{C}$, then

$$\frac{\text{Rate of quality loss at } 140\,^{\circ}\text{C}}{\text{Rate of quality loss at } 125\,^{\circ}\text{C}} = \frac{10^{\left(\dfrac{140-100}{20}\right)}}{10^{\left(\dfrac{125-100}{20}\right)}} = 5.62 \qquad (9.11)$$

So the increased quality loss as a result of moving to the higher temperature is $0.44 \times 5.62 = 2.47$.

The slowness of heat transfer, and the resulting quality losses, thus limit the use of higher temperatures in canning. In addition, the cost of pressure vessels is also limiting; if too high a steam temperature is used the vessel will be too expensive. Canning temperatures are limited to about 120–125 °C.

The advantages of the canning process are considerable, however. The process is basically cheap and straightforward and gives a robust and safe product. More than 100 years of operating experience in canning mean that the process and the factors that result in a safe product are well understood, and the product is also accepted by the consumer. Any alternative process must offer significant improvements if it is to be adopted by the industry.

9.1.2 The ideal food sterilization process

Kinetic data suggest that the higher the process temperature is, the better is the quality of the food when it is sterilized. The analysis of the canning process carried out above indicates that even if a high temperature is used, all parts of the food should be processed simultaneously, or else portions of the food will be overcooked. The requirements for an ideal process are thus:

- instantaneous heat transfer to the food, ensuring that all parts reach the same temperature at the same time, and follow the same time–temperature path;
- high temperatures, allowing sterilization reactions to predominate over the reactions that reduce quality.

Partial solutions are found if the distance over which heat has to be transferred is reduced, lowering the Biot number and decreasing the cooking time. A number of retortable pouches and trays are now available that combine a marketing advantage of a novel and convenient package (from which the food can often be eaten) with the better quality of food that results from faster heat transfer and thus a shorter process. Although these packs improve final product quality, they do not solve the problem; a new approach is needed.

The two 'ideal' concepts form the design philosophy behind **HTST** (high-temperature short time) or **UHT** (ultra-high temperature) processes. These processes are commonly considered to consist of three heat transfer stages, as follows.

1. Food is **heated** rapidly to temperatures around 130–140 °C.
2. It is **held** at high temperature in a holding section for the few seconds necessary to ensure sterilization.
3. It is then rapidly **cooled** before significant product degradation occurs.

This simple picture ignores preheating of ingredients and any packing step. Comparison of typical temperature–time curves for canning and HTST processes shows that much shorter processing times are possible.

9.1.3 Engineering implications of HTST

HTST process plant requires flowsheets that are significantly different from those for canning. Equipment is necessary to provide:

- high rates of heat transfer;
- aseptic packaging;
- uniform residence time distributions.

High rates of heat transfer. The rate of heating required by HTST processes is much faster than in canning. Forced convective heat transfer coefficients are significantly greater than those possible in conduction. Forced convection is also faster than free convection, which is driven by the density difference between hot and cold fluid: within a heated can, for example. It is thus possible to heat flowing food much faster than static food. HTST processes are thus ideally suited to continuous rather than batch operation. Processing is carried out by heating food before packaging through heat exchangers, which give high heat transfer coefficients, rather than after packaging, as in canning.

EXAMPLE 9.6

Calculate the heating rate possible when a food fluid with the thermal properties of water flows at 1 m s^{-1} down a 1 cm pipe, assuming (a) a 5 °C temperature difference between the inside and outside, and (b) that heat transfer coefficients are equal on both sides and that

$$\mathrm{Nu} = 0.023\mathrm{Re}^{0.8}\mathrm{Pr}^{0.4} \qquad (9.12)$$

applies on the inside.

For water at 1 m s^{-1} in a 1 cm pipe

$$Re = 10^6 \times 1 \times 0.01 = 10000$$

and

$$Pr = 5.8$$

so

$$Nu = 0.023 \times 10000^{0.8} \times 5.8^{0.4} = 74.3$$

so

$$h = 74.3\lambda/d = 74.3 \times 0.61/0.01 = 4535\,\mathrm{W\,m^{-2}\,K^{-1}}$$

So the overall heat transfer coefficient can be found from

$$\frac{1}{U} = \frac{1}{4535} + \frac{1}{4535} = 2268\,W\,\mathrm{m^{-2}\,K^{-1}}$$

Then the heat balance on an element of length dx is

$$h.\pi d\,\mathrm{d}x\Delta T = \rho c_p v \frac{\pi d^2}{4}\mathrm{d}T \tag{9.13}$$

and, as $\mathrm{d}x = v\,\mathrm{d}t$, then

$$\frac{\mathrm{d}T}{\mathrm{d}t} = \frac{4h\Delta T}{\rho c_p d} = \frac{4 \times 2268 \times 5}{1000 \times 4200 \times 0.01} = 1.08\,^{\circ}\mathrm{C\,s^{-1}} \tag{9.14}$$

Example 9.6 demonstrates that very rapid heating rates can be obtained for liquid foods using forced convection. However, the need for rapid convective heat transfer limits the applicability of HTST techniques to foods that contain particles. Particles heat by themal conduction. If large particles are processed by conduction, the time required to raise their temperature to the required level is such that, as with Example 9.5, by the time particle centres are sterile the liquid in which they were immersed is unacceptably overcooked. In practice, therefore, conventional heating techniques can only process very small particles, of the order of a couple of millimetres in diameter. Section 9.2 discusses some possible solutions to this problem.

Aseptic packaging. HTST processes involve the sterilization of food materials prior to packaging. Packaging process must therefore be sterile (aseptic): that is, no contamination of the already sterile product can be allowed to occur during filling. Until such processes were developed, HTST techniques could not be commercially used. A number of companies produce aseptic packaging equipment. Common features are:

- a prefill step, in which the unfilled pack or laminate is presterilized using (for example) steam, ultraviolet light or hydrogen peroxide;
- a filling step, in which measured amounts of material are passed to the carton (in a batch fill the nozzle is then flushed out, with steam for example, to ensure sterility);
- a sealing stage.

The requirement for sterility means that aseptic packaging plant requires a lot of complex and expensive mechanical engineering. It is more difficult to pack foods that contain particles than liquids, because of problems in ensuring that the particles do not block the filler head or make the seal

imperfect. Such problems have not been completely solved, but effective commercial solutions are available.

Uniform food residence time distributions. At UHT temperatures, degradation reactions are rapid. The need for product safety implies that the process must be designed to sterilize the material that travels at the fastest speed through the plant. However, if a significant fraction of the food is unacceptably overprocessed, then the equipment is a failure.

The concept of the **residence time distribution** (RTD), introduced and discussed in Chapter 8, is critical in the quality of continuously processed food. In food processing, it is important that all parts of the material are sterilized, and the food produced must reach a required quality standard. Two criteria, for **sterility** and **product quality**, must be met by any process plant. For example, the process objectives may be an F-value of at least 6 min in the whole medium. This process requirement can be termed the F_p value. In achieving this aim, quality changes must take place: beyond a certain C-value, C_{max}, the product will no longer be commercially acceptable. The acceptability of a food product can thus be expressed in terms of sterility and quality ratios, ϕ_F and ϕ_C, defined by

$$\phi_F = \frac{F}{F_p} \qquad \phi C = \frac{C}{C_{max}} \tag{9.15}$$

These two ratios can be used to decide whether the process is producing an acceptable product as, at each point,

- if $\phi_F < 1$, the food is insufficiently sterile;
- if $\phi_C > 1$, it is overcooked.

Figure 9.2 shows schematically the changes in ϕ_F and ϕ_C that take place during processing. Each point in the food must be processed for a time longer than t_{min}, the time at which $\phi_F = 1$, but for no longer than t_{max}, the time at which $\phi_C = 1$, to ensure that the food is of satisfactory quality. The task of the food engineer in designing a process is thus to ensure that residence times of food in the equipment lie between these two values. A possible problem resulting from velocity distributions in a process plant is shown by the next example.

EXAMPLE 9.7

Estimate the distribution in F-value that results when a fluid with the properties of water flows at 130 °C at a mean velocity of 0.02 m s^{-1} through a 5 cm diameter tube 2 m long. Take T_F^{ref} *= 121.6 °C and* z *= 10 °C.*

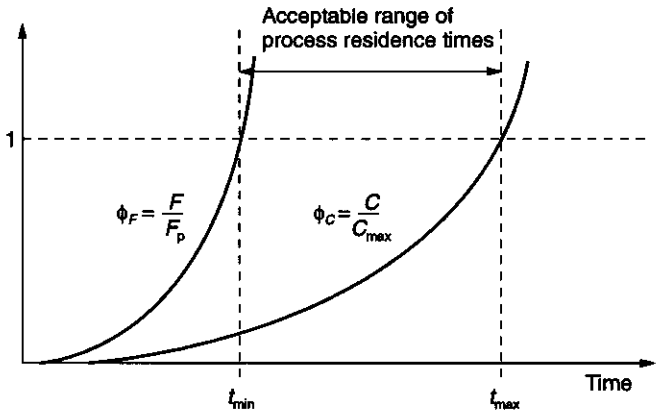


Fig. 9.2 The acceptable process RTD lies between the point where the food is sterile ($\phi_F = 1$) and where it is overcooked ($\phi_C = 1$).

The Reynolds number of the flow is ca. $1000 \times 0.02 \times 0.05/0.001 = 1000$: that is, the flow is laminar. As described in Chapter 2, the velocity profile of such a flow is parabolic:

$$v = 2v_m\left[1-\left(\frac{r}{R}\right)^2\right] \tag{9.16}$$

The residence time for fluid at the wall of the pipe is thus much longer than that for fluid along the axis. The RTD function for such a system is

$$E(t)dt = 2\frac{t_o^2}{t^3}dt \tag{9.17}$$

where t_o is the residence time along the axis. At 130°C, the change in F with time is given by:

$$\frac{dF}{dt} = 10^{(T-T_{ref})/z} = 10^{(130-121.6)/10} = 6.918\,\text{min}^{-1}$$

The velocity profile is such that for $r = R/2$, the velocity is only 75% of that at the axis; fluid at that radius thus spends 33% longer in the tube than fluid on the axis. Nearer the wall the residence time increases further until the layer at the wall is theoretically stationary. Along the centreline, the velocity is $0.04\,\text{m s}^{-1}$, the residence time is $2/0.04 = 0.833$ min, so that F is 5.76 min. At a radius of 4 cm, however, the velocity is $0.04 \times (1 - 0.8 \times 0.8) = 0.0144\,\text{m s}^{-1}$, so that the F accumulated along the tube is 16.0 min. The process must be designed to sterilize the fluid flowing along the centreline; material in the wall region will be significantly overcooked.

Developed parabolic flow gives the worst case; in many cases, hold tubes for particle–liquid mixtures are designed as if the flow were parabolic: that is, it

is possible for material to pass through the system at twice the mean velocity. This is a conservative estimate. A well-designed process plant should have as small a difference between minimum and maximum residence times as possible. It is important that RTD be measured using real process fluids; even for fluids such as milk and water, differences in RTD have been found in plate heat exchangers.

Similar analyses can be carried out on other food processing equipment. For example, the spray drier is widely used to produce powdered products. Here a food is sprayed through a nozzle that produces a dispersion of small particles (much less than 0.25 mm) into a countercurrent flow of heated air. This offers very rapid drying rates because of the high surface area available for mass transfer, and gives minimal heat damage to the food material. Flow patterns are extremely complex, however, and the design of the nozzle is critical. If droplets are too small they produce particles that can be entrained by the air. These drops will stay in the drier for too long or be carried out with the airstream. If droplets are too large they will not dry. If too wide a droplet size distribution is produced, therefore, problems will result with unevenly dried product, owing to a wide product RTD.

The idea of the RTD is a powerful one, which can be widely used in the analysis of process equipment.

9.1.4 Problems in HTST processing

The logic behind HTST processing is, as outlined above, straightforward, and the technical problems discussed above appear easy to solve. In practice, they are not.

Many of the **mechanical engineering design** problems of food processing are highly complex. Hygienic design of equipment is crucial; it is vital to ensure that it is impossible to contaminate the food stream from outside and that process plant is as simple to clean and sterilize as possible. This requires that plant be carefully designed, and made from materials that do not contaminate the food. The design of hygienic valves and the need for sterile cleaning-in-place (CIP) makes food plant very complex. This type of plant is essentially similar to a petrochemical plant; for safety, the status of each valve must be known and the system must be continuously monitored.

In addition, there are a number of problems with **flow and heat transfer**, owing to the nature of food fluids. Data are widespread on petrochemical fluids, which have fairly stable properties, and largely Newtonian viscosities. In contrast, food fluids are complex and thermally unstable, with non-Newtonian rheology, as discussed in Chapter 5. Rheology can also be strongly temperature-dependent as a result of reactions on heating, such as starch gelatinization or protein aggregation within the fluid. During cooling, the viscosity of the product will also change. It is very difficult to predict the flow and heat transfer properties of food fluids. In addition, the thermal

instability of foods makes process plant prone to the formation of fouling deposit: solidified material on the inside surfaces of processing plant, which must be frequently cleaned.

Control of continuous plant is much more important than in conventional batch canning. In canning, the process is well defined; provided a given temperature (or steam pressure) is maintained inside a retort for a given length of time, then all the food material will be sterile. In continuous plant, flowrate, pressure and temperature should be monitored throughout the plant to ensure that each stage of the process is operating correctly. This is especially important when dealing with complex fluids. For control and process validation, the development of **models** for the process is necessary; this requires an understanding of the process and of food physical and engineering properties.

The rest of this chapter will discuss some of these difficulties and their possible solutions.

9.2 Continuous processing: problems and solutions

9.2.1 Design principles for heat transfer equipment

The previous section has developed the arguments that led to the adoption of HTST processing. Continuous rather than batch processing is necessary, and it has been pointed out that the nature of food fluids makes them very prone to fouling and cleaning problems. The food industry uses a very wide range of heat transfer equipment. This section indicates the chemical engineering principles on which process equipment operates, rather than provide a guide to the best available.

Continuous heat exchange is commonplace in the chemical industry. Most of the petrochemical industry uses shell and tube exchangers of the type discussed in Chapter 3, which are fairly simple to design and fabricate and have high heat transfer coefficients. Food fluids are generally of high and non-Newtonian viscosity, and may contain particles. Flow of such a material through a tubular exchanger would give a poor RTD, poor heat transfer coefficients, and probably severe fouling. When selecting a heat exchanger, a number of questions must be answered, as follows.

- Is the RTD narrow enough to give uniform **final product quality**? Does the exchanger give unacceptable product damage (for example, food particles damaged by scraped surface units)?
- Is the **pressure drop** through the unit acceptable? If the food fluid is of high viscosity, very high pressure drops can result.
- Does the **extra value** given to the product by the unit make it cost-effective? For example, if an expensive scraped-surface exchanger is used

to increase product quality, does the increased price of the product generate enough extra revenue to offset the capital expenditure on the plant?

- Is it possible to **recover the energy** used in heating, by (for example) cooling the food down using material that is required at high temperatures?

The types of exchanger used in the industry will now be reviewed briefly.

Direct contact. One simple way of getting efficient heat transfer and minimizing fouling is to inject steam directly into or over the food material, eliminating the heat transfer surface. This principle is used in direct steam injection or infusion sterilizers. In the latter process, steam at a pressure higher than that of the product is injected into the product stream via a suitable nozzle, and its condensation releases latent heat, which increases the product temperature and causes sterilization. The food is then sprayed into a vacuum chamber to reduce its temperature by evaporation of added water, which requires latent heat to be absorbed. These systems offer efficient sterilization, but are highly energy-inefficient and expensive to operate.

Plate heat exchangers. These are the main type used in the food industry. Plate exchangers consist of a series of vertical shaped steel plates, separated by gaskets and held in a metal press, which form parallel corrugated channels through which liquid food and heating media can be passed in various configurations. High film heat transfer coefficients are possible even for viscous liquids, as thermal boundary layers are very thin and the plates are designed to enhance fluid mixing. Unlike tubular exchangers, few design correlations of the type described in the heat transfer section are available in the literature. The precise form of the correlations are particular to each manufacturer's design. One published equation for a heat exchanger is (Rene *et al.*, 1991)

$$Nu = 0.352Re^{0.639}Pr^{1/3} \qquad \text{for } Re > 5$$

where the hydraulic diameter (defined in equation (2.77)) is used as the length term in the Nusselt and Reynolds numbers. The form of this equation is similar to those given in the section on heat transfer and can be compared with the other equation for plate exchangers (p. 131). Care should be taken when calculating heat transfer coefficients in plate exchangers unless manufacturer's data are available for the particular case.

Plate exchangers are compact and are well suited to energy recovery. In pasteurizers, it is common for hot product to be cooled against inlet liquid; this makes them inherently more efficient than injection systems. They are, however, commonly limited by the pressure at which the gaskets burst (at

about 750 kPa) which means that only low flowrates are possible. Welded units are now available that can operate at higher pressures.

Scraped-surface exchangers. Many food materials are highly viscous and foul heavily; they are thus difficult to process in more conventional plant. The principle of the scraped surface exchanger is simple: food is passed through a heated chamber, which contains a rotating blade. This prevents deposition on the heat transfer surface and also stirs the food. Scraped-surface units are widely used for very viscous solid–liquid mixtures and for materials such as ice-cream and spreads. However, if they are incorrectly operated they can damage the food. Owing to the complexity of the flow patterns, it is difficult to design and model scraped-surface systems. A wide range of residence times may also be encountered in operating plant.

Tubular exchangers. Simple shell-and-tube exchangers are used to a small extent for food processing, but are limited to low-viscosity foods (up to about 2 N s m^{-2}). However, most food plant will contain such exchangers, for example in boiler or evaporator plant. Their design is straightforward and is discussed in Chapter 3. A number of companies offer modifications of the conventional design, such as corrugated tubes, which give better heat transfer than a straight tube while not involving excessive turbulence. This sort of system may be better for fragile foods. Enhanced heat transfer can be obtained by using mixing elements inside the tube, but these significantly increase the pressure drop through the unit.

Evaporators. All the above units operate on liquids or liquid–solid mixtures. Different designs are required where vapour is produced, as in evaporation. Evaporation can be conducted in multitube heat exchangers in which liquid travels either up or down tubes heated by condensing steam on the outside, or inside plate exchangers.

The operation of an evaporator is limited by the temperature difference that is possible between the hot surface and the liquid: if the wall temperature is too high, heavy fouling will result. The increase in viscosity as the solid concentration increases will also lower the heat transfer coefficient. The energy efficiency of evaporators can be increased by using the vapour evaporated in one device as the material that condenses in the next; such multiple-effect systems have a much higher steam efficiency. Two possible configurations of multiple-effect evaporator are shown in Fig. 9.3. In forward feed the hottest steam contacts the least concentrated solution, whereas in backward feed the hottest steam contacts the most concentrated solution. If the product is thermally stable, backward feed has an advantage. The viscosity of the most concentrated fluid will reduce on heating; the heat

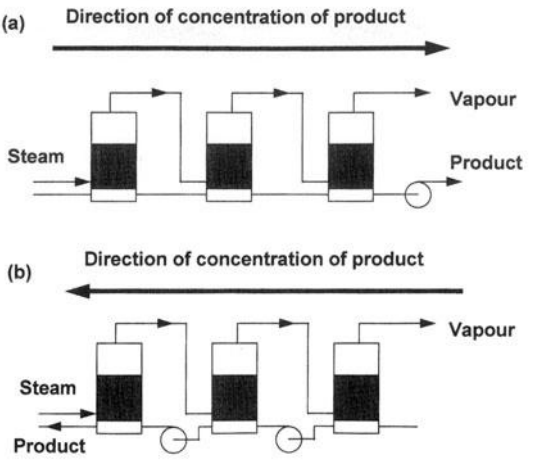


Fig. 9.3 Multiple-effect evaporators: (a) forward feed; (b) backward feed.

transfer coefficient in the final effect will thus be higher than for a cold fluid. For food fluids, however, forward feed is usual, as concentrated fluids will foul heavily at high temperatures.

9.2.2 The processing of solid–liquid food mixtures

The techniques discussed in section 9.1.1 all use conventional heat transfer. The slowness of conduction has already been noted; because of the time taken for conduction, it is impossible for particles greater than about 4 mm to be sterilized in the same time as liquids. The possible commercial advantage offered by the HTST processing of foods containing larger particles, up to the 25 mm commonly used in the home, has stimulated a number of innovations.

Separate processing of solids and liquids. The time required to process large solids and liquids by conductive and convective heat transfer will always be different. One ingenious solution is to treat solids and liquids separately, as in the Jupiter technology developed by APV Baker. Liquid is sterilized in conventional plate exchangers, while solids are processed more slowly in a rotating vessel, which has both steam-heated walls and provision for steam injection. The governing heat transfer processes are thus convection for the liquid phase and conduction for the solid phase. The two sterile phases can then be mixed prior to aseptic packaging. The technique, although elegant, is not an ideal solution, as it requires a large amount of complex pipework.

Heat generation processes. Rather than use conventional heat transfer, a number of techniques exist that exploit heat generation rather than heat transfer. Here, the heat required for sterilization is generated within the food rather than using thermal conduction. Heat generation has the advantage that, if the process is correctly designed, solid and liquid can be heated at the same rates. The heating process is thus much faster than conventional processes, although cooling still depends on heat conduction. Heat generation processes depend on the passage either of an electric field or an electric current through the food material. Heating occurs as a result of friction during molecular rotation, in microwave processing, or by the electrical resistance of the food, in ohmic heating. Unlike conventional processing, where the highest possible temperature is the temperature of the heating medium, there is no theoretical upper limit to the temperature that can be reached by this sort of heating. Long residence times are thus more damaging to product quality than in conventional cooking.

Microwave processing is widespread both in the home and in industry. When high-frequency electric fields are applied to foods, the electric dipoles of water molecules are excited into rapid oscillation. Some of this energy is converted into heat. The heat generation rate per unit volume for heat generation alone is given by

$$Q_G = 0.556E^2\omega\varepsilon' \tan\delta \times 10^{-19} \tag{9.18}$$

where ω is the frequency, E is the electric field strength, ε' is a measure of the number and strength of the dipoles and $\tan\delta$ is the loss tangent. The latter two parameters are functions of the material being heated, and will vary with temperature. Ice has an entirely different response from that of liquid water; as a material thaws the liquid will absorb much more strongly than the ice, causing rapid heating of the liquid portion. Microwave thawing processes must thus be controlled carefully.

The practical efficiency of microwave heating depends on the depth to which the microwaves penetrate into the material. This is a strong function of frequency: as shown by the equation:

$$d = \frac{\lambda}{2\pi\sqrt{\varepsilon_r \tan\delta}} \tag{9.19}$$

where d is the depth where the intensity decays to 1/e of its surface value, λ is the wavelength of the microwave radiation and ε_r the relative permittivity. This equation applies strictly only to the case where microwave energy decays exponentially; more complex behaviour is generally found because of the shape of foods, but the equation is a good guide.

The expression suggests that greater penetration and thus more uniform heating is obtained by low-frequency (high-wavelength) systems. Commer-

cial microwave heating thus uses frequencies in the region of 900 MHz. It is arguable that the main influence of microwave processing on the food industry is home microwaving, and the introduction of meals that can be cooked in the microwave. However, for safety reasons, domestic microwaves use higher frequencies than industrial systems, commonly 2450 MHz, at which the penetration depth is low (commonly of the order of 10 mm). Domestic microwaves thus do not give very uniform heating. In addition, rapid heating to 100 °C is unlikely to sterilize the food. The way in which the consumer uses microwave processing is thus not as safe as conduction cooking; microbial contamination that would be destroyed by half an hour in the oven may survive reheating in the microwave. Microwave ovens have added to the safety problems of the food industry.

Electrical heating is conceptually simple. A schematic diagram of a process is given in Fig. 9.4. A continuous stream of flowing fluid is passed vertically through a tube, which contains a series of electrodes. A voltage is applied between the electrodes; the fluid is rapidly sterilized by heat generated within it due to its electrical resistance. Once sterilized, the material is

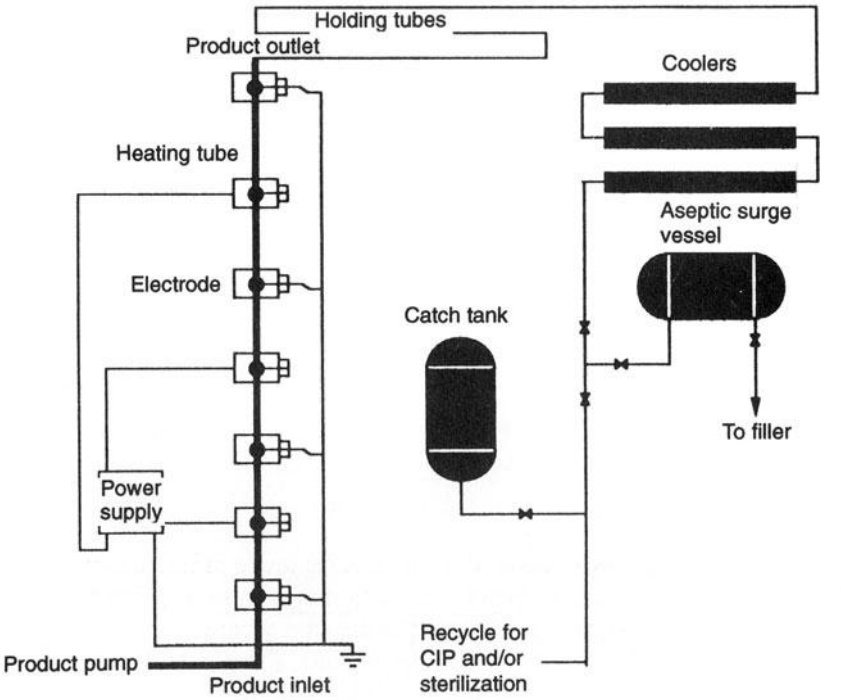


Fig. 9.4 APV Baker 'ohmic heater' unit. Liquid flows through a tube containing a series of electrodes connected to three-phase supply.

cooled and passed to an aseptic packaging system. Such an 'ohmic heating' process has been commercially developed by APV Baker and plants are now operational in the UK.

The process has a greater energy efficiency than microwave heating, as almost all of the electrical power supplied is transformed directly into heat, and it uses low-frequency a.c. (50 or 60 Hz) to eliminate electrolysis. As heat generation rather than heat transfer governs sterilization, irrespective of the particle size, if solid and liquid have identical electrical conductivities, both phases will generate heat at the same rate. In practice, as shown in Fig. 9.5, solids can often heat faster than liquids. This is not possible using conventional techniques.

Electrical heating is unfamiliar by comparison with conventional sterilization processes, and requires new skills of the food technologist and chemical engineer. The controlling factor in the process is the electrical conductivity: both the overall conductivity, which controls the power consumption of the process, and the variation of local conductivity, which affects the local temperature. As with any process, it is necessary to develop operating procedures that assure product safety. The process can be confirmed, in terms both of safety and product quality, by a four-step procedure, as follows.

1. Find the electrical conductivity of all the components as a function of temperature, and measure their heating rates under static conditions as a function of the orientation to the electric field.
2. Thus identify the 'worst-case' particle: the one that heats slowest. The process parameters should be set to sterilize the worst-case particle in the worst orientation.

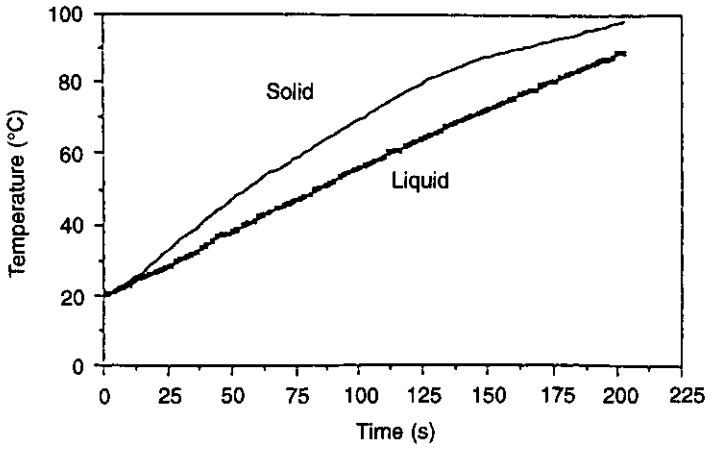


Fig. 9.5 Electrical heating of a potato piece, 3 cm × 4 cm × 0.75 cm, showing particle heating faster than the liquid.

3. Once the process parameters have been determined, perform tests to determine the effect of the process on the fastest-heating component, to check that no unwanted degradation occurs.
4. Also, perform calculations and tests on any component that is known to be thermally fragile.

Once the fastest and slowest cases have been checked and found satisfactory, and provided no intermediate case suffers unacceptable damage, the process can be considered satisfactory. Similar approaches must be adopted for other techniques, such as microwaving, where the dielectric properties of the material, rather than the electrical conductivity alone, control the heating rate.

9.2.3 Flow of food materials

Ideally, all the food material should remain within the process plant for the same time. It is vital to be able to predict the flow properties of foods, but there are two major problems:

- the non-Newtonian rheology, both of food fluids and mixtures;
- the possibility of different behaviour of solid and liquid foods in mixtures.

Although a range of complex flow effects can occur, the techniques used to analyse the flows are straightforward, and are based on the principles outlined in Chapters 2 and 5 on Newtonian and non-Newtonian flows. Turbulent flow, in which the velocity profile is described by the one-seventh power law, offers the nearest approach to ideal plug flow. However, to minimize product damage, solid–liquid mixtures are usually processed in laminar flow, in which a strong velocity profile exists that is dependent on the liquid viscosity. A range of velocities will therefore be expected in commercial plant, giving rise to a range of product qualities and sterilities; this will be especially true in volumetric heating processes, in which the temperature of the food is directly dependent on the time spent in the heater.

The addition of solids to a viscous fluid tends to flatten the velocity profile. In addition, the strong temperature-dependence of viscosity may help the flow of solid–liquid mixtures, as the hot region near the wall may have the lowest viscosity, in which case the central region could move as more of a plug. For efficient operation, however, solids and liquids should flow together, without separation. In practice, particles may be conveyed in plugs (**capsule flow**) or in layers (**heterogeneous** or **saltation flows**). Figure 9.6 shows photographs of flows of carrot particles in water at different solids concentrations, showing the sedimented bed; as the solids fraction increases, the flow becomes more uniform. Processes must be designed to minimize sedimentation: that is, for high-volume solids fractions and vis-

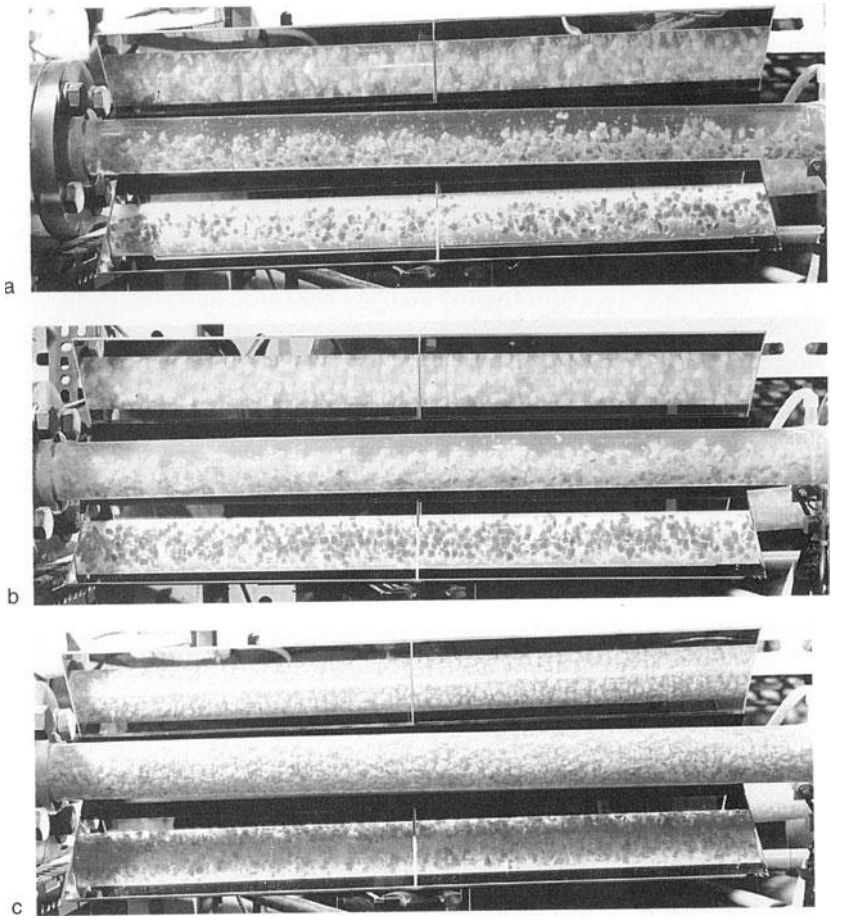


Fig. 9.6 Flow of carrot particles in water at different concentrations of solids: (a) 8%; (b) 14%; (c) 35% (Liu, 1993, PhD thesis, Cambridge University.)

cous carrier fluids. The pressure drop characteristics of such flows are unknown, making it difficult to predict the pressure required to allow the food to flow.

To ensure the microbiological safety of the process, the speed of the fastest-moving particle relative to the average particle velocity must be known. Any process must be designed to sterilize the fastest-moving part of the system while minimizing the cooking of the slowest-moving part. The fastest-moving particle, with the shortest residence time, thus determines the process time, while the particle residence time distribution determines the quality of the final product. Some estimate of the variation in velocity is needed. Applying the technique of dimensional analysis (discussed in sec-

tion 2.4) to the flow of a single particle suggests that eight parameters are involved: particle and mean fluid flow velocities v_p and v_m, particle and pipe diameters d and D, particle and fluid densities ρ_p and ρ_f, the fluid viscosity μ, and the acceleration due to gravity, g:

$$v_p = f(d, D, \rho_p, \rho_f, \mu, v_m, g) \tag{9.20}$$

which, as these involve the three dimensions of mass, length and time, can be rewritten in terms of five dimensionless groups:

$$v_r = f\left(Fr_p, Re, \frac{d}{D}, \rho_r\right) \tag{9.21}$$

Here v_r is the velocity ratio = v_p/v_m (that is, the ratio of particle velocity to mean fluid velocity), Re is the Reynolds number based on pipe diameter, ρ_r is the ratio of particle to fluid density, and Fr_p is the particle **Froude number**, defined as:

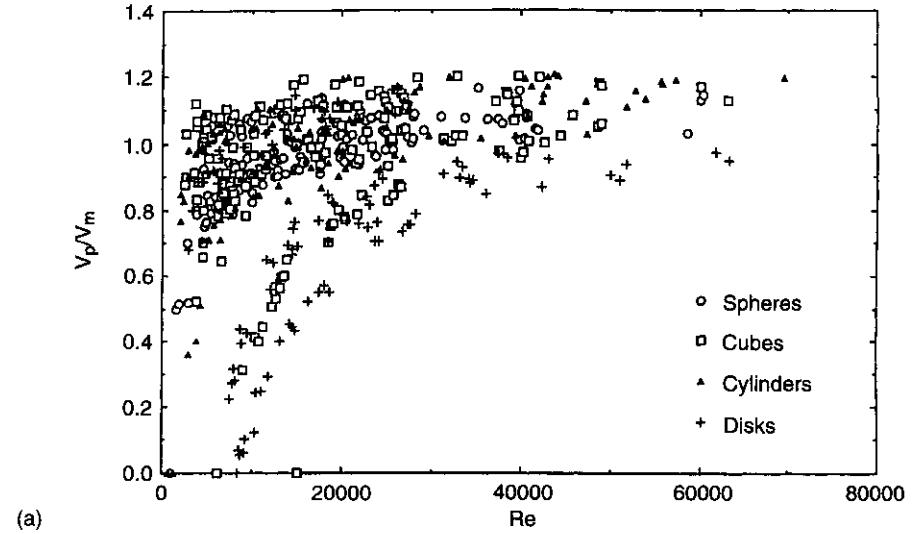
$$Fr_p = \frac{v_m}{\sqrt{gd(\rho_r - 1)}} \tag{9.22}$$

To demonstrate the use of dimensional analysis, Fig. 9.7 shows results for the flow of single particles of different shapes in water, expressed as the variation of v_r as a function both of the particle Froude number defined in equation (9.22) and of the pipe Reynolds number. It can be seen that the Reynolds number is a poor fitting parameter, but when the particle Froude number is used a single curve results. For $Fr_p > 5$ the particle velocity is comparable to that to the fluid, values of v_r between 0.9 and 1.2 being found. For $Fr_p < 5$, however, the mean particle velocity decreases rapidly, until for $Fr_p < 2$ the flow is insufficient to move the particle, leading to a velocity ratio of zero. The empirical equation

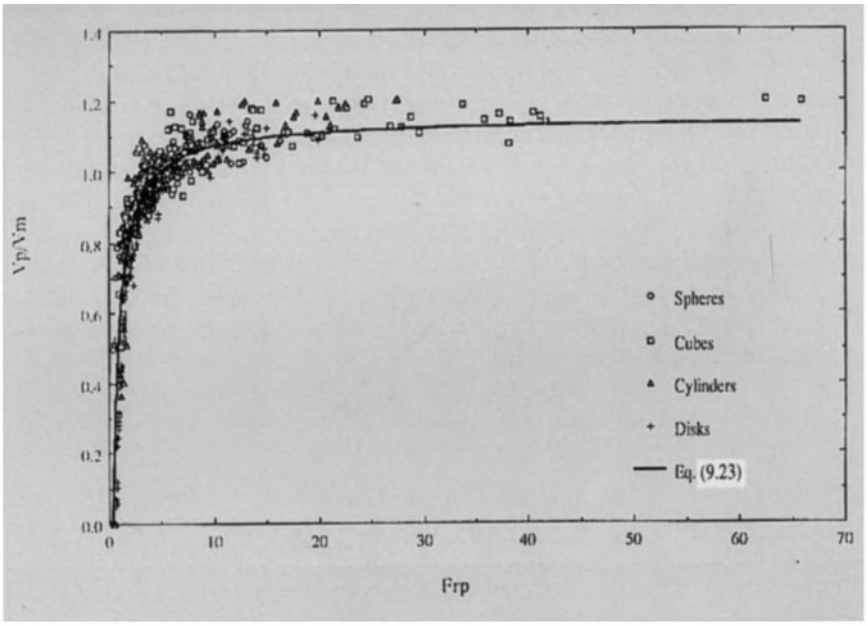
$$v_r = 1.16 - \frac{0.7234}{Fr_p} \tag{9.23}$$

gave a good fit to the data. Similar equations can be found for other sizes of particle and fluids, but the same physical principles apply. In non-Newtonian flow the process is more complex. The velocity profile of a laminar flow is such that the centreline velocity is significantly greater than the mean. If a particle is suspended in the centre of a pipe in a laminar flow it might be expected to travel at a higher v_r than if it were centred in a turbulent flow. The lighter the particle is, therefore, the more likely it is to be suspended at the velocity of the bulk fluid.

The range of velocities possible in a flow can be estimated by following the lightest and heaviest particles. The variation of carrot densities in the flows shown in Fig. 9.6 is between 1010 and 1080 kg m^{-3}. Figure 9.8 shows data for the velocity of two tracer particles of density 1010 and 1040 kg m^{-3}



(a)



(b)

Fig. 9.7 Data for the flow of single particles in water as a function of: (a) tube Reynolds number; (b) particle Froude number.

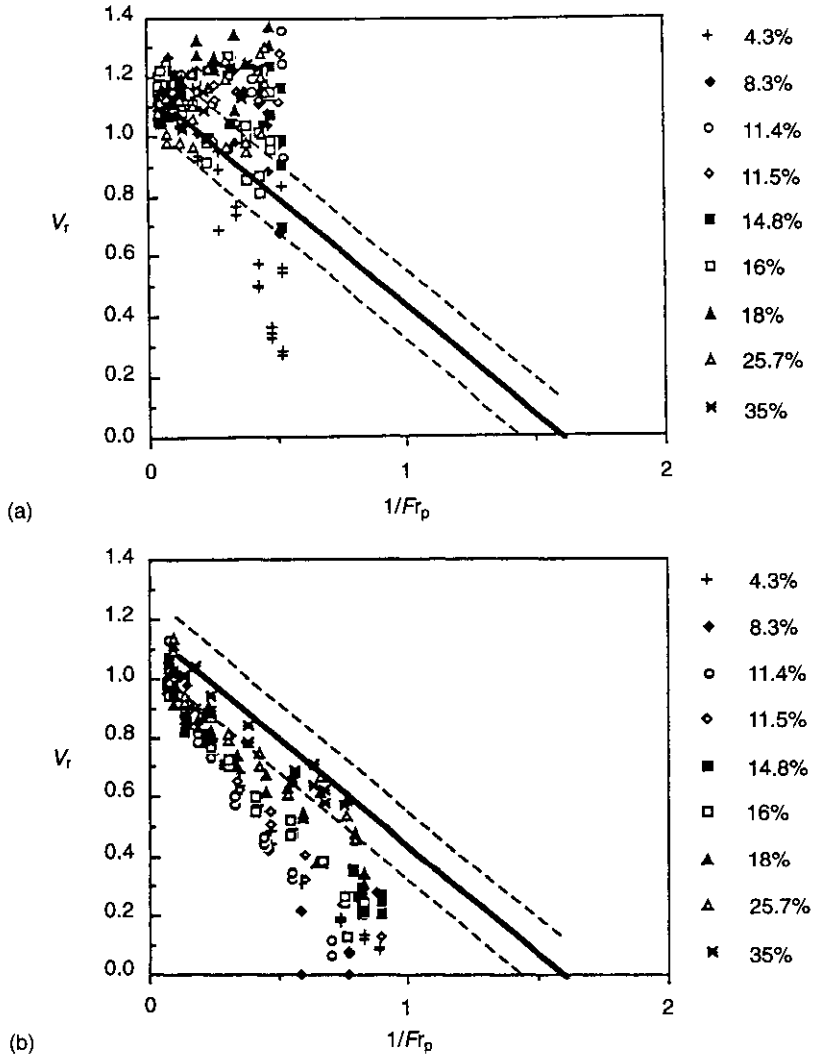


Fig. 9.8 Data for the flow of tracer particles in carrot–water mixtures as a function of solids fraction: (a) tracer specific gravity 1.01; (b) tracer specific gravity 1.04. Black line is equation (9.23); dotted lines are ±10%.

in flows of carrots of various solids fractions in water, together with equation (9.23) for comparison. The lightest particle spent most of the time in the region of low solids fraction at the top of the pipe, and the heaviest spent most time in the sedimented bed. To maintain continuity, if the sedimented bed is travelling at less than the mean velocity then the flow of the liquid above the bed must be faster than the mean velocity; any particles suspended or saltated by the liquid will thus travel significantly faster than

the bed. For the heavier particle, at all solids fractions, the particle velocity is less than that of a single particle, and the data are not widely scattered. Much greater scatter is seen for the lighter particle. The particle can travel up to about 1.5 times the mean flow velocity, well above that of single particles. The highest velocities are found for solids fractions between 10 and 20%. At low solids fraction the effect of the particles is minimal, so that the velocity of the particles can be modelled by the single-particle correlation. However, the depth of the sedimented bed becomes significant for solids fractions greater than about 10%, increasing the liquid phase velocity. At solids fractions above about 20%, the whole pipe is filled by the high solids fraction phase. There is thus no room for a fast-moving stream, and the flow can again be predicted by the single-particle equation. The top of the bed continues to move at higher velocities than the base, but the difference decreases at higher solids fractions. In practice, care must be taken in designing any system that contains a two-phase mixture.

9.2.4 *Control and modelling of heat transfer equipment*

Food plant must be operated safely, and produce a safe product. To do that, it is necessary to ensure that every part of the food has been sterilized, which requires a model for the heat transfer process that can be used to calculate F and C values. The principles used to develop equations for these systems are those of thermal balance and heat transfer, described in earlier sections.

Modelling conduction and convention. Chapter 3 has discussed the basic principles of heat conduction, both steady and unsteady state. The unsteady-state heat conduction equation has been derived in Section 3.1.4. In general, modelling the conduction heating of a three-dimensional body requires the solution of the three-dimensional version of the equation:

$$\rho c_P \frac{\partial T}{\partial t} = k\nabla^2 T = k\left(\frac{\partial^2 T}{\partial x^2} + \frac{\partial^2 T}{\partial y^2} + \frac{\partial^2 T}{\partial z^2}\right) \tag{9.24}$$

at each point in the material, where T is temperature, k is the thermal conductivity, and ρ and c_P are the density and specific heat of the solid, with the requisite boundary conditions, such as

$$q = h\left(T_{\text{liq}} - T_{\text{surf}}\right) = \left(k\nabla T\right)_{\text{surf}} \tag{9.25}$$

that is, external heat transfer to the surface is the same as internal conduction from the surface. At high Biot number, conduction dominates. The rate of heating does not depend on the magnitude of the heat transfer coefficient, and the boundary condition becomes simpler:

$$T_{surf} = T_{liq} \tag{9.26}$$

In practice in the food industry, Biot numbers are often so high that this latter condition applies (Example 9.3), and so it is not possible to increase the heating or cooling rate by increasing the external heat transfer coefficient, for example by increasing the flowrate over the surface.

Equation (9.24), with whichever boundary condition, is difficult to solve. Analytical solutions are rare, and simple ones rarer still; the charts given in Chapter 3 apply to the very limited situation of constant heating temperature and for constant physical properties and simple shapes. It is much more likely that the heating temperature is a function of time, that the physical properties are not constant and that the shape is complex. The simplifications that are necessary to produce a simple equation can be demonstrated by deriving the Plank equation for the freezing time of a solid, already used in Chapter 3.

EXAMPLE 9.8

To derive the equation, it is necessary to assume: (a) the specific heat of the water can be neglected with respect to latent heat, i.e. all the heat supplied goes to freeze water; (b) heat transfer is one-dimensional, through a layer of thickness x *from a medium of cooling temperature* T_b *to a freezing front at* T_f, *where* $\Delta T = T_f - T_b$; *(c) the physical properties are constant. The rate of convective heat transfer from the solid will equal that due to conduction from the freezing front:*

$$q = h(T_i - T_b) = \frac{\lambda}{x}(T_f - T_i) \tag{9.27}$$

where T_i is the solid–coolant interface temperature and the interfacial heat transfer coefficient is h. This heat is evolved by latent heat due to freezing:

$$q = \theta_w h_{fg} \frac{dx}{dt} \tag{9.28}$$

where h_{fg} is the latent heat of water and θ_w is the fraction of water in the material. Eliminating the unknown temperature gives

$$q = \theta_w h_{fg} \frac{dx}{dt} = \Delta T \left(\frac{1}{h} + \frac{x}{\lambda} \right)^{-1} \tag{9.29}$$

which, when integrated, gives

$$t = \frac{h_{fg}\theta_w}{\Delta T} \left[\frac{x}{h} + \frac{x^2}{2\lambda} \right] \tag{9.30}$$

from which the freezing time can be estimated, as in Chapter 3.

The equation derived in Example 9.8 is useful only for a first estimate. Solutions for heat transfer where changes in temperature occur throughout the material have been discussed in Chapter 3, and are best expressed graphically. More accurate models are needed to cope with different shapes and the complex variation of physical properties with temperature found in real systems. Numerical models are thus used, in which the differential equation (9.24) is approximated at a series of points or regions. The numerical accuracy of the techniques depends on the spacing of these points: the more points, the more accurate the calculation, but the more the computing power that is needed. It is also vital to be able to predict the physical properties of the system. Thermal properties (density, specific heat, thermal conductivity) of most common foods are now reasonably well documented.

For the control of retort processes, much work has gone into the development of mathematical models of heat penetration into a can or retortable pouch. Such models solve the conduction equation to calculate the temperature, and hence the F-value, at the centre of the package, using the monitored temperature–time profile in the retort. Problems arise in modelling the effect of the headspace, for example between the food and the lid in retortable trays; here, the interfacial heat transfer coefficient is more important than it is in slow-moving liquid–liquid flows. These models enable the plant operator to determine when the cans in a retort are sterile, and minimize process losses.

Modelling and control of HTST processes. Control of a continuous aseptic HTST process requires a different approach from the above. Canning effectively requires the measurement of one temperature as a function of time. In continuous HTST processes, a number of temperatures must be measured continuously to ensure that every part of the plant maintains sterility and that cooling is effective; at the very least, the start and finish of the heating, holding, and cooling sections should be monitored. In addition, the flowrate and the pressure in the system must be monitored: if pressure falls too low in the heating section, the product may begin to boil, and if it falls too low at the filler, insufficient pressure will be available to enable filling.

The sterilizing effect of the heating regime on the food must be calculated. For a single-phase fluid the heat balances are as given in Chapter 3, but a two-phase fluid requires the use of an enthalpy balance on both solid and liquid phases. For each particle in a fluid, the interchange between the particle and the fluid is given by equation (9.25). The balance on the liquid side is more complex, as there could be i particles exchanging heat with the liquid, each with its own physical properties and velocities. Assuming (a) a heat transfer coefficient of h_w between wall and fluid, (b) a fully stirred liquid, and (c) a constant velocity v_L for the whole fluid, then the liquid phase heat balance on a length dx of tube of diameter D becomes

Heat transfer from wall + Heat transfer from i particles = Heat change in fluid

$$\pi D \mathrm{d}x.h_w\left(T_w - T_L\right) + \sum_i h_i a_i \frac{\pi D^2}{4} dx\left(T_{si}^0 - T_L\right)$$

$$= \left(1-\phi\right)\frac{\pi D^2}{4}\mathrm{d}x\left(\rho c_P\right)_L \frac{\mathrm{d}T_L}{\mathrm{d}t} \tag{9.31}$$

where a_i is the surface area of the ith particle per unit volume and ϕ is the total solid fraction occupied by the particles. The accuracy of the assumption that the liquid is well stirred is not known; it is likely to be true for turbulent fluids. The length of the holding tube must be calculated using a model that calculates the temperature distribution in the particle as a function of the bulk temperature, and determines when it is sterile.

Once the holding time is known, the velocity of the food through the hold tube must be calculated. This requires an understanding of the material RTD. If the RTD is not known, very conservative estimates must be used, such as the assumption of fully developed Newtonian flow. One ingenious technique for determining the sterilization given to foods by a process has been developed by Campden Food RA. Particles of sodium alginate containing spores of *B. stereothermophilus* are passed through the heater, and their activity before and after heating is compared, allowing the F-value of the process to be calculated directly. This allows direct confirmation of the effects of the process.

Modelling heat generation. Heat transfer processes require knowledge of physical properties such as the thermal conductivity of the solid and liquid phases, specific heats, densities, and (in some cases, although conduction normally dominates in liquid–solid processing) interphase heat transfer coefficients. Other factors control heat generation: in electrical heating it is vital to know electrical conductivity; in microwave heating the dielectric properties of foods.

To illustrate the difficulties of measuring physical properties, the following was found during an investigation of food electrical conductivity.

- It can be a strong function of frequency. As commercial conductivity meters work at much higher frequencies than the 50 Hz used in ohmic heating, errors can result unless measurements are made at 50 Hz.
- It can be anisotropic, presumably reflecting the structure of the material. For example, the conductivity of carrot at 25 °C is 0.25 mS cm^{-1} across the axis and 0.42 mS cm^{-1} parallel to the axis.
- Conductivity–temperature profiles for some foods can differ significantly between ohmic and conventional heating.

In electrical processing, the heat generation rate at any point is given by $Q_G = \kappa E^2$, where E is the local field strength and κ is the local electrical conductivity. In microwave heating, the heat generation rate is given by

equation (9.19). The heat generation term must be combined with the conduction equation to give

$$\rho c_P \frac{dT}{dt} = Q_G + k\nabla^2 T = \kappa E^2 + k\nabla^2 T \tag{9.32}$$

For electrical heating, if the local electric field strength and the physical properties of the material are known, the heating rate can be predicted. The field distribution must be calculated to find E. This requires the solution of Laplace's equation for voltage V:

$$\nabla(\kappa \nabla V) = 0$$

which reduces to

$$\nabla^2 V = 0 \tag{9.33}$$

if electrical conductivity is constant. Maxwell's equations must be solved for microwave heating; this will not be dealt with here. The voltage distribution in electrical heating will depend on the distribution of electrical conductivities within the fluid, and thus on particle shape and its orientation.

If the physical properties of the medium are not uniform, temperature differences can result. This is clearly demonstrated in Fig. 9.9, which shows

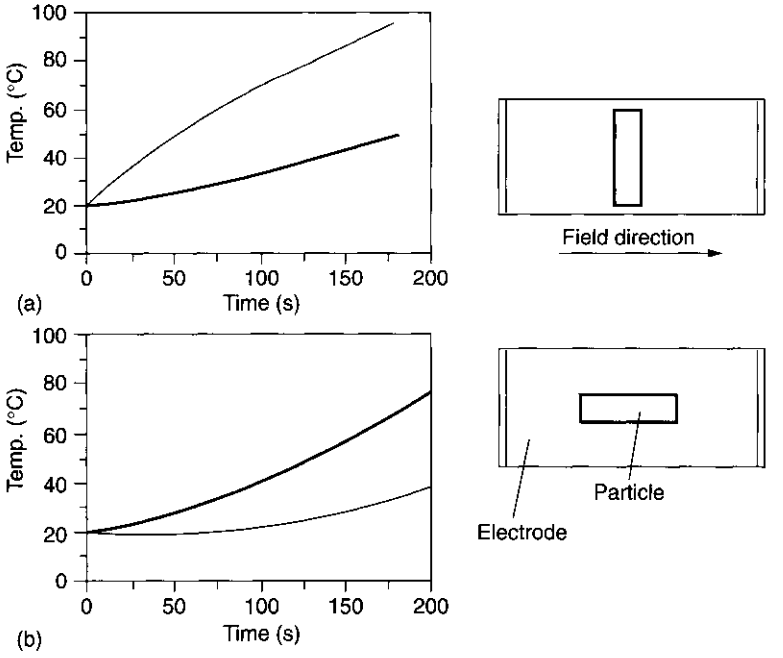


Fig. 9.9 The change in heating rate with particle orientation for potato particles undergoing electrical heating: (a) perpendicular, (b) parallel to the field. Heavy line shows the liquid temperature.

the heating of two identically shaped potato slices in brine, one perpendicular and one parallel to the field. The difference in heating rate between the two is readily seen; when the particle is perpendicular to the field it heats faster than the liquid, but when parallel it heats slower. In practice, this will not matter if the process is designed so that (a) a particle that travels through the heater in the slowest-heating configuration is sterile, and (b) a particle that travels through in the fastest-heating configuration is not unacceptably overcooked: that is, that the system operates between the limits of Fig. 9.3. It is vital to calculate the heating rate to determine the differences that may result.

Calculation of the electrical heat generation rate is difficult and requires either experiment or complex numerical modelling. If it is assumed that the heat generation rate in the liquid and solid phases is uniform and equal to Q_L and Q_S, however, a simple model is possible. As heat generation is much faster than heat transfer it is possible to assume that the particle temperatures are uniform. For a solid particle of temperature T_S moving at velocity v_S in a fully developed flow of temperature T_L the heat balance can be written as

$$-ha\left(T_S - T_L\right) + Q_S = v_S\left(\rho c_P\right)_S \frac{dT_S}{dx} \tag{9.34}$$

where h is the convective heat transfer coefficient, and a is the area of the particle per unit volume ($6/d_p$ for a sphere of diameter d_p). The balance on the liquid side is more complex, as there could be i particles exchanging heat with the liquid, each with its own physical properties and velocities. In the holding section, the liquid phase heat balance becomes

$$\frac{\sum_i \phi_i h_i a_i \left(T_{Si} - T_L\right)}{\left(1 - \sum_i \phi_i\right)} + Q_L = v_L\left(\rho c_P\right)_L \frac{dT_L}{dx} \tag{9.35}$$

where ϕ_i is the fraction of the volume of the system occupied by solid i. If all the particles are travelling with the same velocity, this can be rewritten

$$\frac{\phi ha\left(T_S - T_L\right)}{\left(1-\phi\right)} + Q_L = v_L\left(\rho c_P\right)_L \frac{dT_L}{dx} \tag{9.36}$$

Manipulation gives

$$\begin{aligned} -H_S\left(T_S - T_L\right) + G_S &= v_S \frac{dT_S}{dx} \\ H_L\left(T_S - T_L\right) + G_L &= v_L \frac{dT_L}{dx} \end{aligned} \tag{9.37}$$

The equations can thus be written in terms of G, the inherent heat generation in each phase together with two modified heat transfer terms, $H_S = ha/(\rho c_P)_S$ and $H_L = [\phi/(1-\phi)][ha/(\rho c_P)_L]$ and the phase velocities.

Equation (9.37) can be solved numerically, but an analytical solution is available if all the factors in the equation are constants. The difference between the two phase temperatures is then found to be

$$\Delta T = T_S - T_L = \Delta T_\infty + \left(\Delta T_0 - \Delta T_\infty\right)\exp\left(-\beta x\right) \tag{9.38}$$

where

$$\begin{aligned} \alpha &= \frac{G_S}{\nu_S} - \frac{G_L}{\nu_L} \\ \beta &= \frac{H_L}{\nu_L} + \frac{H_S}{\nu_S} \\ \Delta T_0 &= T_S^0 - T_L^0 \end{aligned} \tag{9.39}$$

and the final temperature difference between the two phases, ΔT_∞, is α/β. Individual phase temperatures can then be calculated using the following equations:

$$T_S = T_S^0 + \frac{H_S}{\beta \nu_S}\left(\Delta T_0 - \Delta T_\infty\right)\left(\exp\left(-\beta x\right) - 1\right) + \left(\frac{G_S}{\nu_S} - \Delta T_\infty \frac{H_S}{\nu_S}\right)x \tag{9.40}$$

and

$$T_L = T_S - \left(\Delta T_\infty + \left(\Delta T_0 - \Delta T_\infty\right)\exp\left(-\beta x\right)\right) \tag{9.41}$$

Using these equations to find temperatures, F and C for each phase can then be found by integrating equations (9.2) and (9.3) numerically. Figure 9.10 shows the type of temperature data which can be produced by the model; here the solid first underheats and then overheats the liquid (Zhang and Fryer, 1994).

As with all analytical solutions, the applicability of the equations is limited; however, they can be used to see how a real system could behave. The equations show that the thermal response of the two phases is in two parts; a term that is linear in distance along the heater, and is governed largely by the rate of heat generation; and an exponential heat transfer term that includes β, and thus takes account of the magnitude of h. The rate at which ΔT_∞ is approached depends on the heat transfer coefficient.

Heat transfer models of this sort are essential for process design and control; although the equation can be difficult to solve numerically, the physical principles are straightforward.

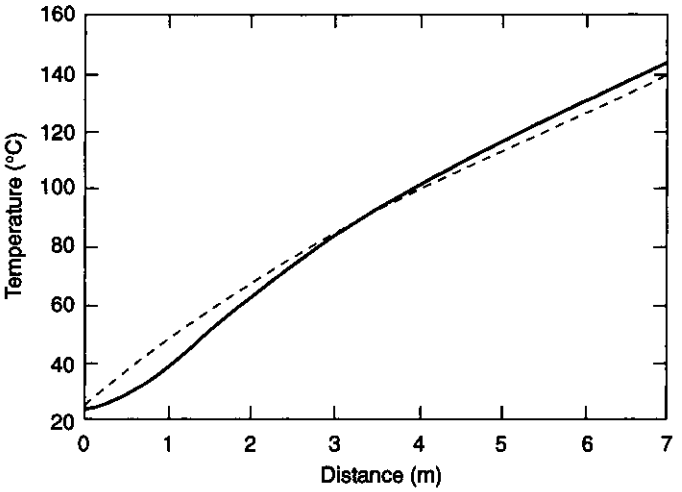


Fig. 9.10 Variation in temperature of a solid (—) and a liquid (---) using the model of (9.34)–(9.41).

9.3 Fouling and cleaning in food process plant

9.3.1 Introduction: the fouling problem

The operation of food processing plant is complicated by the thermal instability of food material. During operation, the inner surfaces of food plant gradually become covered with a solid fouling deposit. Deposition is most rapid in heating equipment such as heat exchangers and evaporators, but can occur elsewhere, for example in high-temperature holding tubes. Fouling causes a number of problems, as follows.

- By adding an insulating solid layer to the heat transfer surface, it creates a barrier to heat transfer.
- By decreasing the area available to flow and changing the roughness of the surface, it increases pressure drop.
- By providing areas in which microbes can adhere and survive, it threatens plant sterility.

Fouling is common in the process industries, but food fouling is especially severe: for example, it is possible to operate an oil refinery for several months between cleanings, but food plant commonly has to be cleaned daily. Food fouling has been solved empirically by the food industry. Many of the types of heat exchanger used by the food industry, such as the scraped-surface exchanger, have an antifouling action that allows run times to be extended before excessive deposit build-up is reached. In addition, it

is necessary to clean process plant regularly, and cleaning-in-place (CIP) equipment is widely used to ensure that this is done efficiently. The aim of this section is to describe fouling and cleaning problems and how they can be reduced.

The effect of fouling is included in the basic heat transfer equation by the inclusion of the **fouling resistance** (or **fouling factor**) R_F:

$$\frac{1}{U} = \frac{1}{U^0} + R_F \tag{9.42}$$

where U and U^0 are the overall heat transfer coefficients in the presence and absence of fouling respectively. Although U^0 may be calculated fairly accurately from correlations, such as those described in Chapter 3, R_F cannot; this makes heat transfer equipment difficult to design.

In addition, fouling is a transient process; the surface starts clean and ends up fouled in a number of ways, as shown in Fig. 9.11. Before fouling there may be an induction period during which heat transfer and pressure drop change only slightly; indeed, the extra roughness given to the surface by the first layer of fouling can increase the heat transfer coefficient for a short time. At the end of this period the surface fouls rapidly; the fouling rate may slow to produce a final equilibrium deposit, but this will probably occur at too high a value to be acceptable in industrial plant. The transient nature of fouling causes problems, because equipment oversized to cope with a high fouling resistance may overprocess food when it is clean; control systems must cope with the changes that occur during a process run.

Mathematically, the rate of fouling is commonly expressed as a balance between deposition and removal processes:

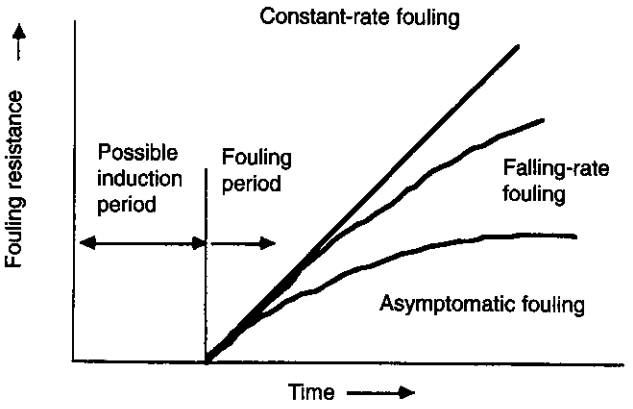


Fig. 9.11 The types of fouling curve found in practice.

$$\frac{dR_f}{dt} = \phi_d - \phi_r \tag{9.43}$$

To operate process plant it is obviously important to be able to predict the rate and severity of fouling in a given environment: that is, the deposition and removal rates ϕ_d and ϕ_r in the above equation. Most research in the food area has been done on milk, a model system of relevance to many areas of food processing; much less information is available on other systems of industrial importance. Even for milk, results are sometimes contradictory, and milk fouling has been found to vary with the time of the year, the age and pH of the milk and the lactation stage of the cow, as well as with the pretreatment that the milk has received.

Some basic characteristics of the fouling process can be inferred from observation of industrial plant. That fouling is **temperature**- and **concentration-dependent** is demonstrated by multiple-effect evaporators: fouling is concentrated in the first effect, where the temperature is highest, and in the last, where the liquid is most concentrated. **Velocity dependence** is shown by individual tubes in multitube evaporators, which can clog solid with deposit. If small amounts of deposit form in an individual tube, the flow through that tube will be reduced, resulting in increased deposition and eventually in total blockage. Multitube evaporators thus foul by having most of the tubes clean and a few blocked solid.

The study of fouling has a number of aims: to interpret experimental data in terms of basic mechanisms; to develop operating procedures that minimize fouling; and to develop new types of heat exchanger. Some model for the reactions that give rise to fouling is required. It is common in process engineering to consider the sequence of steps that give rise to a final effect, such as in the construction of an overall heat transfer coefficient from individual film coefficients. The end result of fouling is the deposition of solids on the heat transfer surface. Fouling from foods may result from a combination of diffusion and reaction steps. The material that becomes fouling deposit will either be generated in the bulk and then transferred to the surface, or its precursor will be transported from the bulk to the surface, where reactions that produce the deposit will take place. The process will be at least a two-stage one: diffusion to the surface followed by reaction on it, as shown in Fig. 9.12. The flux of foulant N will be given both by

$$N = k(c_b - c_i) \tag{9.44}$$

where k is a mass transfer coefficient and c_b and c_i are the bulk and interfacial concentrations of foulant respectively, and by the surface reaction rate

$$N = k_r c_i \tag{9.45}$$

if the reaction is first order, where k_r is a reaction rate constant (a strong Arrhenius function of temperature, not a function of flow conditions). The

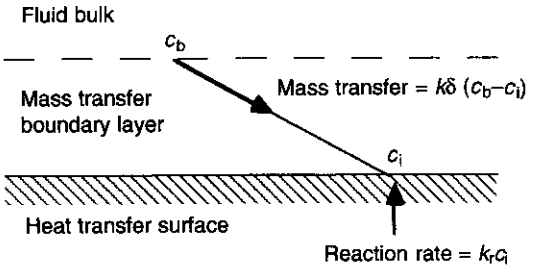


Fig. 9.12 Schematic diagram of consecutive mass transfer and reaction steps in fouling.

overall rate can be found by combination of the two equations to eliminate the unknown interfacial concentration:

$$N = \frac{c_b}{1/k_r + 1/k} \tag{9.46}$$

Depending on the relative magnitude of the mass transfer and reaction coefficients, fouling can be controlled either by mass transfer or by surface reaction. The mass transfer coefficient will be a strong function of the flow conditions, but is less affected by temperature, while the rate of any reaction will be an Arrhenius function of temperature but not a function of the flowrate. Although this analysis is a simple one, it shows how the net rate of fouling may thus depend on either or both of the temperature or flowrate.

9.3.2 The chemistry of fouling

The fouling from milk can be used as an example of food fouling processes. Reflecting its industrial importance, fouling from milk fluids has been studied by a number of workers. Two types of deposit from milk fluids are found, as follows.

- **Type A** or **milk film**: found at temperatures below 110°C, this deposit is creamy and white, and consists of 50–60% protein and 30–35% minerals.
- **Type B** or **milk stone**: found at temperatures above 110°C, this consists of 15–20% protein and up to 70% minerals.

Deposition results from the degradation of thermally unstable components of the fluid, milk proteins and calcium phosphate. Calcium phosphate becomes less soluble with increasing temperature and thus will precipitate out onto heated surfaces to form a mineral scale. On heating whey proteins, denaturation and aggregation reactions can occur. Denaturation describes the unfolding of the complex three-dimensional shape of the protein chain,

which thus loses the activity that it normally possesses as a result of its shape. This process may be reversible; however, if reactive groups are exposed in denaturation, individual strands can polymerize in an irreversible reaction to give insoluble aggregates. Figure 9.13 shows an electron micrograph of a section of deposit, which is composed of a mixture of deposited mineral salts and aggregated proteins. The most thermally labile milk protein, β-lactoglobulin, makes up only 10% of raw milk protein but up to half of the protein content of type A deposit. At higher temperatures, a similar process occurs. In type B deposit, the amount of calcium phosphate in the deposit is higher because the saturation concentration is lower, and different proteins deposit, including some casein fragments.

The thermal behaviour of β-lactoglobulin is complex; it is shown schematically in Fig. 9.14. On heating to about 70 °C, the protein structure partially unfolds in molecular denaturation. This exposes reactive

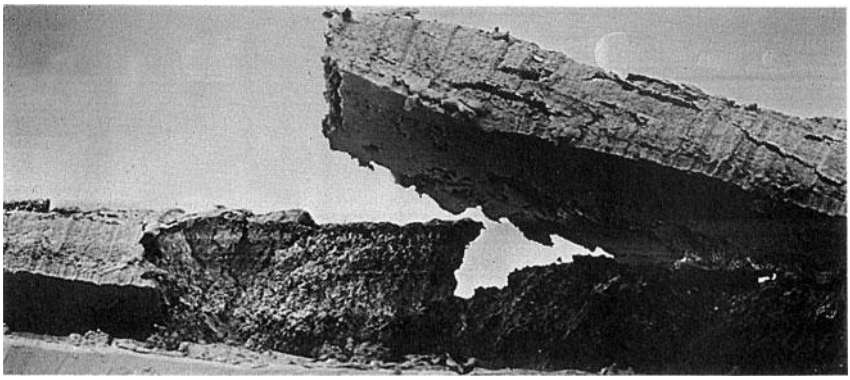


Fig. 9.13 SEM image of a fouling deposit. (Reproduced from Belmar-Beiny and Fryer, 1993.)

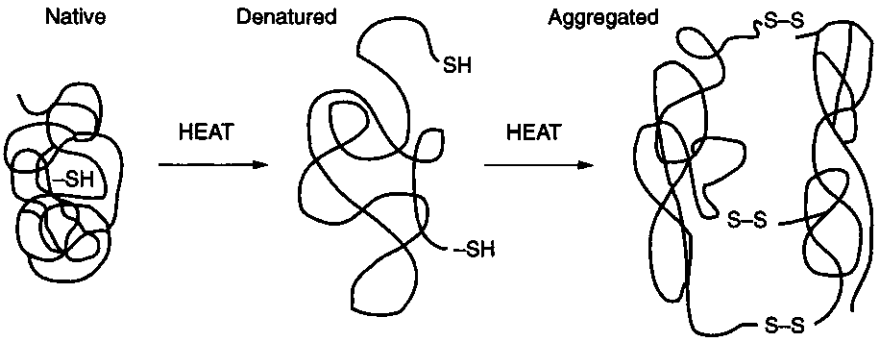


Fig. 9.14 The thermal behaviour of β-lactoglobulin.

sulphydryl groups, which are normally concealed within the core of the protein. Groups in different denatured molecules can then react to form large polymerized and eventually insoluble protein aggregates; this process happens rapidly above about 74 °C. The link between protein denaturation and aggregation and fouling is well established; addition of a sulphydryl oxidizing agent, which makes it impossible for aggregates to form, reduces the amount of deposit formed.

The change in fouling with pH suggests that protein aggregation, rather than denaturation, is the controlling process in type A fouling. Denaturation of β-lactoglobulin increases above pH 6.25, but fouling in pasteurization and UHT plants increases rapidly below pH 6.6. If β-lactoglobulin denaturation were the limiting step, fouling should increase at higher pH, but the reverse is true. However, β-lactoglobulin aggregation increases significantly below pH 6.0, reflecting increased molecular unfolding during denaturation at this pH, resulting in increased disulphide exchange reactions. Although the **rate** of β-lactoglobulin denaturation increases above pH 6.5, the amount of molecular unfolding associated with denaturation is reduced, decreasing the concentration of free sulphydryl groups. This prevents intermolecular disulphide exchange reactions, and thus aggregation. Below pH 6.5, although the denaturation rate is slow, it results in a high concentration of sulphydryl groups, and rapid aggregation. Although denaturation is necessary for fouling to occur, the stage which directly leads to the formation of deposit thus appears to be the formation of insoluble aggregates.

Protein aggregates will be formed wherever the temperature is hot enough, and yet induction periods of several hours can be found in industrial plant before the effect of fouling is noticed. Some of this is due to the fact that the fouling resistance will be initially small and will make little change in the overall heat transfer coefficient, but it is likely that some conditioning of the surface must occur before heavy fouling can take place.

Experimental results thus imply that the fouling process is two-stage.

1. The **induction period**, during which little change in heat transfer coefficient takes place, corresponds to the time taken to condition the surface or a region of it so that formation of heavy deposit can begin.
2. The **fouling period** starts when the surface is conditioned so that rapid fouling can begin.

This model reflects the types of deposit formed in commercial milk processing. Different deposits will be found in different situations. When faced with a fouling problem, analysis of the deposit should first be conducted, to determine which components of the food material are causing the problem. The next step is to determine where the processes which result in fouling are taking place.

9.3.3 Fouling as a problem in reaction engineering

Section 9.3.2 has shown that the chemistry of the fouling process is fairly well understood. To the process engineer, an understanding of the chemistry of the process is insufficient; it is necessary to know where the controlling reaction takes place, and what can be done to reduce its effects. It is not easy to determine basic mechanisms from experiments on industrial plant. Practical equipment such as plate heat exchangers contains a range of different temperatures and surface shear stresses, which will give different local fouling rates: so the fouling resistance measured in such plant will be a composite of these different rates. This section describes some laboratory experiments to elucidate fouling mechanisms.

Induction period. If the processes that control the first layer of deposition are known, it might be possible to modify the surface to resist fouling. When the final deposit, of the type of Fig. 9.13, is analysed, a thin film of calcium phosphate (less than 20 μm) can be seen next to the heat transfer surface. On top of this layer is a much thicker layer of protein aggregates, sometimes clustered around crystalline protrusions from the mineral layer. This might suggest that minerals are the first species to be deposited. However, if experiments are carried out over much shorter periods, it can be seen that proteins are the first layer to be absorbed. Figure 9.15 shows the XPS spectra obtained for unfouled and fouled stainless steel with a contact time of 40 s. XPS (X-ray photoelection spectroscopy) is a technique that measures the presence of particular atoms. Figure 9.15(a) gives the basic spectrum of the steel; after contact with fluid for only 40 s, no stainless steel peaks remain, showing that the surface has become covered. No peaks corresponding to calcium or phosphorus can be found; the surface becomes covered with a smooth layer of protein. Electron micrographs of the surface show that protein aggregates are not seen until after a longer period, as in Fig. 9.16. It seems likely that the layer of calcium phosphate found on the surface after extended periods of time is caused by diffusion of the material through the deposit after it has formed, i.e. that ageing of the deposit changes its local composition.

Fouling period. From the two-stage model above it is clear that the rate of fouling could be controlled by diffusion or wall processes. If it is assumed that deposition results from a combination of mass transfer and chemical reactions, in any situation one of these processes – the slowest – will be the rate-controlling step. Using the simple ideas developed in earlier chapters, we can develop a possible mechanism for fouling.

Fouling may be **mass-transfer controlled**: that is, the transfer of reacted protein to the wall may be the slowest step. Here, deposit formation will not be a strong function of temperature. However, if the process is **reaction-**

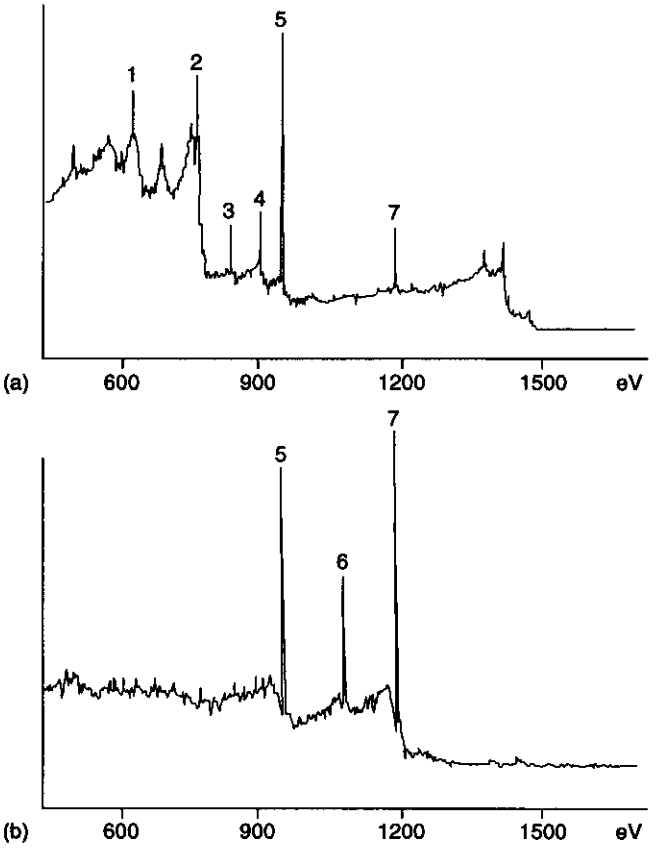


Fig. 9.15 XPS spectra on stainless steel surfaces (AISI 321): (a) clean surface; (b) surface fouled with whey protein concentrate for 4 s. Temperatures: inlet, 73 °C; outlet, 75 °C; wall 96 °C. Main characteristic peaks: 1, Fe_{Auger}; 2, Fe_{2p}; 3, Ni_{Auger}; 4, Cr_{2p}; 5, O_{1s}; 6, N_{1s}; 7, C_{1s}; 8, S_{2p}. (Reproduced from Belmar-Beiny and Fryer, 1993.)

controlled, deposition will be a function of wall or bulk temperature, depending on where the controlling reaction takes place. Chapter 3 has described the film model of a turbulent flow, in which the rate of heat transfer is modelled by conduction through a fluid thermal boundary layer that is near the wall temperature. Any reaction controlling it could take place in two possible places and in three different ways, as follows.

1. **Surface reaction**. If fouling is controlled only by surface processes, deposition will occur wherever the surface temperature is high enough for protein denaturation and aggregation to occur. The fouling rate will be a function of surface rather than bulk temperature.
2. **Bulk reaction**. If the controlling reaction for fouling takes place in the fluid bulk, then two cases can be envisaged.

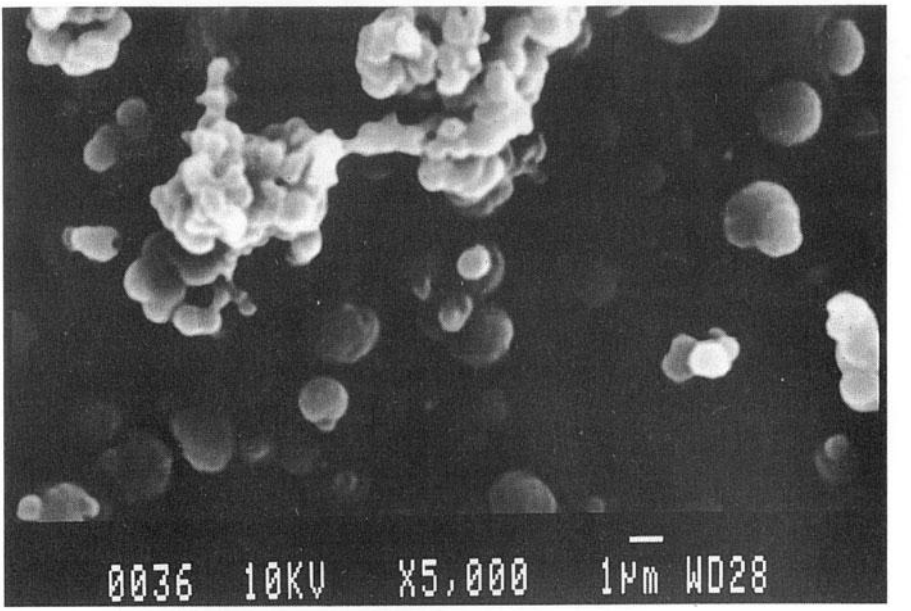


Fig. 9.16 SEM image of the beginning of the adhesion of protein aggregates. (Reproduced from Belmar-Beiny and Fryer, 1993.)

(a) If the wall and bulk temperatures are such that protein denaturation and aggregation will occur at the wall but not in the bulk, fouling will only result from deposition of protein that has denatured and aggregated in the **thermal boundary layer** adjacent to the wall.
(b) If both the boundary layer and the **turbulent core** are hot enough for protein denaturation and aggregation, protein denatured and aggregated in both regions will contribute to deposit formation.

If a surface reaction is responsible for fouling, the amount should depend only on the wall temperature. If bulk processes contribute, then the amount of fouling should increase when the fluid bulk becomes hot enough to produce denatured and aggregated protein.

Figure 9.17 shows the result of experiments showing the initial rate of fouling for a system corresponding to case 2(a). The rate of fouling is temperature-dependent, showing an activation energy of about 90 kJ mol^{-1}, but decreases with increasing *Re*. The rate law for the initial rate of fouling was correlated as

$$\frac{dR_F}{dt} = \frac{k}{Re}\exp\left(\frac{-E}{RT_w}\right) \tag{9.47}$$

where *Re* is the fluid Reynolds number and T_w is the wall temperature. An equation of this form can be obtained if the variation of the thickness of the

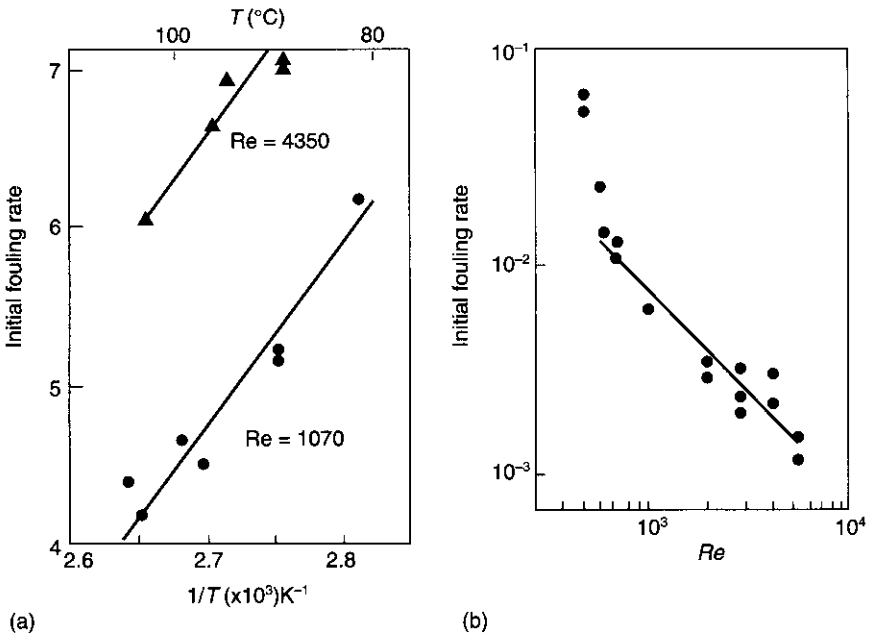


Fig. 9.17 (a) Temperature and (b) flow dependence of the initial rate of fouling for a wall temperature of about 90 °C and bulk temperature constantly below 65 °C. Fouling rate is d*Bi*/d*t*; dimensionless.

thermal boundary layer of the fluid is considered as a reactor at the temperature of the heat transfer surface. Further evidence for the involvement of bulk processes comes from experiments such as that shown in Fig. 9.18, where the wall temperature is constant and above the aggregation temperature, and the bulk fluid is below the temperature at inlet and above it at outlet. When the bulk temperature exceeds the point where aggregation occurs significantly, an increase in deposition results.

Although the final step in the fouling process is the adhesion of aggregates to the wall, the fouling rate is critically influenced by the generation of denatured and aggregated protein in the bulk of the fluid: that is, c_b in equation (9.46) is the concentration of reacted rather than native protein. In many industrial situations the temperature difference between the wall and the bulk is small and this effect might be obscured.

9.3.4 Implications of the fouling model

The above experiments have demonstrated the controlling mechanisms. Fouling results from a series of processes:

1. deposition of proteins on the clean surface to give an initial layer; when nucleation sites are available,

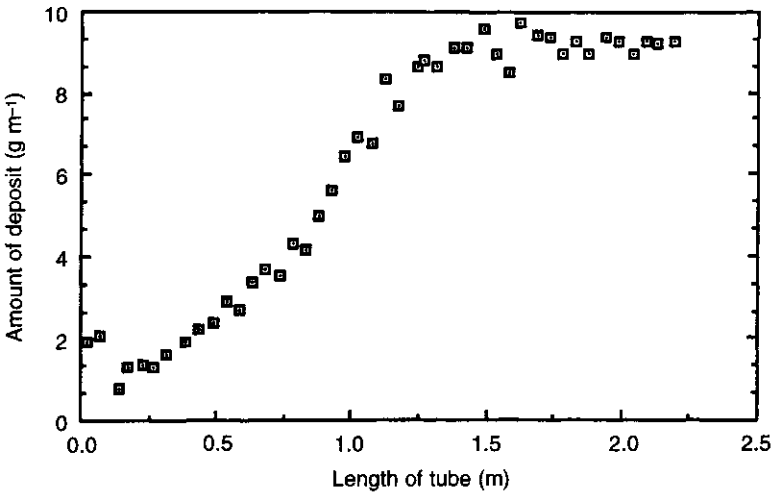


Fig. 9.18 Fouling from 1% protein: Re = 7500; protein inlet 73 °C and outlet 83 °C; oil inlet 97 °C and outlet 95 °C. Fluid enters below the temperature at which β-lactoglobulin thermal instability is significant, and leaves above that temperature.

2. deposition of proteins that have reacted in the hot region of the fluid, either the bulk of the fluid or the wall layer, can occur.

The mechanistic model can be used to examine industrial heat exchangers. Equipment should be designed to reduce adhesion, avoid high temperatures, and minimize surface fluid residence times.

- **Reduce adhesion**. Fouling is reduced by high surface shear stresses and smooth surfaces, which do not contain crevices in which deposition can start. Good hygienic design (that is, providing no points where microbial contamination can occur) will thus reduce fouling. Once the heat transfer surface is covered with a layer of protein, fouling on different types of surfaces will be the same; this is found in practice.
- **Avoid high temperatures**. The formation of denatured and aggregated protein depends only on the fluid temperature, but salt deposition depends on the difference between wall and bulk temperatures: that is, on the amount by which the salt is supersaturated at the surface. Fouling thus depends on both T_w and T_b; low temperature differences between bulk and wall will reduce salt formation, and low temperatures will reduce protein aggregation.
- **Minimize surface fluid residence times**. Surface bonding of proteins to the wall will take a finite time. If the residence time of fluid at the wall is kept low, by high turbulence or good mixing, the chance of adhesion will be reduced. However, if the surface reaction is rapid, and mass transfer controls deposition, increasing mixing may enhance fouling by increasing k_m in equation (9.46). Some plate heat exchangers contain points of

contact between plates; these will be low-shear regions, and are generally the points at which fouling begins.

Fouling models can also be used in the design of process plant. To do this, a computer model of the flow and temperature profiles in the heat exchanger is needed, together with a model for the kinetics of fouling as a function of process variables. Such models are not available as yet. Under these circumstance it might be possible to model fouling using an equation of the type

$$\frac{dR_f}{dt} = k_d \exp\left(-E/RT_w\right) - k_r \tau R_f \tag{9.48}$$

where both constants k_d and k_r are functions of process variables such as surface shear stress. The direct dependence of the removal rate on the surface shear stress has been noted in a number of fouling experiments.

9.3.5 *Cleaning fouled surfaces*

In commercial practice, fouling must be lived with. Once a deposit has formed it must be removed, and this requires frequent and expensive cleaning. Cleaning involves several types of cost, of which the actual cost of cleaning chemical may not be the most severe; the production time lost both during cleaning and in preparing for it may be the most expensive loss in a plant designed for continuous operation.

Current cleaning techniques are largely empirical. This is largely due to the poorly understood nature of fouling; before cleaning can be optimized, it is necessary to understand both the nature of the deposit that is to be removed and the processes that give rise to it. Cleaning is necessary both to remove fouled deposit and to control possible contamination from microorganisms. The limit on the operation of a plant is generally the pressure drop through the system due to deposition; a plant is operated until fouling is so severe that it is no longer possible to maintain full flowrate through the equipment. If the plant is operated until the pressure drop is too great for the pumping capacity of the unit, the resulting fouling deposit is very difficult to remove. It may be that operation for a shorter period, producing a deposit that takes a shorter time to remove, would give higher overall production. Such calculations are difficult to do, and are rarely if ever done in practice; cleaning cycles either arise through trial and error or, more frequently, are arranged to fit in with existing shift patterns.

Industrial cleaning-in-place systems are highly developed and automated. Two types of chemical treatment are used.

- **Two-stage acid and alkali cleaners**: typically sodium hydroxide and nitric acid are used. Their use reflects the structure of the fouled deposit described in section 9.3.2; the alkali is added first to remove the protein

deposit and expose the thin mineral layer, which is then dissolved by the acid.

- **Single-stage commercial cleaners**, usually detergent based, and which contain surface-active agents to increase the wetting properties of the solution, decrease its surface tension and emulsify and disperse soil. Chelating agents may also be used to maintain removed material in solution.

Two-stage cleaners are more complex to use in practice, requiring extra dosage equipment and more rinsing steps, which are not needed by single-stage cleaners. Although two-stage alkali and acid sequences are designed to cope with organic and inorganic soils, caustic and acid alone have been shown to be insufficient to achieve a completely physically and chemically clean surface. It is, of course, difficult to define 'clean': in an engineering sense, a surface is clean if its subsequent fouling behaviour is indistinguishable from a surface that has never been used before. If material is left on a surface after cleaning it may provide nucleation sites for future deposition of protein aggregates.

Single-stage cleaners have been developed to produce a clean surface in a short time. Although they are more expensive than base chemicals, they are more efficient under most circumstances. A cleaner surface than is possible using caustic and acid alone can be obtained, together with savings in time, wash water, and energy. Comparisons of single- and double-stage cleaners do not, however, agree on which is most economic.

9.3.6 Stages and kinetics of cleaning

Cleaning is a multistage process. A soiled system consists of three phases: the **heat transfer surface**, the **deposit** and the **cleaning solution** (Fig. 9.19). Cleaning agent:

1. contacts the surface of the material to be removed;
2. wets and penetrates the deposit;
3. reacts and breaks down the deposited material; and
4. disperses the material into the cleaning solution.

Cleaning thus involves processes that are governed by mass transfer (stages 1 and 4) diffusion (stage 2) and reaction (stage 3), any of which could control the overall rate of the process. Problems with any step will result in a deposit which is difficult to clean, as follows.

- Mass transfer from the cleaning solution in low-shear areas – such as points of contact between exchanger plates – will be slow. This will limit both contact and dispersal.
- Non-wetting deposit surfaces will resist the cleaning solution.
- Cleaning material will diffuse only slowly through hard non-porous

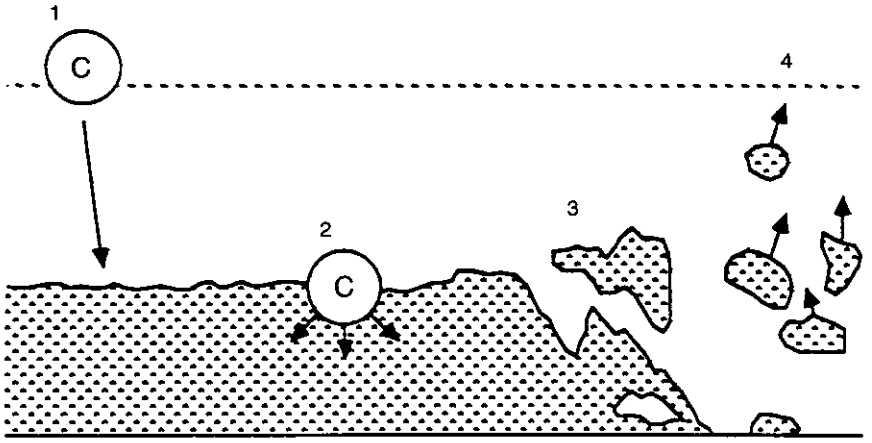


Fig. 9.19 Stages of cleaning: (1) mass transfer to the deposit; (2) penetration and wetting; (3) reaction to loosen deposit; (4) dispersal into fluid bulk.

deposit, such as the largely carbonized deposit that forms by overcooking.

- Some forms of deposit, again perhaps produced by overcooking, may be resistant to chemical attack; the reaction step will be slow.

The fouling model predicts that severe fouling will be found in low-shear areas; it is thus doubly important to ensure their absence.

Various kinetic models have been proposed for cleaning, but they have largely been produced by experiments on the cleaning of large systems, which contain a range of deposit types and thicknesses. These results are difficult to interpret kinetically. Figure 9.20 shows an electron micrograph of the cleaning, by sodium hydroxide, of a surface that has been fouled by whey protein concentrate; by comparison with the deposit seen in Fig. 9.13, a much more open structure can be seen. Deposit behaviour during cleaning from sodium hydroxide has also been observed visually. At time $t = 0$, sodium hydroxide contacts the deposit and weakly bound material is removed immediately. Over the next 5–20 s the surface of the deposit swells and becomes translucent. Over the next 10 min, the thickness of the translucent layer increases and the surface gradually breaks up, with aggregates about 0.1–0.2 mm detaching from the surface and being swept away. This suggests that removal of protein deposit by hydroxyl ions takes place in several stages:

1. contact between hydroxyl and deposit;
2. diffusion of hydroxyl through sponge to unreacted deposit;
3. reaction of hydroxyl ions with deposit to create an expanded deposit;

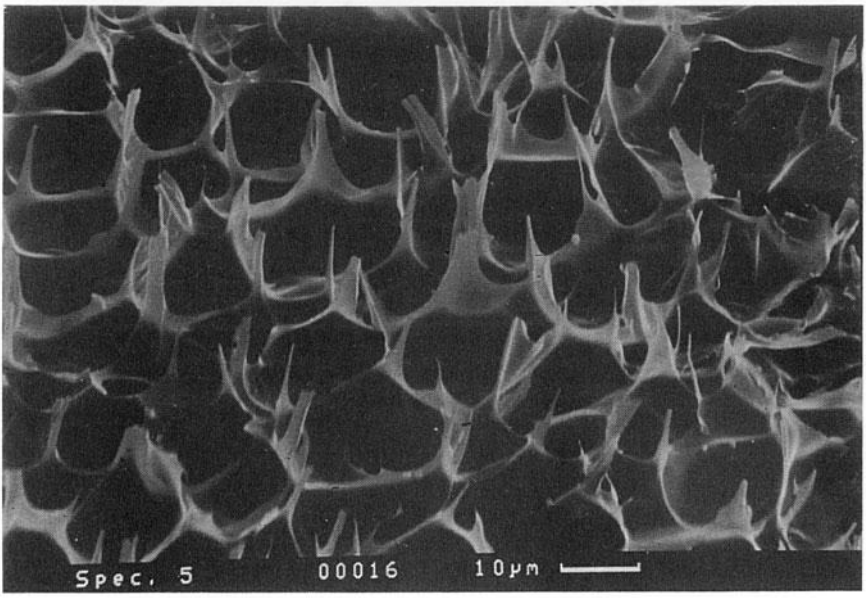


Fig. 9.20 Fouled surface that has been contacted with sodium hydroxide.

4. removal of this expanded deposit by fluid shear and by chemical reaction.

Figure 9.21 shows a typical cleaning curve for a whey protein deposit cleaned by sodium hydroxide. There is a delay before cleaning begins, presumably due to the time required for hydroxyl to diffuse into the deposit. The cleaning rate then builds up steeply, before falling to zero as deposit is removed. Cleaning rates vary as a function of temperature, flowrate and cleaning chemical concentration. Removal rate increase substantially at temperatures above 50 °C, suggesting that chemical reactions are taking place. Removal rate also increases as the flowrate increases. The effect of concentration is the most interesting result. Figure 9.22 shows a plot of the cleaning time for the same deposit as a function of the cleaning chemical concentration. An optimal concentration of cleaning chemical can be seen; the rate of removal of deposit for both 0% and 2% sodium hydroxide is much less than for 0.5% hydroxide. It appears that too high a concentration of cleaning chemical can seal the surface and prevent removal. This effect has been found for both whey protein concentrates and whole milk deposits. Industrially it suggests that the addition of extra cleaning chemical to cope with difficult deposit may increase the cleaning time rather than having the desired effect; increasing the temperature at which cleaning is carried out may prove more effective.

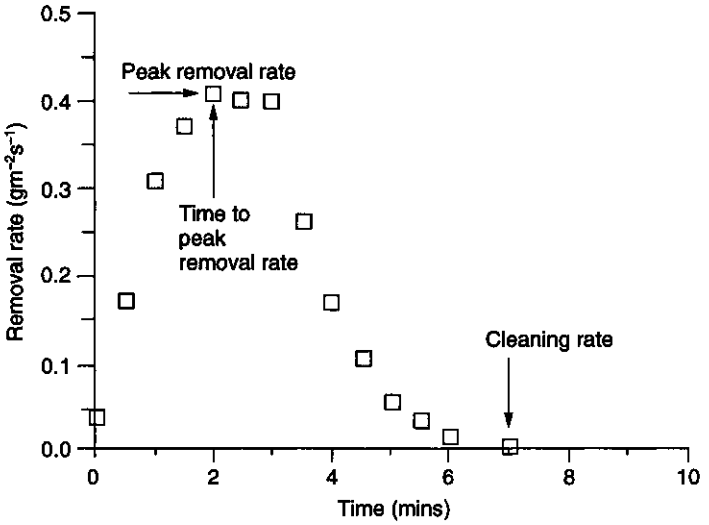


Fig. 9.21 Typical plot of protein removal as a function of time.

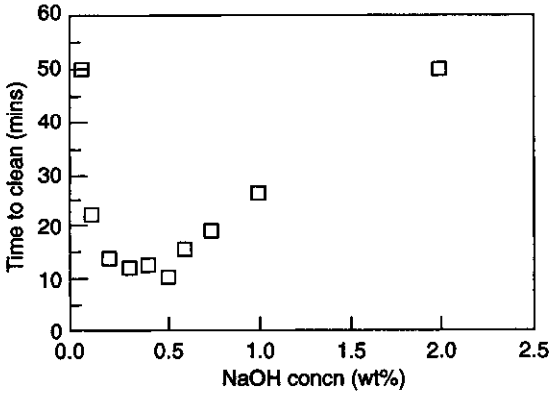


Fig. 9.22 The variation of cleaning time with cleaning agent concentration.

9.3.7 Control: monitoring fouling and cleaning

The above sections have described fouling and cleaning mechanisms. Both processes are complex and involve several stages. Unlike the sterilization and quality processes described earlier in this chapter, mathematical models are not available; this limits the optimization of process designs.

Until accurate kinetic data are available, the best way to ensure that plant is operated at peak efficiency is to monitor the plant and control it correctly. Very few plants are instrumented well enough to detect fouling until it is

large enough to affect outlet temperature or pressure drop to a significant extent. Any monitor developed should be cheap and small enough to be easily serviced and replaced but, above all, reliable and relevant enough for the information it provides to be acted on by plant operators.

Two types of measurement are required. In the latter stages of fouling it is sufficient to measure heat transfer and pressure drop across the whole plant. However, more subtle measurements are required in the early stages of fouling and the last stages of cleaning. Detecting fouling directly is difficult, especially in its early stages. Pressure sensors are least sensitive to the very small initial and final amounts of deposit. It is important to remove all deposit during cleaning or it will act as nucleation sites for fresh deposition when the plant is restarted. At this stage, it is not possible to propose a general approach to ensuring that optimal operating cycles are chosen; process plant must be monitored closely and operated in accordance with the principles described above.

Conclusions

Thermal processing is at the heart of the food industry, and this chapter returned to the subject, following our earlier incursions into energy balances, the analysis and design of heat exchangers and heat integration. As well as drawing on those chapters the discussion here also built on some of the elements discussed in the chapters on mass transfer and reaction engineering.

Several topics which are specific to the food industry and which are not covered in existing texts were touched on. The first section was mainly concerned with how to design batch thermal processes so as to simultaneously achieve the desired level of sterilization and an acceptable level of product quality. The discussion then moved on to some specific engineering problems in the design and operation of continuous processes; this included a discussion of the importance of the residence time distribution (described in the previous chapter) for process operation and a number of other important issues in the selection and heat transfer equipment.

Another section was concerned with the important operations involving mixtures of fluids and solids, where the classical chemical engineering texts have little that is directly relevant to offer: the discussion here touched on aspects of flow and heat transfer behaviour and of modelling of these systems, including microwave and ohmic heating processes.

The final section was concerned with fouling in thermal food processes, and outlined some recent results on the mechanisms of fouling together with a discussion of the significance of fouling for quantitative process design. The results presented here apply only to one particular case, but, as throughout this book, the principles are of general use.

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