

Cooling-Water Fouling in Heat Exchangers

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Abstract

The formation of deposits on heat transfer surfaces is an important problem during convective and nucleate boiling heat transfer to water. Possible deposition mechanisms are crystallization from dissolved salts, particulate fouling from suspended material, and microbial and macrobial growth. To account for unreliable design procedures and operational problems, heat exchangers are typically oversized by 70 to 80%, of which 30 to 50% is attributed to fouling. Further refinement of heat-exchanger design procedures will be possible only in conjunction with more sophisticated methods to include the effects of fouling. While the installation of excess heat transfer surface may extend the operation time of heat exchangers, it provides no remedy against the deposition of dirt.

Heat-exchanger fouling can be reduced by proper heat-exchanger design, by appropriate selection of heat-exchanger type, and by mechanical and/or chemical mitigation methods. All methods require an understanding of the mechanisms of dirt deposition and of the influence of operating parameters on the deposition rate. It is still not possible to predict a fouling resistance versus time relationship for a given water composition. However, it is possible to estimate the effects of flow velocity, foulant concentration and surface temperature on the deposition rate and to use this knowledge to optimize the design of heat exchangers.

In this paper, results of numerous experimental and theoretical investigations during the past 20 years are presented. The state of the art in fouling mitigation is reviewed, and special attention is drawn to areas where future academic and industrial research is required.

I. Introduction

The need for proper energy management has considerably increased the importance of heat-exchanging equipment over the past two to three decades. Today, there is virtually no chemical, food processing, or power-generating facility without a whole variety of heat exchangers. This trend has stimulated an enormous development of heat-exchanger design procedures, which has become possible as a result of improved knowledge of heat transfer fundamentals and the availability of larger and faster computers. Design calculations have shifted from overall heat transfer coefficients through length-averaged film coefficients to local film coefficients with integration along the flow path. Sophisticated software, which sometimes requires a mainframe computer, is available for the most common heat-exchanger types, such as shell-and-tube, plate-and-frame, coiled bundles, and plate-fin heat exchangers. The most widely known and accepted design software was developed by Heat Transfer and Fluid Flow Service (HTFS) in the United Kingdom and by Heat Transfer Research Incorporated (HTRI) in the United States.

A. DESCRIPTION OF PROBLEM

The usefulness of the above design procedures is considerably reduced by the fact that they are valid only for clean fluids. In most practical processes, however, heat-exchanging fluids such as water contain a certain amount of dissolved or suspended material or provide conditions favorable for the growth of biological organisms. This may lead to the growth of deposits on the heat transfer surfaces. Figure 1 shows excessive mineral deposits on the cooling-water side of an ammonia reactor effluent gas cooler.

Since the thermal conductivity of such deposits is low (see Table I), their resistance to heat transfer may well be of the same order of magnitude of that of the clean fluids. As a result, reduced heat-exchanger performance is common, and regular cleaning may be required. This is shown in Fig. 2 for the live steam heater in a bauxite refinery, where sodium aluminum silicate deposits necessitate acid washes at intervals of between 5 and 15 days. A mail survey covering 3000 heat exchangers from about 1100 New Zealand companies [1, 2] showed that more than 90% of heat exchangers had some fouling problems (see Fig. 3).

B. DESIGN PRACTICE

The possibility of deposition on the heat transfer surfaces is considered in the design of heat exchangers by using so-called fouling resistances in the calculation of the overall heat transfer coefficient:

$$\frac{1}{U} = \left(\frac{1}{\alpha_i} + R_{fi} \right) \frac{d_o}{d_i} + \frac{d_o \ln(d_o/d_i)}{2\lambda_w} + \frac{1}{\alpha_o} + R_{fo} \quad (1)$$



FIG. 1. Fouling on the shell side of an ammonia reactor effluent cooler operated with *treated* cooling water.

TABLE I
THERMAL CONDUCTIVITY OF VARIOUS DEPOSITS

Sodium aluminium silicate	0.2–0.4 W/(m · K)
Milk omponents	0.5–0.7 W/(m · K)
Hematite (boiler deposit)	0.6 W/(m · K)
Biofilm	0.7 W/(m · K)
Calcium sulfate (boiler)	0.8–2.2 W/(m · K)
Calcite (boiler deposit)	0.9 W/(m · K)
Serpentine (boiler deposit)	1.0 W/(m · K)
Gypsum (boiler deposit)	1.3 W/(m · K)
Calcium sulfate	2.3 W/(m · K)
Magnesium phosphate	2.3 W/(m · K)
Calcium phosphate	0.6 W/(m · K)
Calcium carbonate	0.9 W/(m · K)
Magnetite iron oxide	2.9 W/(m · K)

The almost exclusive source of fouling resistances is the approximately 100 values suggested by the Tubular Exchanger Manufacturers Association (TEMA) [3], which have not been significantly updated since their first publication in 1947. Since TEMA values are available for only a very limited number of fluids

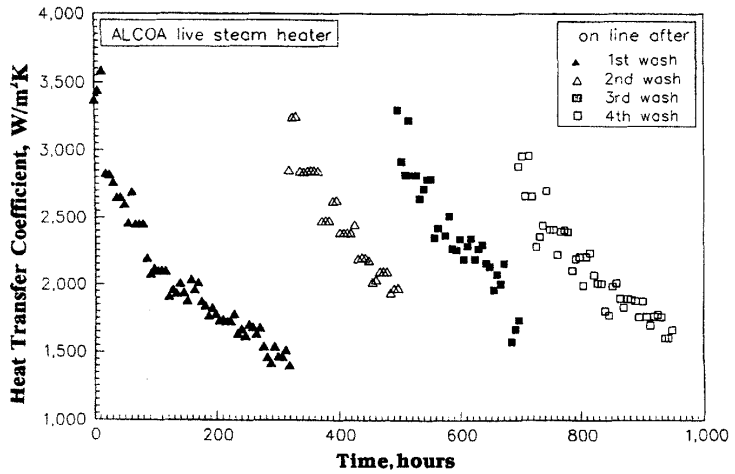


FIG. 2. Reduction in heat transfer due to sodium aluminum silicate deposition in the live steam heat exchanger of a bauxite refinery.

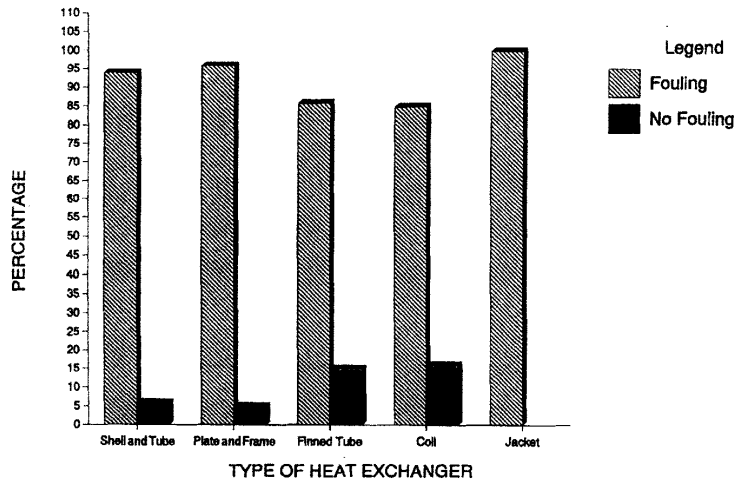


FIG. 3. Fouling problems in various heat-exchanger types [1, 2].

(mainly water and hydrocarbon processing streams) and do not include any correction for the effects of operating parameters (dirt concentration, surface temperature, flow velocity), selection of the appropriate fouling resistance is more or less guesswork or “gut feeling”. In many cases, the user of a heat exchanger overspecifies the fouling resistance because of operational uncertainties, the process designer increases it because of process design uncertainties, and, finally, the design engineer adds a further percentage because of uncertainties in the prediction of the clean heat transfer coefficients. None of these safety factors is related to the mechanisms or the seriousness of the actual fouling problem. It is also common to copy fouling resistances from “similar” plants built by the company in the past, even though operational conditions and fluid compositions may have changed significantly.

For a given heat duty, the following equation can be derived:

$$\frac{A_f}{A_c} = 1 + U_c R_f \quad (2)$$

Table II shows the excess heat transfer surface required for several types of heat exchanger if a typical TEMA fouling resistance of $0.18 \text{ m}^2 \cdot \text{K/kW}$ is used for each of the two heat-exchanging fluids. Obviously, the percentage of excess surface area increases with increasing clean heat transfer coefficient. This puts a heavy penalty on compact heat-exchanger types if, because of ignorance or because of cautiousness, the TEMA fouling resistances for shell-and-tube heat exchangers are used.

C. COST OF FOULING

Despite the enormous costs associated with heat-exchanger fouling, only very limited research has been done to determine accurately the economic penalties caused by fouling and to attribute these costs to the various aspects of heat-exchanger design and operation. However, reliable knowledge of fouling

TABLE II
EXCESS SURFACE AREA FOR VARIOUS HEAT-EXCHANGER APPLICATIONS ($R_f = 0.36 \text{ m}^2 \text{K/kW}$)

Application	Clean overall coefficient	Excess area
Gas-gas shell- and-tube heat exchanger	$50 \text{ W}/(\text{m}^2 \cdot \text{K})$	1.8%
Liquid-gas shell-and-tube heat exchanger	$150 \text{ W}/(\text{m}^2 \cdot \text{K})$	5.4%
Liquid-liquid shell-and-tube heat exchanger	$1000 \text{ W}/(\text{m}^2 \cdot \text{K})$	36%
Liquid-liquid plate-and-frame heat exchanger	$3000 \text{ W}/(\text{m}^2 \cdot \text{K})$	108%
Water-cooled shell-and-tube steam condenser	$4500 \text{ W}/(\text{m}^2 \cdot \text{K})$	162%

economics is desirable in order to evaluate the cost-efficiency of various mitigation strategies. The total fouling-related costs consist of the following:

1. *Capital expenditures.* These include excess surface area, costs for stronger foundations, provisions for extra space, and increased transport and installation costs. There are additional capital costs for antifouling equipment, such as the installation of on-line cleaning devices, pretreatment plants, and cleaning in-place equipment.
2. *Fuel costs.* Costs for extra fuel occur if fouling leads to extra fuel burning in furnaces or boilers or if more secondary energy such as electricity or process steam is needed to overcome the effects of fouling.
3. *Maintenance costs.* Maintenance costs are costs for removing fouling deposits, costs for chemicals, or other operating costs of antifouling devices. According to Pritchard [4] and Thackery [5], about 15% of the maintenance costs of a process plant can be attributed to heat exchangers and boilers, and of that, 50% is probably due to fouling.
4. *Costs Due to Production Loss.* Because of planned and unplanned plant shutdowns due to fouling in heat exchangers, large production losses are possible. These losses are often considered to be the main cost of fouling and are very difficult to estimate.

A detailed analysis of the various cost components is given by Garrett-Price *et al.* [6] and Steinhagen *et al.* [1]. Surprisingly good agreement was found among the results given by Steinhagen *et al.*, Pritchard, and Garrett-Price *et al.* [1, 4, 6] for the total costs due to fouling. While Garrett-Price *et al.* [6] and Pritchard [4] found that total fouling costs for highly industrialized countries such as the United States and the United Kingdom are about 0.25% of the countries' gross national product (GNP), the fouling costs for New Zealand, which is less

TABLE III
TOTAL FOULING COSTS FOR SEVERAL COUNTRIES

Country	Fouling in costs (million U.S. dollars)	1984 GNP (billions U.S. dollars)	Fouling as % of GNP
United Kingdom (1978)	700–930	285	0.20–0.33
United States (1982)	3,860–7,000	3,634	0.12–0.22
	8,000–10,000		0.28–0.35
New Zealand	35	23	0.15
Australia	260	173	0.15
West Germany	1,533	613	0.25
Japan	3,062	1,225	0.25
Total industrialized world	26,850	13,429	0.20

industrialized, are 0.15% of the New Zealand GNP [1]. Using these percentages, Table III lists total fouling related costs for various countries, in 1984 U.S. dollars.

II. Fouling Mechanisms during Heat Transfer to Water

Water is by far the most common fluid used in industry. According to Suito *et al.* [7], 8200 m³/s of cooling water are used in the United States alone. This is about one-third of the country's total water consumption for all purposes. Other significant uses of water are in steam generators and as caustic or acidic process liquors, for example, in the pulp and paper industry and in bauxite refineries. This article concentrates on cooling-water fouling.

Deposits formed from water may consist of crystals, silt, corrosion products, or biological growth.

Crystallization fouling. Precipitation and deposition of salts that are supersaturated under process conditions. The solubility limit may be exceeded as a result of

- Evaporation of the water
- Cooling of solutions with normal solubility (increasing solubility with increasing temperature)—for example, CaCl_2 and NaCl in Fig. 4—below the saturation concentration
- Heating of solutions with inverse solubility (decreasing solubility with increasing temperature)—for example, aqueous solutions of calcium carbonate, calcium sulfate, calcium phosphate, calcium silicate, calcium hydroxide, magnesium silicate, magnesium hydroxide, sodium sulfate, lithium sulfate, and lithium carbonate—above the solubility temperature.
- Mixing of different product streams
- Changes in pH

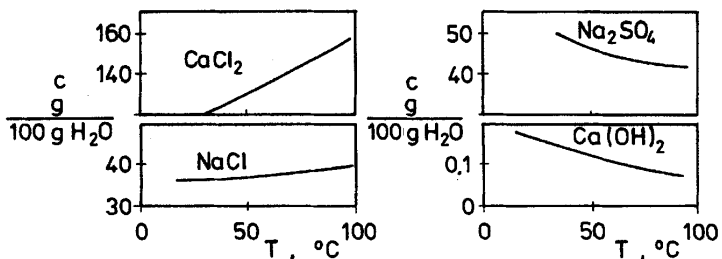


FIG. 4. Normal- and inverse-solubility salts.

Particulate fouling. Accumulation of finely divided solids (sand, clay, dust, rust) suspended in the process fluid on the heat transfer surface. In some instances, settling by gravity prevails, and the process may then be referred to as *sedimentation fouling*.

Corrosion fouling. Formation of an oxide layer on the heat transfer surface. Since these layers are generally thin and have a comparatively high thermal conductivity, they offer only slight resistance to heat transfer. However, the increasing roughness of the heat transfer surface may promote other deposition mechanisms by providing nucleation sites or sheltered zones.

Biological fouling. Biological fouling refers to the development and deposition of organic films consisting of microorganisms and their products such as bacteria (microbial or microbiofouling) and the attachment and growth of macroorganisms such as mussels, algae, etc. (macrobiofouling), on the heat transfer surfaces. Microbial fouling always precedes fouling by macroorganisms, with the microorganisms acting as the nutrient source for the macroorganisms. Suspensions of seaweed and other organic fibers often cause fouling. Many types of bacteria will deposit slime on the heat transfer surfaces, and other types of foulants can adhere to these deposits. Further growth restricts the fluid flow and often causes pitting of the metal. This type of fouling is common with untreated water such as sea, river, or lake water. Microbiological fouling is a particularly serious problem, as the microbes may be introduced into the water cycle not only by the fluid, but also from the ambient air in the cooling tower. Temperatures between 15 and 50°C in cooling towers are ideal for microbial growth. Both dead and live microorganisms adhere to the heat transfer surfaces and form a layer of slime with a thermal conductivity similar to that of water. As a consequence of the filtering effect of biological layers, more suspended particles amass in the deposit.

Fouling mechanisms within each category may be described with similar models. Generally, several fouling mechanisms occur at the same time, and they are nearly always mutually reinforcing. One notable exception is particle deposition together with crystallization, which weakens an otherwise tenacious scale.

III. Sequential Events of Fouling

The above fouling mechanisms generally occur in five consecutive steps:

1. *Initiation period or delay period.* When the new or cleaned heat exchanger has been put into operation, the initially high heat transfer coeffi-

cients may remain unchanged for a certain time. During this time, nuclei for crystallization are formed or nutrients for biological growth are deposited. This delayed period may last anywhere from a few seconds to several days. According to Blöchl and Müller-Steinhagen [8], no delay period occurs for particulate fouling. For crystallization fouling and chemical reaction fouling, the initiation period decreases with increasing surface temperature, as supersaturation and/or reaction rate increases [9]. Generally, it is reported that the delay time before deposition starts decreases with increasing roughness of the heat transfer surface.

2. *Mass transport.* To form a deposit at the heat transfer surface, it is necessary for at least one key component to be transported from the fluid bulk to the heat transfer surface. In most cases, this occurs by diffusion. For the transport of particles to the wall, inertia effects and thermophoretic forces have to be considered, too.
3. *Formation of deposit.* After the foulant has been transported to the heat transfer surface, it must stick to the surface (for particulate fouling) or react to the deposit-forming substance (e.g., CaCO_3).
4. *Removal or autoretardation.* Depending on the strength of the deposit, erosion occurs immediately after the first deposit has been laid down. Furthermore, several mechanisms exist that cause autoretardation of the deposition process. For the thermal boundary condition of constant temperature difference between heated and cooled fluid, the growth of deposit causes a reduction of the driving temperature difference between the heat transfer surface and the fluid.
5. *Aging.* Every deposit is subjected to aging. Aging may increase the strength of the deposit through polymerization, recrystallization, dehydration, etc. Biological deposits are poisoned by metal ions and may be washed away by the bulk flow. Aging is the least investigated and understood step and is usually ignored in modeling attempts.

Depending on the process parameters and the dominant fouling mechanism, the fouling rate can either be constant or decrease with time (see Fig. 5).

For hard, adherent deposits such as silicates or some polymerization products, steps 1 and 5 may be ignored and the growth rate of deposits is constant or continuously decreasing with time. For weaker deposits (for example, particles $> 1 \mu\text{m}$), the fouling resistance approaches a constant (or asymptotic) value, which may or may not allow acceptable operation of the process.

In what follows, the main individual fouling mechanisms are discussed first, and then plant measurements in which several mechanisms may interact in an unidentified manner are outlined.

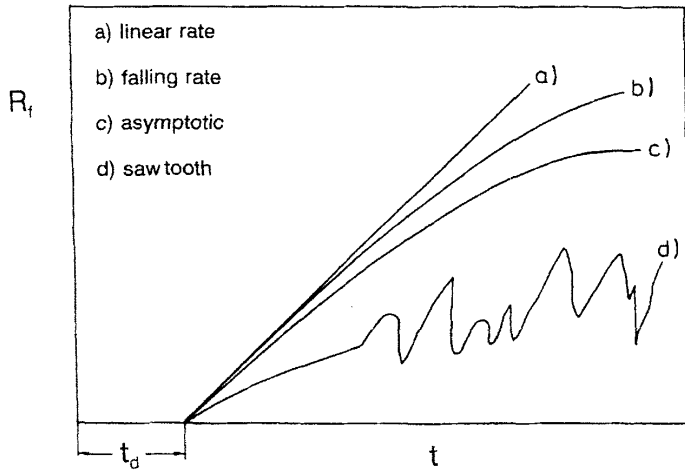


FIG. 5. Possible fouling resistance versus time curves.

IV. General Approach to the Modeling of Heat-Exchanger Fouling

Most fouling models are based on the assumption by Kern and Seaton [10] that the growth of deposit on the heat transfer surface is the net value of a transport flux to the heat transfer surface and a removal flux away from the heat transfer surface:

$$\frac{dm}{dt} = \dot{m}_d - \dot{m}_r = \lambda_d \rho_d \frac{dR_f}{dt} \quad (3)$$

To form a deposit at the heat transfer surface, all or most of the material must be transported from the fluid bulk to the heat transfer surface and there undergo a chemical reaction or a sticking process.

The *transport* step is usually modeled as a diffusion process. The local diffusion flux to the heat transfer surface is calculated from

$$\dot{m}_t = \beta(c_b - c_s) \quad (4)$$

where c_b and c_s are the concentrations of key components in the fluid bulk and at the heat transfer surface, respectively. The mass transfer coefficient β has to be determined from an appropriate Sh-Re-Sc- d/L relationship. In specific cases, there may be other contribution to mass transfer, such as inertia, electrostatic, or thermophoretic forces [11, 12].

Adhesion and/or reaction can be modeled in analogy to a chemical reaction.

$$\dot{m}_a = k_r(c_s - c^*)^n \quad (5)$$

The rate constant k_r is proportional to an Arrhenius term

$$k_r = Ke^{-E/RT} \quad (6)$$

For continuity reasons, $\dot{m}_t$ must be equal to $\dot{m}_a$. Hence Eq. (4) can be combined with Eq. (5) to eliminate the unknown interface concentration c_s . If the adhesion-reaction rate is considerably higher than the diffusion rate, the deposition is transport-controlled: $c_s \approx c^*$ and $\dot{m}_d \approx \dot{m}_r$. If the adhesion rate is considerably slower than the transport rate, $c_s \approx c_b$ and $\dot{m}_d \approx \dot{m}_a$. In this case, deposition is adhesion- or reaction-controlled.

In most models, it is assumed that the *removal rate* is proportional to the fluid shear stress and to the thickness of the deposit:

$$\dot{m}_r = K\tau_w s \quad (7)$$

The constant K in Eq. (7) may be proportional to a “deposit strength”, which is very difficult to evaluate experimentally. In general, it has been found that deposits generated at higher flow velocities are more compact and adhesive [13]. The integration of Eq. (3) yields

$$R_f(t) = R_f^*(1 - e^{-t/t_c}) \quad (8)$$

The asymptotic fouling resistance R_f^* and the time constant t_c can be evaluated from Eqs. (4) to (8).

V. Crystallization Fouling

A. INDICES FOR THE SCALING TENDENCY OF WATER

Crystallization cannot occur unless the solubility of the salt in question is exceeded. Scaling indices use the chemistry of an aqueous solution to indicate whether the solution has the potential for scaling or corrosion. Although more sophisticated methods for the prediction of scaling rates are available, e.g., [14–16], water treatment is usually determined using one of the following indices (*NALCO Water Handbook* [17], *BETZ handbook* [18], *DREW water treatment handbook* [19]).

1. Saturation Index

The saturation index (in this example written for CaCO_3)

$$\text{S.I.} = \frac{[\text{Ca}^{++}] \cdot [\text{CO}_3^{--}]}{[\text{Ca}^{++}]_s \cdot [\text{CO}_3^{--}]_s} \quad (9)$$

uses the degree of saturation to determine the scaling potential of a solution. Scaling may occur only if the solution is supersaturated; hence,

S.I. > 1 scaling tendency

S.I. < 1 corrosion tendency

The thermodynamic activities of the ions to be used in Eq. (9) can be calculated from the chemical analysis of the water if the level of dissolved salts is not too high, using, for example, the expression by Davies [20] for the activity coefficients γ_1 and γ_2 of mono- and divalent ions for ionic strength up to 0.1:

$$\log_{10} \gamma_1 = -A \left(\frac{I^{1/2}}{1 + I^{1/2}} + 0.3I \right) \quad \gamma_2 = \gamma_1^4 \quad (10)$$

where

$$A = \frac{1.8246 \cdot 10^6}{\{78.54T[1 - 0.004579(T - 25) + 1.19 \cdot 10^{-5}(T - 25)^2]\}^{3/2}} \quad (11)$$

with T in degrees Celsius. The ionic strength I is given by

$$I = \frac{1}{2} \sum m_i z_i^2 \quad (12)$$

where the summation is over all ions present and m_i and z_i are the molar concentrations and their charges, respectively. Approximately,

$$I \approx 2.5 \cdot 10^{-5} \text{TDS} \approx 1.6 \cdot 10^{-5} \Lambda \quad (13)$$

where TDS is the concentration of total dissolved solids in milligrams per liter and Λ is the conductivity in microsiemens.

In practice, ionic concentrations are sometimes used in place of the thermodynamic activities to obtain an approximate value of S.I. This approximation is valid in dilute solutions, but can lead to serious errors in more concentrated ones.

2. Langelier Saturation Index

Langelier [21, 22] developed a qualitative formula to predict whether CaCO_3 will be precipitated or dissolved in water. The equation for the Langelier

Saturation Index is based on pH, total alkalinity, hardness, and temperature, all of which affect the solubility of calcium carbonate in water.

$$\text{L.S.I.} = \text{pH} - \text{pH}_s \quad (14)$$

For CaCO_3 solutions, the pH at saturation, pH_s , can be calculated from

$$\text{pH}_s = \text{p}K_2 - \text{p}K_{sp} + \text{p}(\text{T.A.}) \quad (15)$$

with T.A. being the total alkalinity. Scaling occurs if $\text{L.S.I.} > 0$, corrosion if $\text{L.S.I.} < 0$.

The Langelier Saturation Index is very similar to the saturation index. Therefore, similar predictions can be expected with both indices.

3. Ryznar Stability Index

Saturation indices show only the direction of the driving force; they do not indicate whether the supersaturation is high enough to initiate crystallization. Ryznar [23] therefore suggested an index that is said to be a quantitative as well as a qualitative measure of the solutions' scaling potential. The Ryznar Stability Index is defined as

$$\text{R.S.I.} = 2\text{pH}_s - \text{pH} \quad (16)$$

It was shown experimentally that the following criteria can be used to determine the fouling potential of a solution [23]:

$\text{R.S.I.} < 6$ denotes a scaling tendency

$\text{R.S.I.} > 7$ denotes a corrosion tendency

4. Comparison of Indices

The predictions of the three indices were compared for incipient scaling conditions, i.e., for $\text{S.I.} = 1$, $\text{L.S.I.} = 0$, and $\text{R.S.I.} = 6$ [24]. The results may be interpreted as the pH to which the process has to be adjusted for a given composition and temperature to minimize scaling and corrosion. Figure 6 shows the prediction of the three indices as a function of the heat transfer surface temperature.

Scaling occurs above the curves; corrosion is likely for conditions below the curves. Generally, scaling tendency increases with increasing values of pH and surface temperature. As was expected, the saturation index and the Langelier Saturation Index give comparable values and trends. pH values for incipient scaling predicted by the Ryznar Stability Index differ considerably from those predicted by the two saturation indices. For lower solution temperatures, the Ryznar Stability Index predicts higher pH values than the two saturation indices,

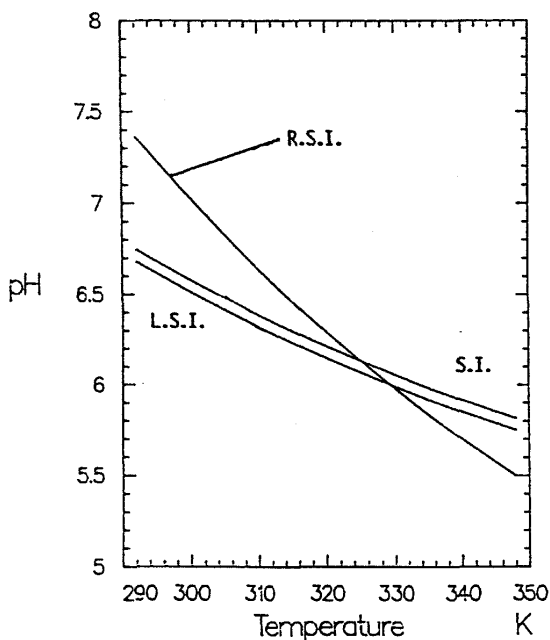


FIG. 6. Comparison of incipient scaling conditions predicted by S.I., L.S.I., and R.S.I. $[Ca^{++}] = 0.01122$ mol/L, T.A. = 0.00483 mol/L.

which indicates that a certain supersaturation is necessary to initiate scaling. However, for higher temperatures, scaling is predicted to start for $CaCO_3$ concentrations well below the saturation concentration. Obviously, this result is not in accordance with the basic understanding of crystallization kinetics. The deviation between the predictions of S.I or L.S.I. and R.S.I. increases with increasing Ca^{++} concentration [24]. Preference should be given to the predictions of saturation indices in situations where the Ryznar index predicts scaling for subsaturated conditions.

B. MODELS FOR SCALE FORMATION IN HEAT EXCHANGERS

The following discussion provides some insight into the mechanisms of scale formation from various salts. It is limited to convective heat transfer. Because of

the mechanism of microlayer evaporation, scale formation is usually more severe during nucleate boiling than without bubble formation [25]. A notable exception is Bayer process liquor, where concentration effects related to bubble formation suppress the formation of silica deposits [25].

Because of the high strength of crystalline deposits, most fouling models predict a linear increase in fouling resistance with time, as long as the salt concentration and the temperature at the solid-liquid interface remain constant. This has been confirmed by most experiments that have been performed under carefully controlled conditions. However, if suspended particles (broken crystals or foreign matter) are present in the fluid, the linear fouling curve will change to a falling rate or even an asymptotic curve, because of the reduced strength of the resulting deposits. This is demonstrated in Fig. 7, which shows measurements of CaSO_4 fouling with and without an in-line filter to remove particulate matter.

A falling-rate curve is also obtained if heat transfer occurs with a constant wall temperature or constant hot fluid inlet temperature boundary condition, rather than with constant heat flux, as is the case in most industrial heat exchangers. In this case, the temperature at the deposit-liquid interface will decrease with time because of the insulating effect of fouling. The reduced interface temperature causes an autoretardation effect, as it reduces the value of the reaction rate constant k_r and the driving concentration difference ($c_s - c^*$).

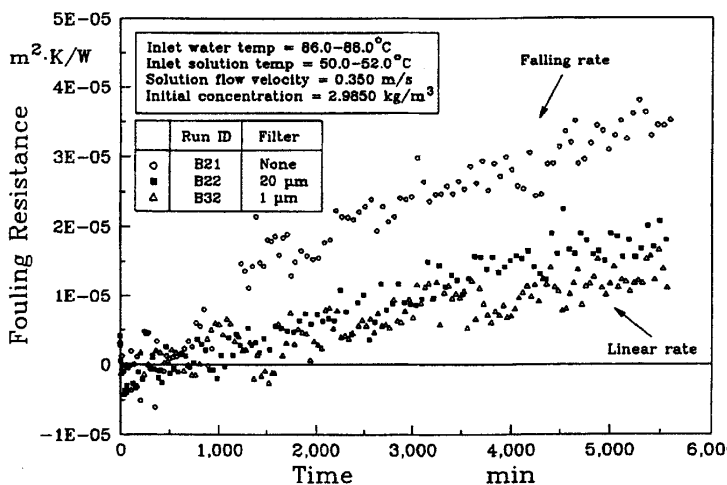


FIG. 7. Effect of in-line filter on CaSO_4 fouling [27].

For the deposit to grow on the heat transfer surface, scale-forming ions have to diffuse from the fluid bulk to the solid-liquid interface, where the crystallization reaction takes place. Depending on the relative values of the rates of diffusion and reaction, the deposition process can be transport-controlled (slow diffusion, fast reaction), reaction-controlled (fast diffusion, slow reaction), or a combination of both. Models can be simplified if only one step is controlling.

1. CaSO_4 Scaling

CaSO_4 fouling is a key problem if seawater is used as a cooling medium. For temperatures above 40°C , CaSO_4 has a decreasing solubility with increasing temperature. As shown in Fig. 8, calcium sulfate deposition from an aqueous solution appears in three forms: gypsum ($\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$), calcium sulfate hemihydrate ($\text{CaSO}_4 \cdot 0.5\text{H}_2\text{O}$), and calcium sulfate anhydride (CaSO_4).

The reaction rate of anhydride, which is the thermodynamically most stable form, is so slow that usually only the other two forms are observed. However, even in this case, the deposition process is reaction-controlled, since the reaction rates are still significantly slower than the diffusion rates. Scale formed from solutions that do not contain any suspended particles is relatively hard and subject to little erosion.

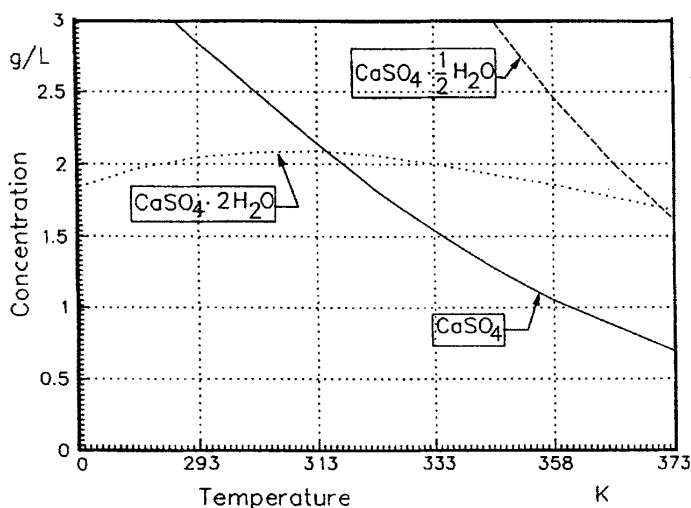


FIG. 8. Solubility of various forms of calcium sulfate in water.

- + Reaction at the heat transfer surface
- Little erosion

This is supported by Fig. 9, which shows the results of experiments on CaSO_4 deposition in a double-pipe heat exchanger. The deposition increases with increasing surface temperature, but shows an almost linear increase with time.

For flow velocities above 0.5 m/s, no influence of velocity on the deposition rate was observed, as long as the wall temperature was maintained constant (see Fig. 10).

CaSO_4 deposition rates can be predicted with good accuracy with the following correlations [15], if it is assumed that the deposition process consists of two consecutive steps. In the first step, calcium and sulfate ions are transported to the heat transfer surface by convection,

$$\dot{m}_t = \beta(c_b - c_s) \quad (17)$$

where they react to CaSO_4 and precipitate in crystalline form:

$$\dot{m}_a = k_r(c_s - c^*)^2 \quad (18)$$

In these equations, c_b , c^* and c_s are the CaSO_4 concentration in the fluid bulk, at saturation (with respect to surface temperature), and at the interface between

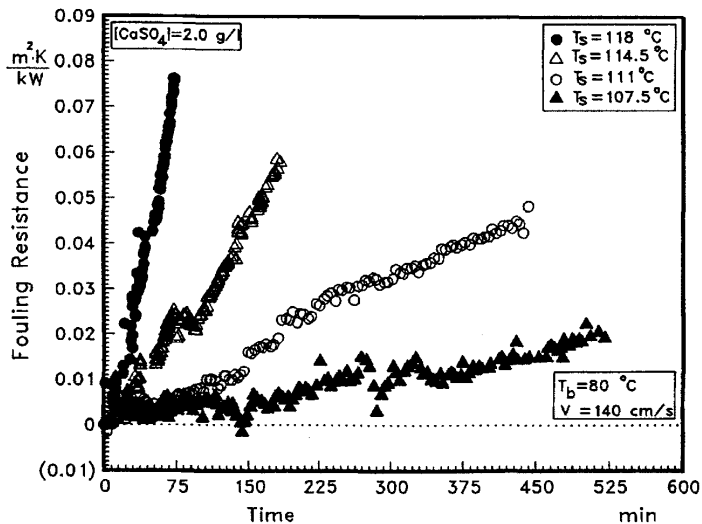


FIG. 9. CaSO_4 Deposition as a function of time and heat transfer surface temperature [15].

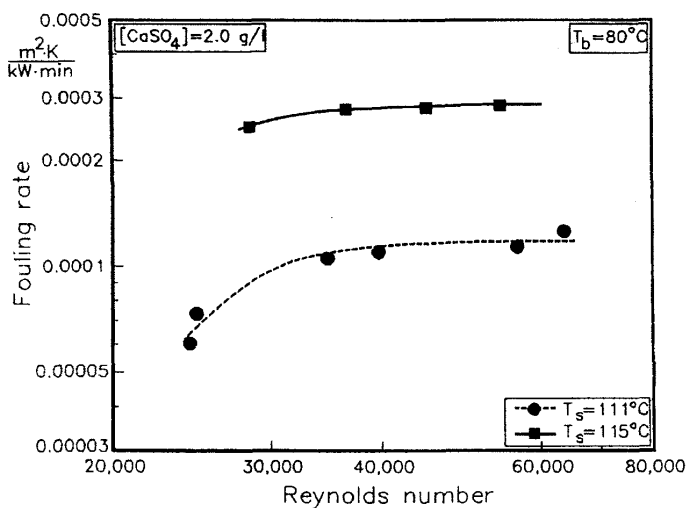


FIG. 10. Influence of flow velocity on the deposition rate for constant initial wall temperature [15].

deposit and fluid, respectively. The reaction rate constant k_r increases exponentially with temperature:

$$k_r = K_{1,0} e^{-E/RT_s} \quad (19)$$

If the interface concentration is eliminated by equating Eqs. (17) and (18),

$$\dot{m} = \beta \left[\frac{1}{2} \left(\frac{\beta}{k_r} \right) + (c_b - c^*) - \sqrt{\frac{1}{4} \left(\frac{\beta}{k_r} \right)^2 + \left(\frac{\beta}{k_r} \right) (c_b - c^*)} \right] \quad (20)$$

The constants required in the above equations have been obtained from nonlinear regression analysis of the experimental data. These values are listed in Table IV. The physical properties for density and thermal conductivity of deposits are the average of the values given by Hasson and Zahavi and by Partridge and White [28, 29]. The mass transfer coefficients are predicted by a Dittus-Bölder [30] type equation using the analogy between heat and mass transfer:

$$\beta = 0.023 \text{ Re}^{0.85} \text{ Sc}^{0.33} \frac{D}{d_h} \quad (21)$$

TABLE IV
PHYSICAL PROPERTIES AND CONSTANTS FOR CaSO_4 DEPOSITION

$K_{1,0}$ [$\text{m}^4/(\text{kg} \cdot \text{s})$]	E (J/mol)	ρ_D (kg/m^3)	λ_d [W/(m · K)]
$5.7 \cdot 10^{11}$	112,500	2160	2.2

From Ref. 15.

Diffusion coefficients for Ca^{++} and SO_4^- are given by Bohnet [31] for an average temperature of 82.5°C [$D(82.5^\circ\text{C}) = 1.0633 \cdot 10^{-9} \text{m}^2/\text{s}$]. Using this information and the approximate relation suggested by Bird *et al.* [32],

$$D_{T_2} = D_{T_1} \left(\frac{T_2}{T_1} \right) \cdot \exp \left[3.8 T^* \left(\frac{1}{T_1} - \frac{1}{T_2} \right) \right] \quad (22)$$

where T^* denotes the boiling temperature of the liquid, the diffusivity can be calculated for any temperature.

Solutions with a small concentration of suspended CaSO_4 particles cause substantially increased deposition, with a decreasing or asymptotic trend (see Fig. 7). In this case, it is possible to reduce the deposit formation substantially by

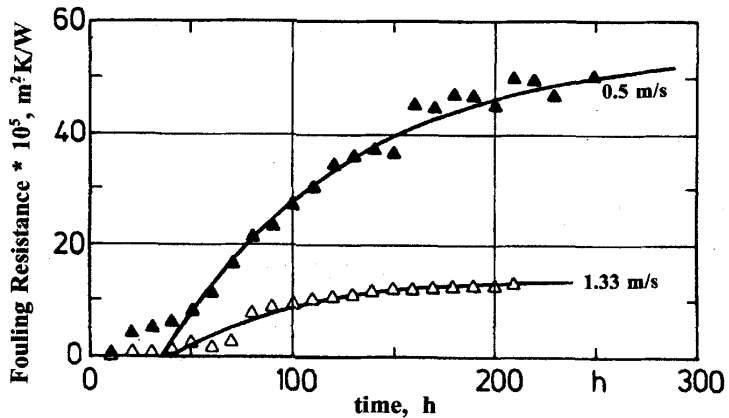


FIG. 11. Influence of flow velocity on unfiltered, saturated CaSO_4 solutions [33].

increasing the flow velocity, as shown in Fig. 11 for the results obtained by Krause [33]. Krause used a saturated CaSO_4 solution with suspended CaSO_4 particles [33]. The relatively soft CaSO_4 deposits can be strengthened through the addition of CaCO_3 particles [34] or weakened through the addition of Al_2O_3 particles [35].

Crystallization fouling from dissolved CaSO_4 in plate and frame heat exchangers was investigated by Müller-Steinhagen and coworkers [35–38]. Information about industrial applications and fouling of plate heat exchangers is given in Sec. VIII. Figure 12 shows a typical picture of a fouled heat-exchanger plate. The flow in the channel is diagonal, entering on the bottom left-hand side and leaving on the top right-hand side of the plate.

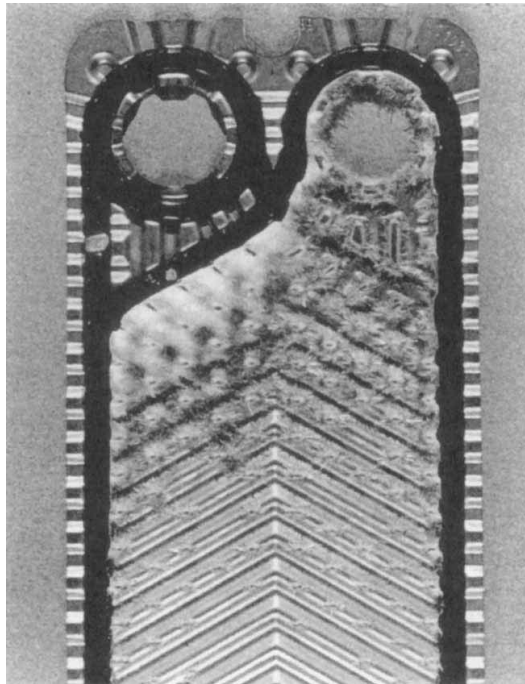


FIG. 12. CaSO_4 deposition in a plate- and -frame heat exchanger.

For the following reasons, the deposition is more severe at the hot end of the plate:

1. The driving concentration difference increases because fluid and wall temperatures increase from inlet to outlet. This decreases the solubility and therefore increases the supersaturation if the bulk concentration is assumed to be constant.
2. The crystallization rate constant increases with increasing temperature.

The diagonal flow results in low-velocity zones, indicated by A and B in Fig. 13. In these areas, shear forces are at a minimum and the wall temperature is close to the temperature of the heating medium. Therefore, most deposition occurs in the top left-hand corner because of the low flow velocity and high fluid-surface temperature in this area. The extent of the stagnant zones, and therefore of the deposition, decreases with increasing flow velocity.

A closer look at the crystal formation in a certain area of the heat transfer plate reveals that the crystals vary in size and amount. Around the contact points

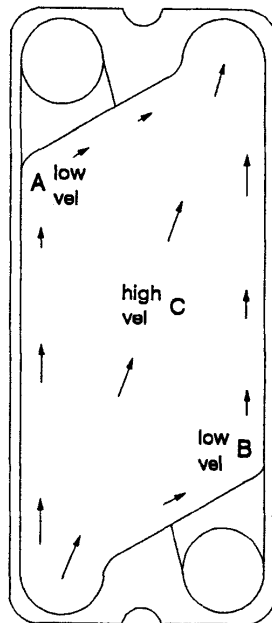


FIG. 13. Velocity distribution in a plate- and -frame heat exchanger.

between the plates, the flow velocity is low, resulting in higher wall temperatures and lower removal rates. Because of the particular geometry at these locations, the temperature profile in the vicinity of the contact points is different from that in the flow channel, since a volume of liquid is heated from several directions. As a result of these effects, most crystal formation is initiated near the contact points. Obviously, this does not apply for the stagnant flow zones in the heat exchanger. Further examination of the crystal formation near a contact point shows that the crystals can be divided into two different size categories: An almost homogeneous deposit of small crystals appears directly on the plate surface, and long single crystals are formed on top of this initial layer.

Since the crystallization of gypsum is reaction-controlled, the rate of deposition increases with increasing wall temperature and bulk concentration (see Figs. 14–16). With increasing flow velocity, both the initial fouling rate and the absolute value of the fouling resistance decrease (see Figs. 17 and 18). The initial fouling rate decreases as a result of lower wall temperature at higher flow velocities, whereas the absolute value of the fouling resistance decreases as a result of a higher removal rate in conjunction with a lower interface temperature.

The CaSO_4 concentration was important only in the initial stages of the fouling process. Later, the channel gap is reduced considerably by the deposit, causing other factors such as higher shear stresses and lower interface temperature to

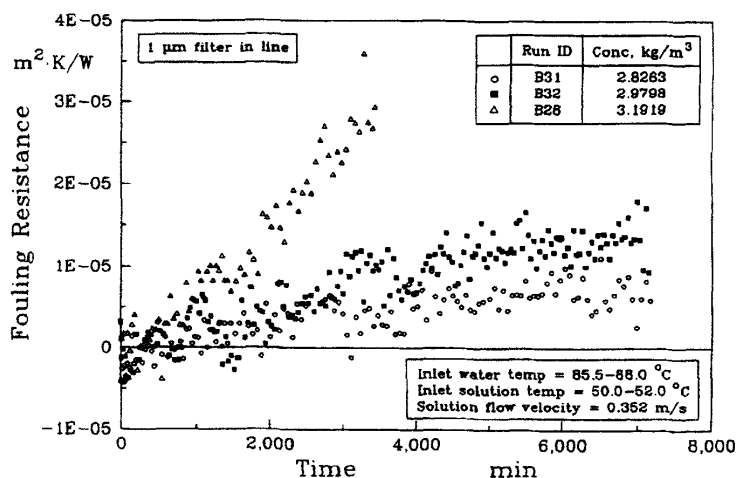


FIG. 14. Effect of CaSO_4 concentration on crystallization fouling in a plate heat exchanger.

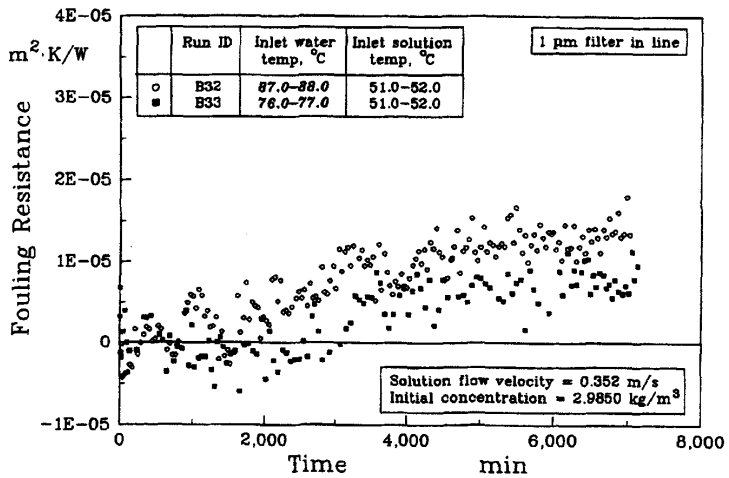


FIG. 15. Effect of wall temperature on crystallization fouling in a plate heat exchanger.

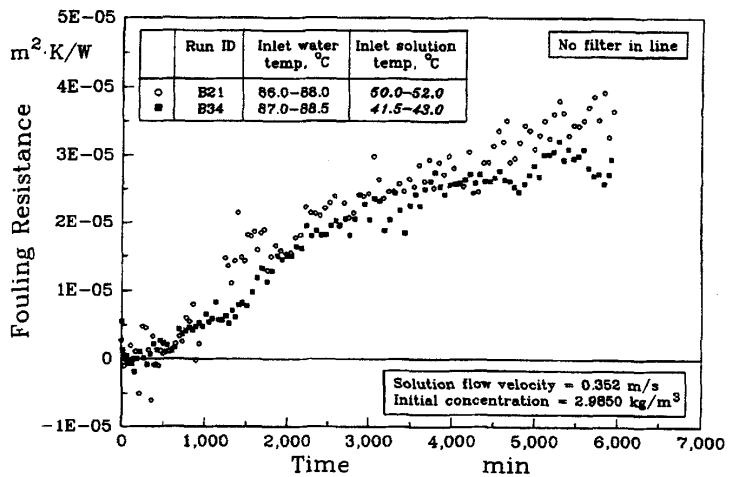


FIG. 16. Effect of fluid temperature on crystallization fouling in a plate heat exchanger.

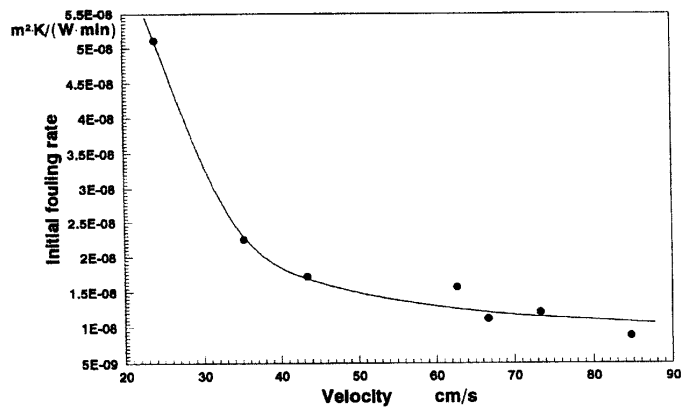


FIG. 17. Effect of flow velocity on the initial fouling rate in a plate heat exchanger.

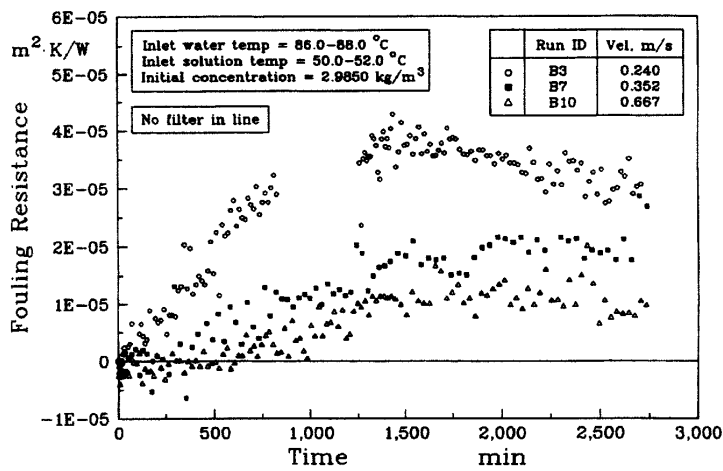


FIG. 18. Effect of flow velocity on the asymptotic fouling resistance in a plate heat exchanger.

become more important. While the difference in geometry between tubular and plate heat exchangers has an effect on the absolute amount of fouling (see Sec. VIII), the above effects of operating conditions on deposition are generally applicable.

2. CaCO_3 Scaling

Calcium carbonate scaling is the main contribution to cooling-water fouling in heat exchangers and pipelines. It is caused by local supersaturation as a result of temperature, pH, and pressure changes. For typical flow velocities, reaction rate and diffusion rate are of the same order of magnitude, and deposits are relatively hard. Hence, the following mechanisms have to be considered in the modeling process:

- + Transport
- + Reaction at the heat transfer surface
- Erosion ($u < 1 \text{ m/s}$)

The influence of Reynolds number on the deposition rate is shown in Fig. 19, whereas Fig. 20 shows the increase in scaling with increasing CaCO_3 concentration.

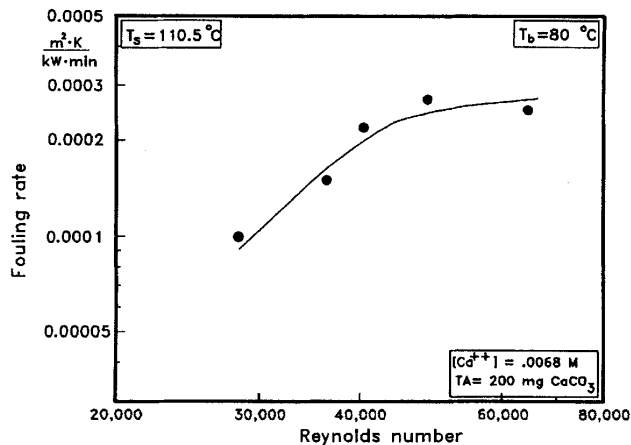


FIG. 19. Influence of Reynolds number on the deposition rate of CaCO_3 .

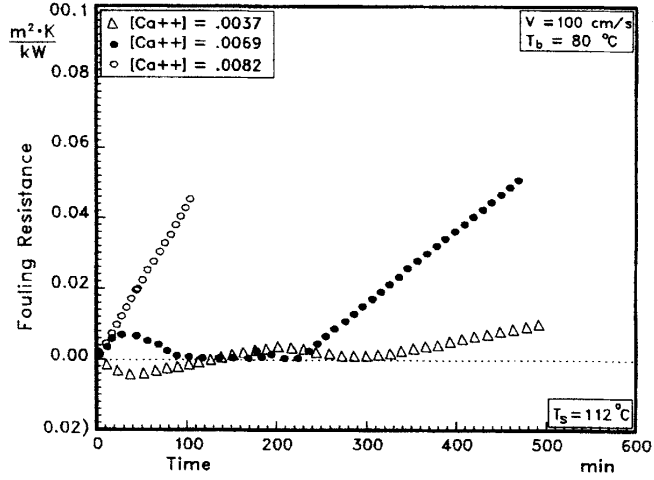


FIG. 20. Influence of Ca^{++} concentration on the fouling resistance.

Hasson's ionic diffusion model combines equations for solubility, reaction, and diffusion [14]:

$$\dot{m}_d = \frac{\beta[Ca^{++}][(1 + 4ac/b^2)^{0.5} - 1]b}{2a} \quad (23)$$

with

$$a = 1 - \frac{4K_2k_r[Ca^{++}]}{\beta K_1} \quad (24)$$

$$b = \frac{[CO_2]}{[Ca^{++}]} + \frac{4K_2k_r[HCO_3^-]}{\beta K_1} + \frac{K_{sp}k_r}{\beta[Ca^{++}]} \quad (25)$$

$$c = \frac{K_2k_r[HCO_3^-]^2}{K_1\beta[Ca^{++}]} - \frac{K_{sp}k_r[CO_2]}{\beta[Ca^{++}]} \quad (26)$$

Correlations for equilibrium constant, reaction rate constant and solubility of $CaCO_3$ in water are given in Branch and Müller-Steinhagen [39]. The mass

transfer coefficient β is calculated with Eq. (21). From the deposition flux, the fouling rate can be determined:

$$\frac{dR_f}{dt} = \left(\frac{\dot{m}}{\lambda \rho} \right)_{\text{scale}} \quad (27)$$

Watkinson [40] found an average value of $5,000 \text{ kg} \cdot \text{W}/\{\text{m}^4 \cdot \text{K}\}$ for $(\lambda \rho)_{\text{scale}}$.

Hence, if pH, T.A., and $[\text{Ca}^{++}]$ are known, variables a , b , and c can be determined and Eqs. (23) and (27) can be solved to give a value for the fouling rate. The predictions of Hasson's ionic diffusion model agreed well with measurements by Watkinson [40], as long as the fouling rates remained below $4 \cdot 10^{-6} \text{ m}^2 \cdot \text{K}/\text{kJ}$ (see Fig. 21).

Hasson's model predicts that scale formation increases with increasing flow velocity. This peculiar effect is caused by the relatively fast reaction between the ions at the heat transfer surface, which leaves the rate-determining step to the diffusion from the fluid bulk to the liquid-solid interface. It is expected that this will change for higher flow velocities. For most sparingly soluble salts, the reaction rate constant is rather small and deposition is controlled by the rate of reaction, e.g., the surface temperature. Equation (23) does not account for any deposit removal mechanisms because of the strength of CaCO_3 deposits. Hasson

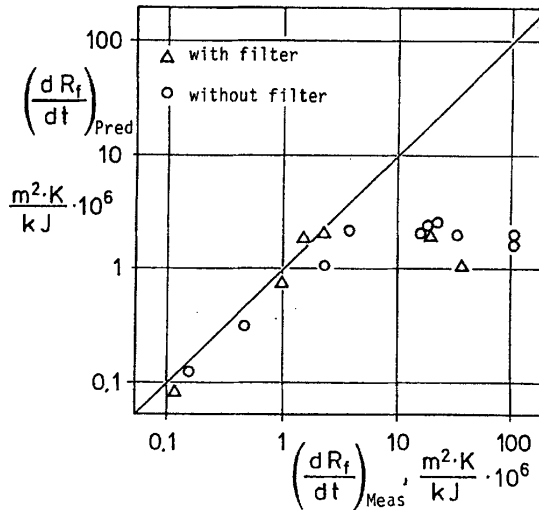
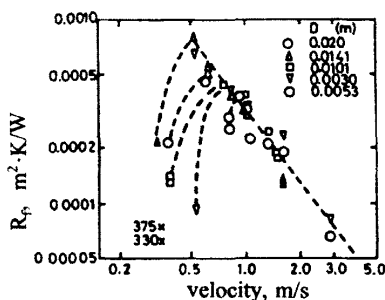


FIG. 21. Measured and predicted CaCO_3 scaling rates [40].

FIG. 22. Effect of flow velocity on CaCO_3 scaling [39].

[14] claims that this assumption is valid for flow velocities below 0.8 m/s. For higher flow velocities (and weaker deposits), a significant reduction in the asymptotic fouling resistance is observed (see Fig. 22).

Chan and Ghassemi [42, 43] recommend a model that is based on the conservation equations. For an electrically heated cylindrical rod, one obtains

$$\frac{\partial c_i}{\partial t} + (u \nabla c_i) = D_{i\text{eff}} \nabla^2 c_i + R_A \quad (28)$$

Assuming a pseudo-steady state and neglecting reactions in the bulk flow, this equation simplifies to

$$u \nabla c_i = D_{i\text{eff}} \nabla^2 c_i \quad (29)$$

and with cylindrical coordinates,

$$u_r \frac{\partial c_i}{\partial r} + u_\theta \frac{1}{r} \frac{\partial c_i}{\partial \theta} + u_z \frac{\partial c_i}{\partial z} = D_{i\text{eff}} \left[\frac{1}{r} \frac{\partial}{\partial r} \left(r \frac{\partial c_i}{\partial r} \right) + \frac{1}{r^2} \frac{\partial^2 c_i}{\partial \theta^2} + \frac{\partial^2 c_i}{\partial z^2} \right] \quad (30)$$

Flow components in the radial and tangential directions can be neglected, and the diffusion flux of CaCO_3 is much lower than the convective flux, because of its low concentration. With these assumptions,

$$u_z(r) \frac{\partial c_i}{\partial z} = D_{i\text{eff}} \left[\frac{1}{r} \frac{\partial}{\partial r} \left(r \frac{\partial c_i}{\partial r} \right) \right] \quad (31)$$

This equation can be solved with the following boundary conditions:

$$\text{At } r = kR: \quad D_i \frac{\partial c_i}{\partial r} = k_r ([Ca^{++}] - [Ca^{++}]^*)^2 \quad (32)$$

$$\text{At } r = R \quad \frac{\partial c_i}{\partial r} = 0$$

$$\text{At } z = 0 \quad c_i = c_i^0$$

where kR is the radius of the heater rod. The velocity distribution in the annulus, can for developed flow [$u_z = u_z(r)$] be approximated according to Bird *et al.* [32] by

$$u_{z,\max} - v_z = \frac{1}{k_1} \left(\frac{\tau_0 \lambda^2 - k^2}{\rho k} \right)^{1/2} \ln \frac{(\lambda - k)R}{r - kR} \quad \text{for } r < \lambda R \quad (33)$$

$$u_{z,\max} - v_z = \frac{1}{k_1} \left[\frac{\tau_0}{\rho} (1 - \lambda^2) \right]^{1/2} \ln \frac{R - \lambda R}{R - r} \quad \text{for } r > \lambda R \quad (34)$$

where

$$\lambda^2 = \frac{1 - k}{2 \ln(1/k)} \quad (35)$$

The shear stress at the wall τ_0 can be obtained if the friction factor for the inner wall is known. This factor, f_1 , for annuli has been correlated with the flow Reynolds number by Perry and Chilton [44]:

$$f_i = f_o \frac{\lambda^2 - 2k^2}{k(1 - \lambda^2)} \quad (36)$$

where f_o is the friction factor based on the outside diameter of the annulus.

$$f_2 = \frac{0.0791}{[Re_f(1 - \lambda^2)]^{1/4}} \quad (37)$$

Now the wall shear stress may be calculated from

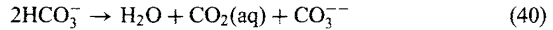
$$\tau_0 = f_1 \rho (u_z)^2 \quad (38)$$

The above equations can be solved for the appropriate boundary conditions with a finite difference method. The deposition rate can then be evaluated from

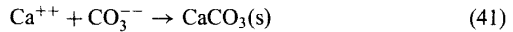
$$\dot{m}_t = D_{i,\text{eff}} \left(\frac{\partial c_i}{\partial r} \right) \quad (39)$$

A comparison between predicted values and deposition rates measured by Najibi *et al.* [16] showed an average relative error of 44%.

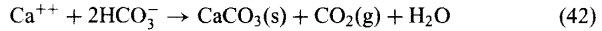
A new approach to the prediction of CaCO_3 deposition during convective heat transfer and subcooled flow boiling has been suggested by Najibi *et al.* [16], in analogy to the deposition of CaSO_4 . Again it was assumed that deposition occurs in two steps. Because of the concentration gradient, Ca^{++} and HCO_3^- ions will diffuse from the fluid bulk to the heat transfer surface. The following equilibrium conditions exist for HCO_3^- :



CO_2 produced during the reaction will desorb at the gas–liquid interface. The CO_3^{--} ions diffuse to the heat transfer surface to react with Ca^{++} to $\text{CaCO}_3(\text{s})$:



The above reaction equations are frequently combined in the following form [43]:



Diffusion of calcium and bicarbonate ions and the crystallization reaction at the heat transfer surface can be predicted with Eqs. (17) to (22) in complete analogy to CaSO_4 deposition. The reaction rate constant and the activation energy for CaCO_3 deposition have been fitted to the data measured by Hasson and Perl [45]. All values are given in Table V.

The density (2705 kg/m^3) and thermal conductivity [$1.942 \text{ W}/(\text{m} \cdot \text{K})$] of the calcium carbonate deposits are the average of the values measured by Hasson and Perl [45] and by Sheikholeslami and Watkinson [46]. The diffusion coefficients of Ca^{++} and HCO_3^- at 336 K have been calculated according to Reid *et al.* [47]. The value of $1.14563 \text{ E-}9 \text{ m}^2/\text{s}$ has to be extrapolated to other temperatures using Eq. (22). The average relative error between measured and predicted scaling rates is 5%. Comparison of the performance of various models for prediction of CaCO_3 scaling for convective and subcooled flow boiling heat transfer is given in Najibi *et al.* [16].

Depending on the position of the inlet and outlet nozzles (see Fig. 23), shell-and-tube heat exchangers can have ineffective sections or sections in which heating and cooling streams become reversed. These cells have a significant influence on the overall performance of the heat exchanger and its fouling behavior.

TABLE V
PHYSICAL PROPERTIES AND COEFFICIENTS FOR CaCO_3 DEPOSITION

$k_{1,0}$ [$\text{m}^4/(\text{kg} \cdot \text{s})$]	e [J/mol]	λ_D [kg/m^3]	ρ_d [$\text{W}/(\text{m} \cdot \text{K})$]
9.8×10^{11}	122,150	1.942	2705

From Ref. 16.

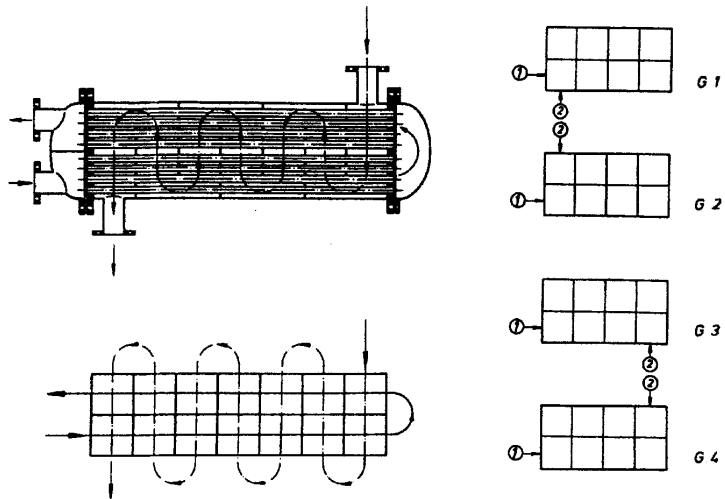


FIG. 23. Possible locations of inlet nozzles of tube-side and shell-side fluid in a 1-2 shell-and-tube heat exchanger.

Branch and Müller-Steinhagen [39] modeled CaCO_3 fouling in shell-and-tube heat exchangers by combining Hasson's ionic diffusion model [14] with the Schlünder-Gaddis model [48] for the prediction of the temperature distribution. The computed results allow assessment of clean heat-exchanger design rules and of the effects of fouling on the efficiency of heat-exchanger configurations. For the calculations described below, it was assumed that a tube-side cooling-water flow is heated by a nonfouling fluid on the shell side. The following solutions were considered for the investigation:

	Solution			
	1	2	3	4
$[\text{Ca}^{++}]$ (mol/L)	0.004	0.006	0.008	0.01
T.A. (mol/L)	0.004	0.004	0.004	0.004

Of the above four solutions (1) is subsaturated, (2) is just saturated, and (3) and (4) are supersaturated. Figure 24 shows the effect of fouling on the effectiveness of the four possible configurations of inlet and outlet nozzles.

Fouling reduces the differences between these configurations. All four configurations tend to approach the same effectiveness after a long period of

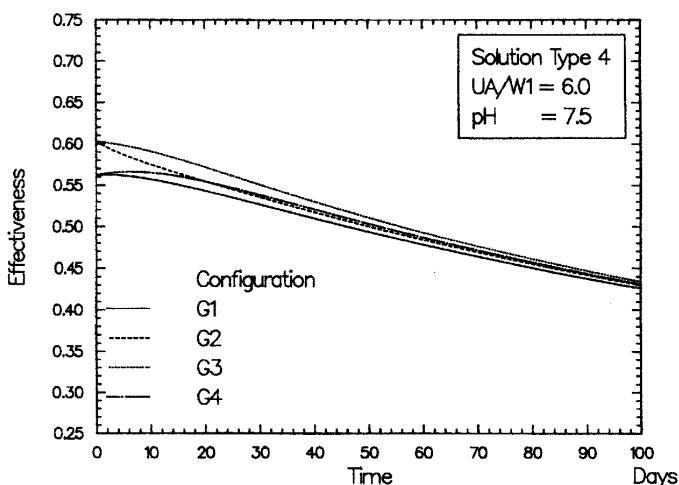


FIG. 24. Effect of CaCO_3 scaling on the effectiveness of the four possible flow arrangements.

operation. The effect of the CaCO_3 saturation on the effectiveness is shown in Fig. 25 for configuration G4.

Increasing the degree of supersaturation (in this case, increasing $[\text{Ca}^{++}]$) increases the extent of fouling. It was shown above that fouling reduces the differences between the four configurations. Therefore, the results shown in this diagram are applicable to any of the four configurations. The supersaturation can be altered in a number of ways. A common fouling mitigation procedure is to reduce the degree of supersaturation by decreasing the pH of the solution. Figure 26 shows how variation of the pH affects the fouling behavior of the heat exchanger.

The configuration shown, G4, has several ineffective sections. Reducing heat transfer in these areas initially improves the overall effectiveness of the heat exchanger. A maximum effectiveness is observed for each solution. As the pH is decreased, this maximum moves to the right. Since overtreating a solution can lead to severe corrosion problems, an optimum pH must exist for each solution and configuration. As shown in Fig. 26, a small amount of deposition may be advantageous for the overall performance of the heat exchanger. Therefore, the best result is obtained for $\text{pH} = 6.5$. The thin CaCO_3 deposit produced with this pH also acts as a protection against corrosion. Reducing the pH below a value of 6.3 does not cause any reduction in fouling. Heat-exchanger configurations that

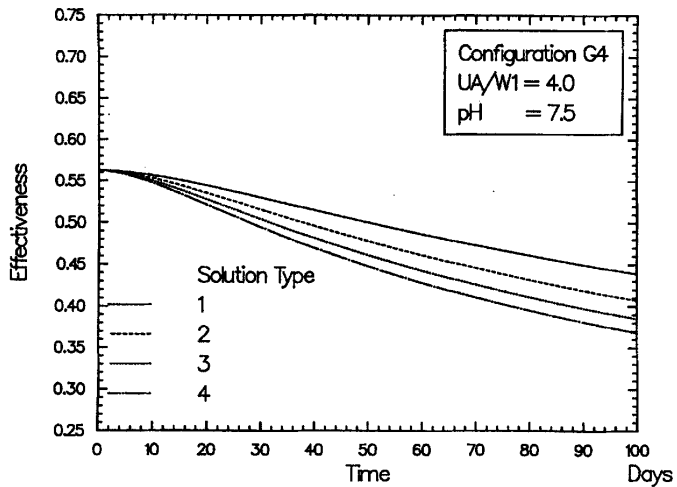


FIG. 25. Effect of CaCO_3 concentration on the effectiveness of a shell-and-tube heat exchanger.

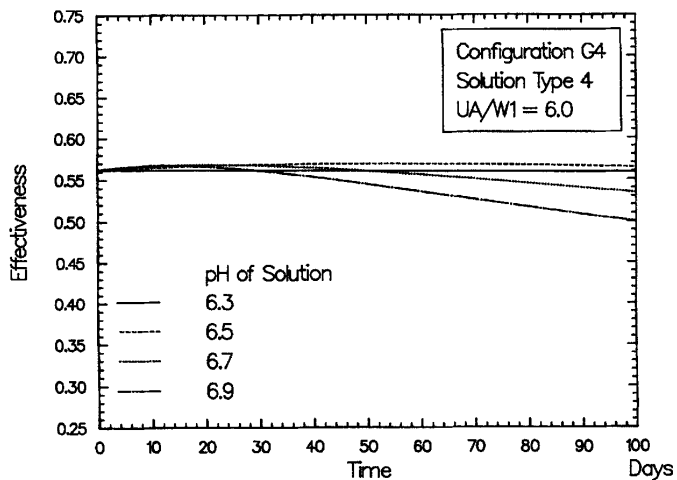


FIG. 26. Effect of solution pH on the fouling-related reduction in effectiveness.

do not show a maximum in their effectiveness curve also have an optimum pH. The optimum pH can be found by variation of the water treatment program or, less expensively, by computer modeling.

While no major differences between the four heat-exchanger configurations were found for scaling from supersaturated solutions (see Fig. 24), considerable deviations are predicted as the solution approaches equilibrium conditions. Figure 27 shows all four configurations being fouled by a solution close to saturation (solution 2). This graph shows that the design rules developed by Gaddis and Schlünder [48] are applicable only to clean heat exchangers, whereas a solution with low fouling potential can reverse their rules. As an example, Fig. 27 shows that configurations G2 and G3 should be selected based on clean heat transfer. If one considers the effectiveness after 100 days, a different conclusion is reached. In this case, G3 has the best performance, with G4 coming in second. Since there are numerous combinations of fouling solution, fouling mechanisms, and heat-exchanger configuration, selection of optimum exchangers will be specific to the problem at hand. However, the above results apply qualitatively not only to CaCO_3 scaling, but to any fouling in which deposition increases with wall superheat.

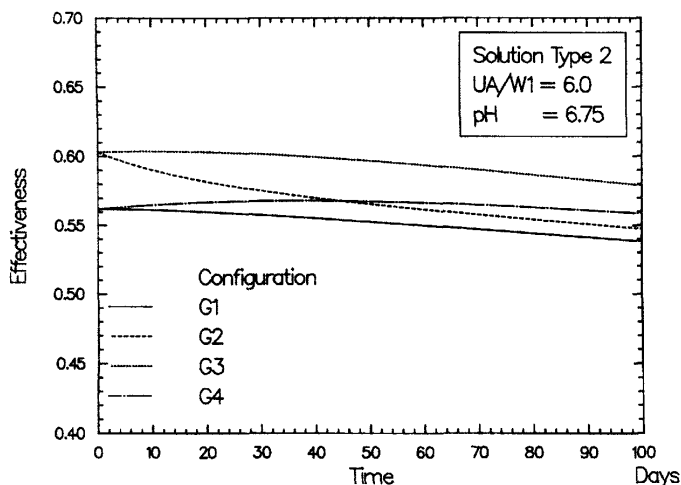


FIG. 27. Effectiveness as a function of time for a bulk subsaturated CaCO_3 solution.

VI. Particulate Fouling

Most process streams contain a certain amount of suspended solids as a result of corrosion, dust intake, or imperfect cleaning. Typical particle sizes vary from 1 to 20 μm , even though larger agglomerates are also possible. These particles are inevitably deposited on the surfaces of heat exchangers. An excellent summary of most of the investigations and models relating to particulate fouling is given by Gudmundsson [49]. This survey paper refers to about 150 investigations on particle deposition from gas and liquid streams with and without phase change.

Particulate deposits are not very strong and are subject to considerable removal rates as a result of the fluid shear. This leads to asymptotic fouling behavior.

- + Transport of particles (diffusion, inertia, thermophoresis)
- + adhesion (van der Waals forces, surface charges)
- + erosion of the deposit

The Kern and Seaton model [10]

$$\frac{dm_d}{dt} = \frac{dR_f}{\lambda_d \rho_d dt} = \dot{m}_d - \dot{m}_r \quad (43)$$

exactly applies to this situation. In integrated form, this model yields

$$R_f(t) = R_f^*(1 - e^{-\Phi t}) \quad (44)$$

where the asymptotic fouling resistance R_f^* and the time constant Φ depend on the mass transfer coefficient, the shear forces on the particles, and the adhesion forces between particles and between particles and the wall. Most models suggested in the literature are based on Eq. (43) and recommend different approaches for the above forces and flow rates. The effect of the main parameters on the deposition rate and on the asymptotic fouling resistance for particulate fouling will be discussed in the following sections.

A. EFFECT OF FLOW VELOCITY

Figure 28 shows the effect of flow velocity on fouling during heat transfer to Bayer liquor that was highly supersaturated with respect to silica [50]. Deposition for these experiments was mainly due to suspended particles, and it can be seen that the asymptotic fouling resistance decreases with increasing flow velocity, whereas the initial deposition rate increases. This result, which is typical for particulate fouling, is caused by the effect of flow velocity on the mass transfer coefficient and on the shear forces. While the removal rates are low as long as the deposit is thin (short times), they approach the deposition rates for higher deposit thickness.

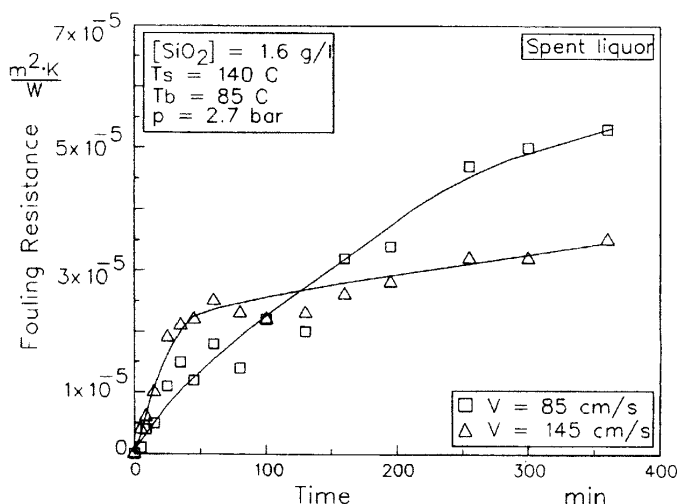


FIG. 28. Effect of flow velocity on particulate fouling from Bayer liquor [50].

Watkinson [51, 52] investigated deposition from *sand-water suspensions*. Sand slurries with a concentration of about 4ppm were circulated through a heated tube with 8.7mm inside diameter. The sand particles were in the size range 3 to 17 μm , the water temperatures were around 60°C, and the wall temperatures were around 77°C. Some of the most important results are given in Figs. 29 and 30, which show the initial deposition rate and the asymptotic fouling resistance as a function of the flow rate through the pipe.

With reference to Eqs. (43) and (44), Watkinson found the following relationships:

$$\dot{m}_d \sim u^{0.92} \quad (45)$$

$$R_f^* \sim u^{-1.2} \quad (46)$$

$$\Phi \sim u^{1.9} \quad (47)$$

However, Figs. 29 and 30 indicate that Eqs. (45) and (47) are applicable only for velocities below approximately 2.3 m/s. For higher velocities, a sharp drop in deposition rate and time constant is observed. Deposition was found to be both

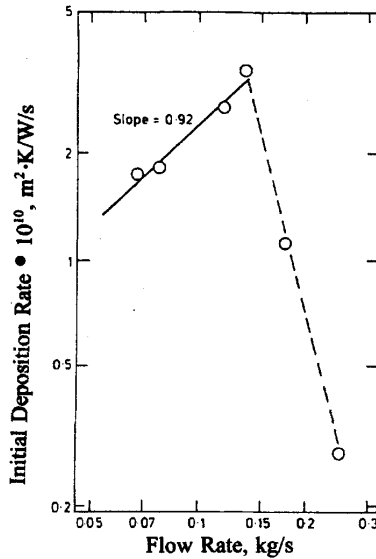


FIG. 29. Initial deposition rate as a function of flow rate [51, 52].

mass transfer- and adhesion-controlled and was modeled with a modified approach:

$$R_f(t) = \frac{C_1(c_b - c_s)\beta e^{-E/RT_s}}{C_2\lambda_d\tau_w^2} (1 - e^{C_2\tau_w t}) \quad (48)$$

where the constants C_1 and C_2 are system-specific. Below the critical velocity of 2.3 m/s, deposition is mass transfer-controlled; for higher velocities, adhesion may be the controlling step. Reported values of the thermal conductivity of sand deposits were between 5 and 13 W/(m·K), which seems somewhat high.

Deposition of *iron oxide particles* from aqueous suspension frequently occurs because of upstream corrosion. For this deposition process, a different velocity dependence was reported. For convective heat transfer to suspensions of 0.5 μm hematite particles, Hopkins and Epstein [53] found

$$\dot{m}_d \sim u^{0.3} \quad (49)$$

and linked this to the mechanism of crevice corrosion:

$$R_f^* \sim u^{-0.9} \quad (50)$$

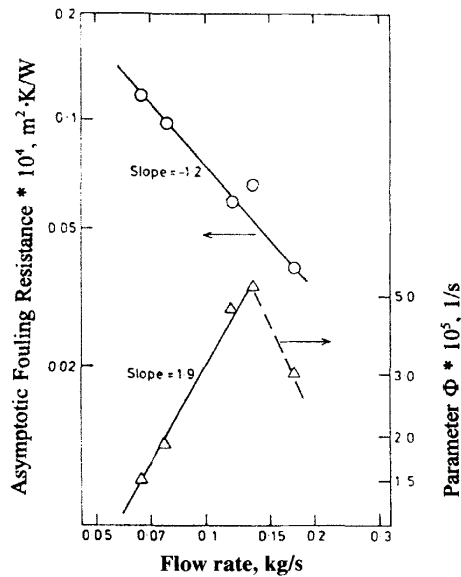


FIG. 30. Asymptotic fouling resistance and time constant as a function of flow rate [51, 52].

For submicron magnetite particles during nonboiling heat transfer in boiler plants, Thomas and Grigull [54] found

$$\dot{m}_d \sim u^{1.074} \quad (51)$$

and claimed that the deposition process was associated with the filling of natural cavities on the heat transfer surfaces. Reduced heat transfer from the smooth surfaces contributed to the reduced heat transfer efficiency as well as the thermal resistance of the deposit.

Obviously, there are significant disagreements among the results reported by various authors. The most likely explanation is the different size of particles, with particles in excess of 1 μm producing transport- and attachment-controlled fouling similar to that predicted by the Kern and Seaton model, whereas submicron particles directly affect the surface of the pipes.

B. EFFECT OF PARTICLE CONCENTRATION

Figure 31 shows that the asymptotic fouling resistance increases linearly with increasing particle concentration before leveling off to a constant value [8]. This

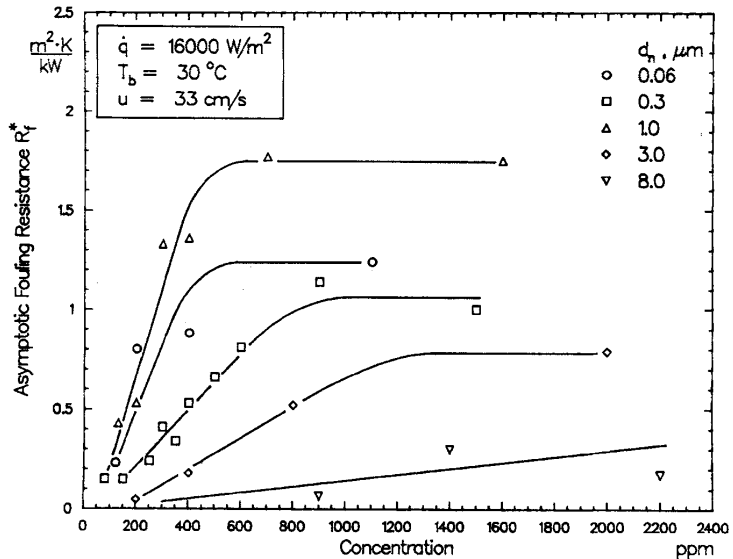


FIG. 31. Effect of particle size and concentration on the asymptotic fouling resistance from alumina particles suspended in X-2 [8].

effect, which is more distinct for smaller particles, is accompanied by the formation of larger particle agglomerates in the bulk fluid. It may be assumed that particulate fouling increases linearly with particle concentration as long as no agglomeration of particles occurs.

C. EFFECT OF SURFACE TEMPERATURE

The effect of surface temperature is correlated through the changing physical properties of the carrier liquid and an Arrhenius term for the adhesion "reaction" of the particles to the surface. While the effect of physical property change is usually not significant, the effect on the adhesion can be substantial. To stick to the surface, particles have to overcome an energy barrier due to the opposing direction and different length dependency of electrical double-layer forces and van der Waals forces (see Fig. 32) [55].

With increasing temperature, the kinetic energy of particles increases, and it becomes more likely that they will reach the primary minimum. The activation

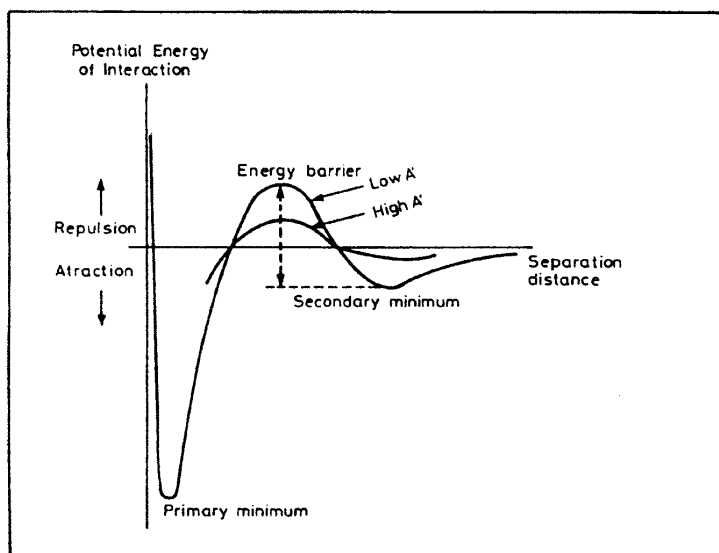


FIG. 32. Potential energy of interaction [55].

energy for sand–water fouling was found to be 64 kJ/mol [51], which agreed with measurements for Al_2O_3 /heptane systems by Müller-Steinhagen *et al.* [12].

D. EFFECT OF HEAT FLUX

The influence of the heat flux on the asymptotic fouling resistance due to particulate fouling is demonstrated in Fig. 33 [56]. For both test geometries used, the fouling resistance has a maximum for a certain heat flux. This maximum decreases in amplitude and shifts toward higher heat fluxes if the wall shear stress is increased [12]. It is hypothesized that this effect is caused by thermophoretic forces acting away from the heated surface. McNab and Meisen [57] developed the following correlation for the thermophoretic velocity:

$$v_t = 0.26 \frac{1}{2\lambda_1 + \lambda_p} \cdot \frac{v_1}{T} \cdot \dot{q} \quad (52)$$

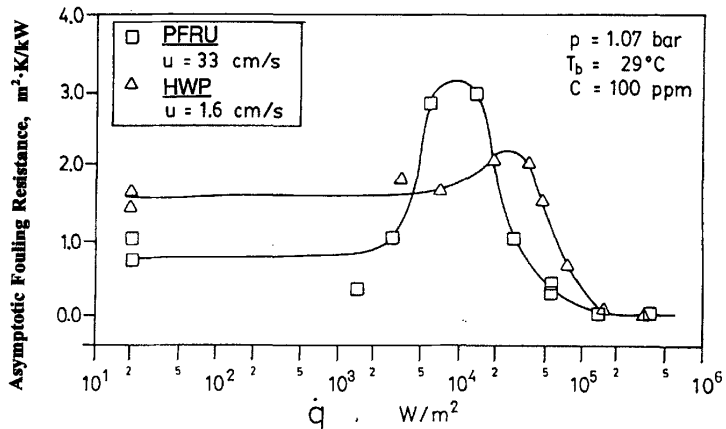


FIG. 33. Effect of heat flux on the asymptotic fouling resistance from Al_2O_3 /heptane suspensions. PFRU = annular probe, HWP = hot wire probe [56].

Hence

$$\dot{m}_d = \left(\beta - \frac{v_i}{2}\right)c_b - \left(\beta + \frac{v_i}{2}\right)c_w \quad (53)$$

E. EFFECT OF PARTICLE SIZE

Figure 34 shows the asymptotic fouling resistance as a function of the actual particle (or agglomerate) size, normalized with respect to the asymptotic fouling resistance and particle diameter for 1- μm particles [8].

With increasing particle diameter, the asymptotic fouling resistance decreases rapidly. The solid line fitted to the experimental data gives a slope of -1.6 in the double-logarithmic plot, which is significantly more than the prediction of the Watkinson and Epstein [52] model, also shown in this diagram. It should be emphasized that these results do not apply in cases where gravitational settling of particles contributes to the deposition process.

F. EFFECT OF SUSPENSION pH

Attachment and agglomeration of particles is mainly due to van der Waals forces and electrical double-layer forces. Electrical double-layer repulsion is due to the charges building up on the surface of metal oxides and hydroxides in aqueous solution. The sign and magnitude of these charges depend on the pH and

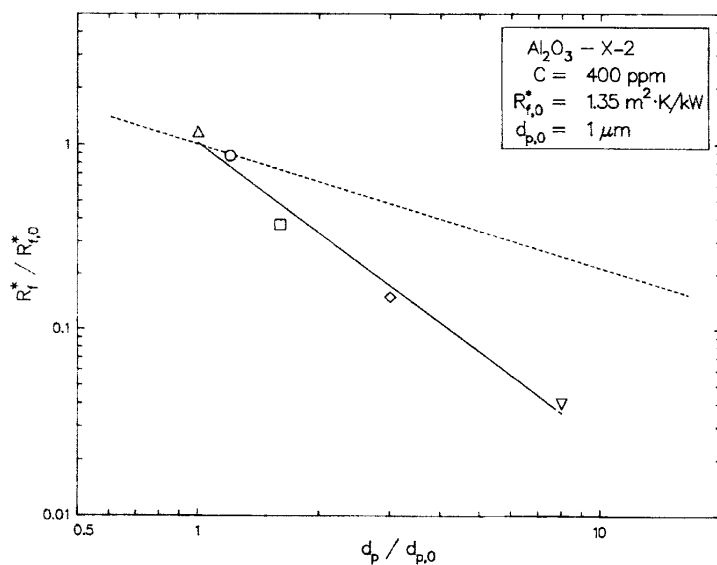


FIG. 34. Effect of particle diameter on the asymptotic fouling resistance from $\text{Al}_2\text{O}_3/\text{X}-2$ suspensions [8]. 0 indicates reference values, i.e. $d_{p,0} = 1 \mu\text{m}$.

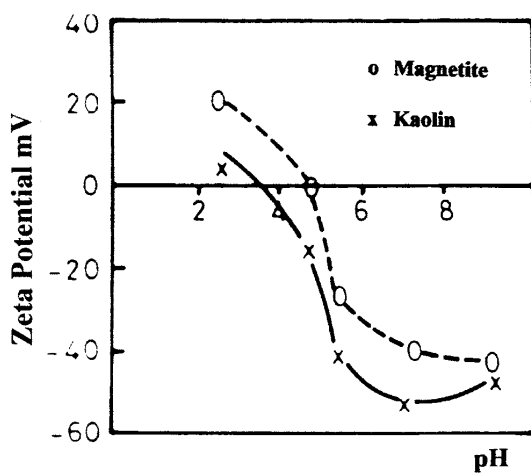


FIG. 35. Zeta potential of magnetite and kaolin suspensions in water [55].

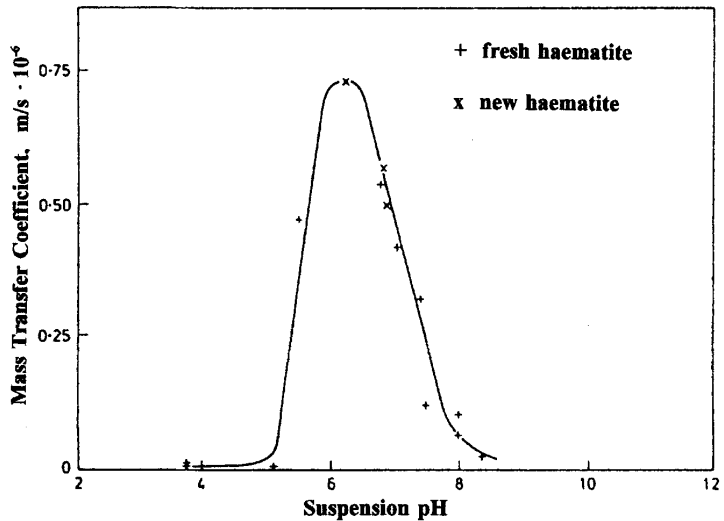


FIG. 36. Measured mass transfer coefficient against pH for small haematite particles depositing on stainless steel tubes at $Re = 1,000$ [58].

ionic content of the solution and the affinity of the surface for H_3O^+ and OH^- ions, which also affect the surface charge. The electrokinetic or zeta potential of particles in suspension can be measured from electrophoresis measurements. This is shown in Fig. 35 for magnetite and kaolin particles in water [55].

For a certain value of pH, the curves pass through the isoelectric point, where minimum repulsion between particles occurs. Obviously, this corresponds to the maximum potential for agglomeration, as shown in Fig. 36 from Williamson *et al.* [58].

VII. Biological Fouling

The following discussion on the effect of biological fouling on heat transfer is limited to microbial films, because they are most common in industrial cooling-water systems. These films or slimes are composed of a wide variety of microorganisms surrounded by a gelatinous film [59]. According to Characklis [60], the development of a microbial film takes place in six stages:

1. Adsorption of omnipresent glycoproteinaceous material to the heat transfer surface.
2. Diffusion of bacteria to the surface.
3. Attachment of the bacteria to the surface. Figure 37 shows the attachment of some bacteria, according to Marshall [61].
4. Growth of the film. This requires transport of nutrients from the bulk fluid for reproduction and production of extracellular polysaccharides.
5. Attachment of organic and inorganic particles to the film.
6. Removal of parts of the film by shear forces from the fluid.

The biofilms have physical properties similar to those of a static layer of water, i.e., a thermal conductivity of $0.55\text{--}0.7\text{ W/(m}\cdot\text{K)}$ as compared to $0.6\text{ W/(m}\cdot\text{K)}$ for water at 22°C .

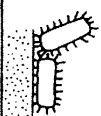
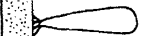
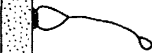
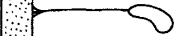
BACTERIUM	ATTACHMENT	
	FORM	ORIENTATION
 <u>PSEUDOMONAS</u>	PERMANENT	RANDOM
 <u>FLEXIBACTER</u>	PERMANENT	EDGE-TO-FACE
 <u>HYPHOMICROBIUM</u>	PERMANENT	EDGE-TO-FACE
 <u>CAULOBACTER</u>	PERMANENT	EDGE-TO-FACE
 <u>CYTOPHAGA</u> (GLIDING)	TEMPORARY	FACE-TO-FACE

FIG. 37. Attachment of bacteria on surfaces [61].

The amount of slime produced is directly related to the concentration of nutrients and oxygen available in the water. Abundant supply leads to large deposits that slough off readily, whereas poor supply generates dense, tenacious slimes [62]. Most microorganisms have optimum growth at pH values around 7, as shown by Hussain [64] in Fig. 38. This diagram also shows that the growth of biodeposits is increased by surface roughness, because of the larger effective contact area and stagnant flow zones in the wake of roughness elements.

The attachment and removal of bacteria at surfaces are strongly influenced by the fluid shear forces, which are a function of flow velocity and geometry. Above a critical shear stress, only a few bacteria can attach to the surface. Once bacteria are established on the surface, a significantly higher shear stress is required to remove the deposit. The growth of the biofilm increases with increasing flow velocity as a result of improved mass transfer of nutrients, until the shear stress becomes large enough to cause sloughing of parts of the film. Slimes formed at higher velocity are usually more adherent than those formed at low flow velocities. The effect of surface shear stress on the attachment of *Pseudomonas fluorescens* bacteria is shown in Fig. 39 [63].

Temperature has an important influence on the growth of microorganisms. Different microbial species have different temperature zones in which they grow preferably. At temperatures above 90°C, growth of microorganisms in heat exchangers is impossible. Figure 47 shows the effect of temperature on biological

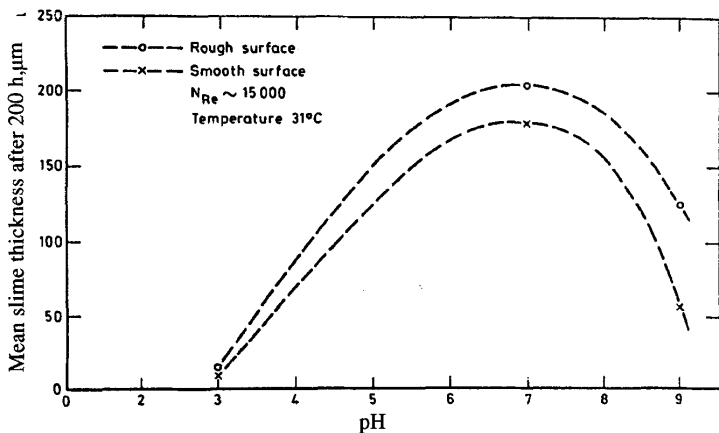


FIG. 38. Change in mean slime thickness with pH [58].

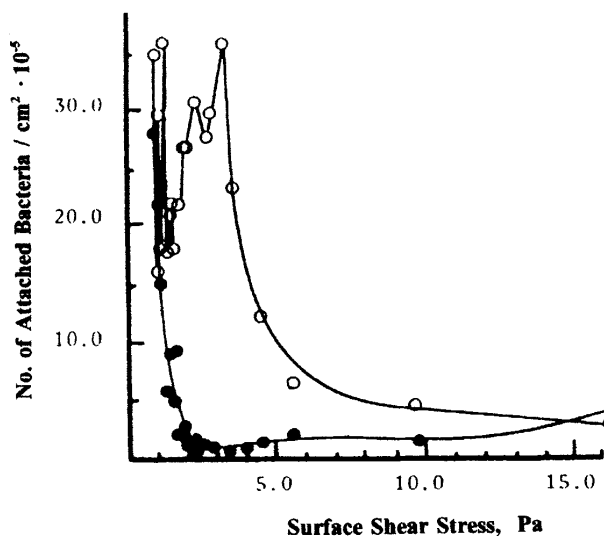


FIG. 39. Bacterial attachment versus surface shear stress. Full symbols, 48 h; open symbols, 168 h [63].

growth in an industrial heat exchanger. A pronounced maximum is found around 35°C.

VIII. Industrial Cooling-Water Fouling

A. SHELL-AND-TUBE HEAT EXCHANGERS

1. Fouling Resistances

Fouling resistances for industrial cooling-water applications are usually selected from the TEMA tables [3], an example of which is shown in Table VI. One of the many weaknesses of the TEMA tables is the fact that they differentiate only very approximately for the effects of water quality, flow velocity, and surface temperature. Figures 40 and 41 show the effect of these parameters on cooling-water fouling. HTRI has developed an empirical prediction procedure for the fouling resistance as a function of these parameters that is proprietary to members of this organization.

The pipe material itself may have a substantial influence on the fouling rate. This will be discussed in more detail in Sec. IX.B.2.

TABLE VI
TEMA FOULING RESISTANCES FOR WATER IN SHELL-AND-TUBE HEAT EXCHANGERS (m² · K/kW)

Temperature of heating medium	Below 120°C		120–200°C	
Water temperature	Below 50°C		Above 50°C	
	Flow velocity		Flow velocity	
Type of water	≤1 m/s	> 1 m/s	≤1 m/s	> 1 m/s
Seawater	0.09	0.09	0.18	0.18
Brackish water	0.35	0.18	0.53	0.35
Cooling tower				
Treated	0.18	0.18	0.35	0.35
Untreated	0.53	0.53	0.90	0.75
City water	0.18	0.18	0.35	0.35
River water				
Minimum	0.35	0.18	0.53	0.35
Average	0.53	0.35	0.70	0.53
Waste water	1.41	0.90	1.75	0.41
Cloudy or silty water	0.53	0.35	.75	0.53
Very hard water	0.53	0.53	0.90	0.90
Engine cooling water	0.18	0.18	0.18	0.18
Distilled water or condensate	0.09	.09	0.09	0.09
Treated boiler feed water	0.18	0.09	0.18	0.18
Boiler blow-down	0.35	0.35	0.35	0.35
Type of water	Fouling resistance			
Quench water	0.45			
Desalination plant (operating time 3 to 4 months)				
Polyphosphate treatment	0.15			
Acid treatment	0.08			
Brine	0.2–0.33			
Aqueous salt solutions	0.2–0.33			

From Ref. 3.

2. Effect of Fouling on Pressure Drop

The formation of deposits on the heat transfer surfaces causes an increase in the frictional pressure drop as a result of increased surface roughness and restricted cross-sectional flow area. According to Chenoweth [67], more heat exchangers are taken out of service because of excessive pressure drop than because of reduced heat transfer. A rough estimate of the tube-side pressure drop can be made if it is assumed that the deposit is distributed evenly over the tube inside. The frictional pressure drop in cylindrical tubes is calculated from Eq. (54):

$$\frac{\Delta p}{\Delta L} = \xi \frac{\rho u^2}{2d} \quad (54)$$

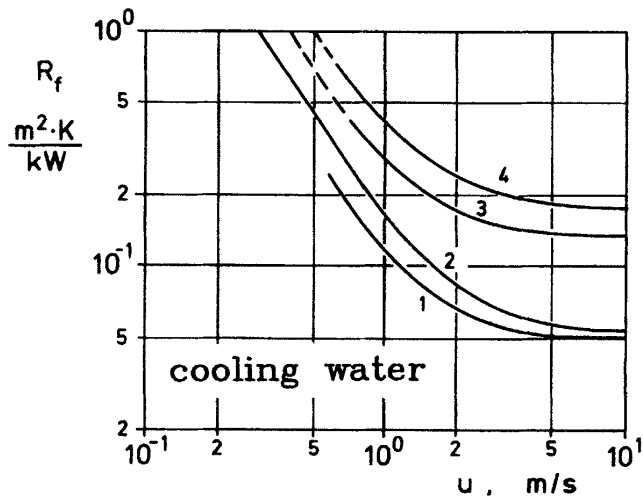


FIG. 40. Fouling resistance in shell-and-tube heat exchangers as a function of flow velocity and water quality [65]. Water quality decreasing from 1 to 4.

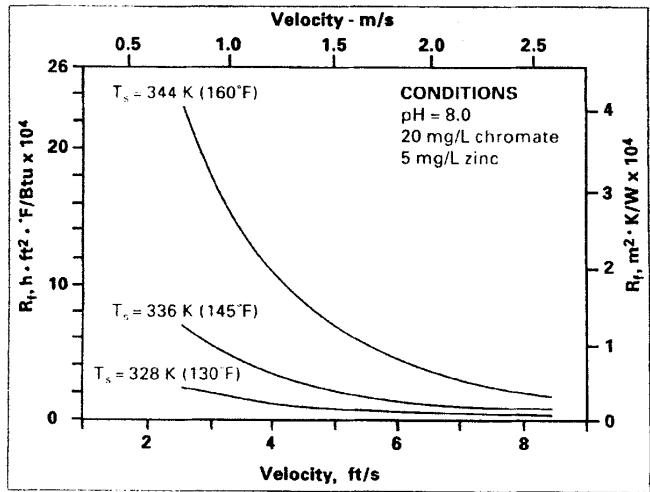


FIG. 41. Fouling resistance in shell-and-tube heat exchangers as a function of flow velocity and surface temperature [66].

The friction factor for smooth tubes is

$$\xi = 0.00056 + 0.5\text{Re}^{-0.32} \quad (55)$$

and that for rough tubes is

$$\xi = 0.014 + 1.056\text{Re}^{-0.42} \quad (56)$$

If the fouling resistance and the thermal conductivity of the deposit are known, the inside diameter of the fouled pipe can be determined from Eqs. (57) and (58):

$$R_f = \frac{d_i \ln(d_i/d_f)}{2\lambda_d} \quad (57)$$

$$d_f = d_i e^{-2\lambda_d R_f / d_i} \quad (58)$$

The pressure drop for the fouled tube is obtained by using Eq. (54) in conjunction with Eqs. (56) and (58).

The effect of fouling on the shell-side pressure drop can be estimated using Table VII, from Coulson *et al.* [68].

B. PLATE-AND-FRAME HEAT EXCHANGERS

1. Fouling Resistances

Plate-and-frame heat exchangers were originally developed for the dairy industry. However, their use is increasing rapidly in the chemical process industry, where they are beginning to replace tubular heat exchangers in several traditional applications. A typical example is the new BASF steam cracker in Antwerp,

TABLE VII
RATIO OF FOULED TO CLEAN SHELL-SIDE PRESSURE DROP

Deposite heat transfer coefficient	Shell diameter/baffle spacing		
	1.0	2.0	5.0
Laminar flow:			
$1/R_f = 6000 \text{ W}/(\text{m}^2 \cdot \text{K})$	1.06	1.20	1.28
$1/R_f = 2000 \text{ W}/(\text{m}^2 \cdot \text{K})$	1.19	1.44	1.55
$1/R_f < 1000 \text{ W}/(\text{m}^2 \cdot \text{K})$	1.32	1.99	2.38
Turbulent flow:			
$1/R_f = 6000 \text{ W}/(\text{m}^2 \cdot \text{K})$	1.12	1.38	1.55
$1/R_f = 2000 \text{ W}/(\text{m}^2 \cdot \text{K})$	1.37	2.31	2.96
$1/R_f < 1000 \text{ W}/(\text{m}^2 \cdot \text{K})$	1.64	3.44	4.77

From Ref. 68.

where several large plate-and-frame heat exchangers are used to transfer heat from the closed cooling-water cycle to brackwater. Plate heat exchangers are attractive because of their higher heat transfer efficiency, lower space requirements, easy accessibility to all areas, and, if expensive materials are required, lower costs [69, 70]. However, if severe fouling is anticipated, several of these advantages may not be valid any more. While there is plenty of information about the governing equations for clean operation [69, 71–76], information on fouling conditions is scarce.

As shown in Eq. (2), the percentage excess surface area increases with increasing clean heat transfer coefficient. This puts a heavy penalty on compact type of heat-exchangers such as plate-and-frame heat exchangers if, because of ignorance or because of cautiousness, the TEMA fouling resistances for shell-and-tube heat exchangers are used. Typical overall heat transfer coefficients for plate-and-frame heat exchangers are about $6000 \text{ W}/(\text{m}^2 \cdot \text{K})$, and those for shell-and-tube heat exchangers, about $1500 \text{ W}/(\text{m}^2 \cdot \text{K})$. As an example, a design fouling resistance of $0.15 \text{ m}^2 \cdot \text{K}/\text{kW}$ corresponds to a 22% overdesign for a shell-and-tube heat exchanger and to a 90% overdesign for a plate-and-frame heat exchanger. Most manufacturers of plate-and-frame heat exchangers recommend that the excess surface not exceed 25% of the heat transfer surface area calculated for the clean duty. The fouling resistances listed in Table VIII have been recommended by Marriott [77] for plate-and-frame heat exchangers.

TABLE VIII
FOULING RESISTANCES FOR PLATE-AND-FRAME
HEAT EXCHANGERS

Fluid	Fouling resistance ($\text{m}^2 \cdot \text{K}/\text{kW}$)
Water	
Demineralized or distilled	0.0009
Soft	0.017
Hard	0.043
Treated cooling-tower water	0.034
Coastal seawater	0.043
Ocean seawater	0.026
River water	0.043
Engine jacket	0.052
Lube oil	0.017–0.043
Vegetable oil	0.017–0.052
Organic solvents	0.0009–0.026
Steam	0.009
General process fluids	0.009–0.052

From Ref. 77.

2. Increase In Pressure Drop Due to Fouling

The estimation procedure for the pressure drop under fouling conditions outlined in Sec. VIII.A.2 does not apply for plate-and-frame heat exchangers. Because of the nonuniformity of flow distribution and deposit formation, measured pressure drop increases are significantly higher than the values predicted using the measured fouling resistances (see Fig. 42).

3. Effect of Process Parameters on Fouling

Most plate-and-frame heat exchangers use water as the cooling medium. In cooling water, dissolved CaCO_3 and CaSO_4 are common, and scaling occurs if the solubility of these salts is exceeded for conditions at the heat transfer surface. The important factors for this type of fouling are temperature, concentration of salts, pH value, and Reynolds number. Scaling is reduced by appropriate water treatment, but two important design features of plate heat exchangers also contribute to the reduction of fouling: the high turbulence level (and thus high shear stresses) and the absence of low-velocity zones.

Cooper *et al.* [78] investigated cooling-water fouling using an APV model R405 plate heat exchanger with APV type R40 stainless steel plates. Two small stainless steel shell-and-tube heat exchangers were installed in parallel for comparison. The water was chemically treated before entering the heat

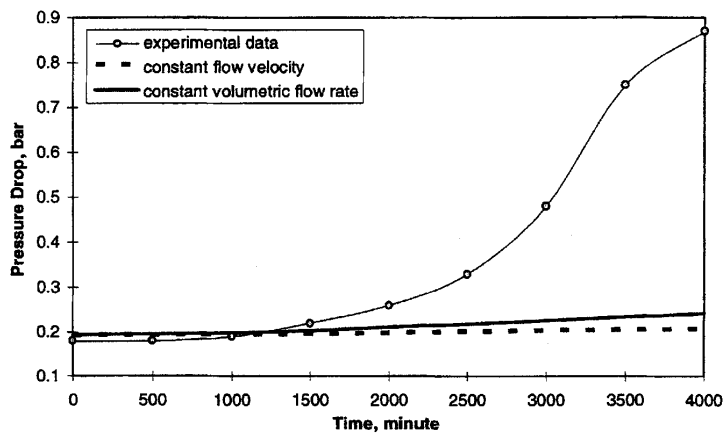


FIG. 42. Measured and calculated pressure drop in plate-and-frame heat exchangers during CaSO_4 scale formation.

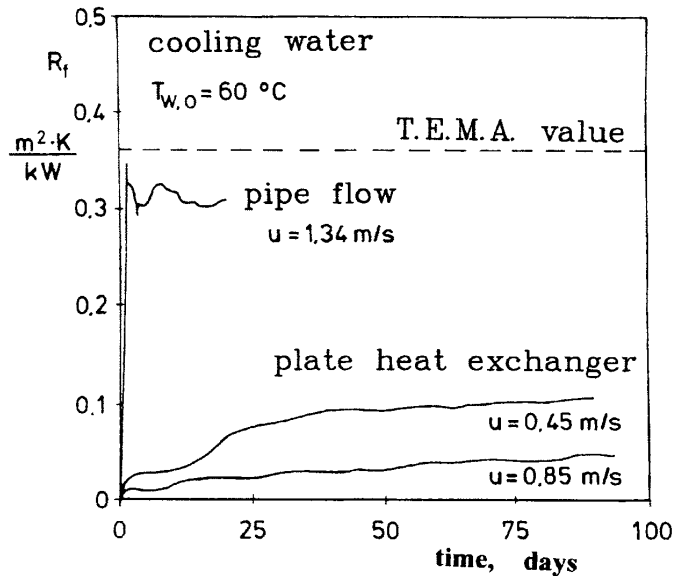


FIG. 43. Cooling-water fouling in a plate-and-frame heat exchanger [78].

exchangers. The various stages of the water treatment and the characteristics of the cooling water are reported. Some of the important results of this investigation are given in Fig. 43.

The fouling resistance in the plate-and-frame heat exchanger is significantly lower than that in the shell-and-tube heat exchanger, despite the typically lower flow velocities. If the flow velocity is increased, the fouling resistance decreases overproportionally. With increasing time of operation, the fouling resistance seems to level off and to approach an asymptotic value. This is also shown in Fig. 44, which shows the asymptotic value as a function of the surface temperature halfway up the plates. Especially at low flow velocities, increasing the surface temperature results in a significant increase in fouling.

Analysis of deposits formed during cooling-water fouling in the experiments by Cooper *et al.* [78] showed that the amount of silica (sand) found at the shell side of the shell-and-tube heat exchanger was significantly higher than that for the parallel plate heat exchanger. The high silica concentration was caused by the sedimentation in the stagnant flow zones at the corners between baffles and shell, whereas the high turbulence promoted by the corrugated plates in the plate heat

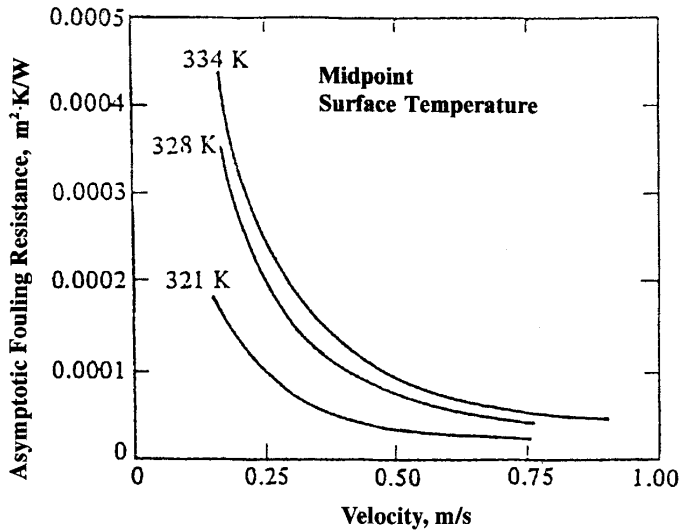


FIG. 44. Fouling resistance as a function of flow velocity and surface temperature [78].

exchanger minimized sedimentation fouling. The smooth surface of the plates further reduces the stickability of the particles.

To investigate cooling-water fouling in plate heat exchangers, Novak [79] studied the fouling behavior of Rhine River water near Ludwigshafen (Germany) and of Öresund water in Sweden. For both waters, mainly biological fouling was observed. Figure 45 shows the measured fouling resistances as a function of time for various flow velocities. In most cases, the fouling resistance increased almost linearly over the observed period of time.

Figures 46 and 47 show the influence of heat transfer surface temperature and of the wall shear stress on the buildup of biological deposits at the heat transfer surface. Table IX summarizes the measured fouling rates.

C. APPROXIMATE INFLUENCE OF PROCESS PARAMETERS ON INDUSTRIAL HEAT-EXCHANGER FOULING

Although some deviations exist, the following approximate influence of process parameters on industrial fouling has been found:

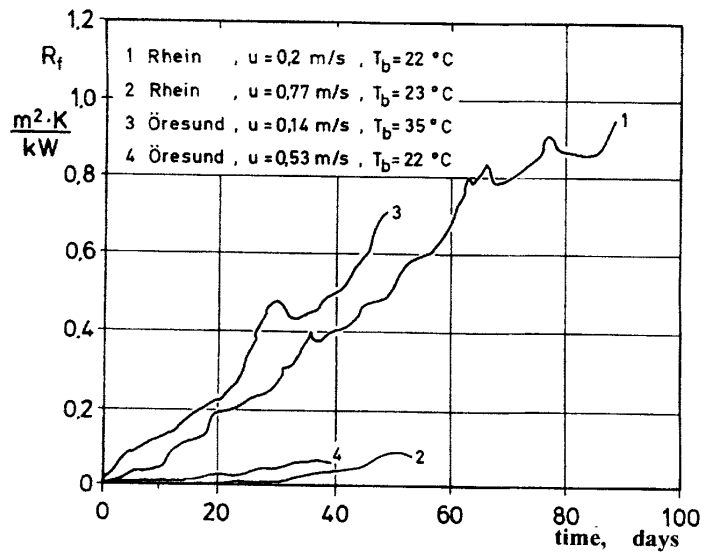


FIG. 45. Fouling resistance as a function of time [79].

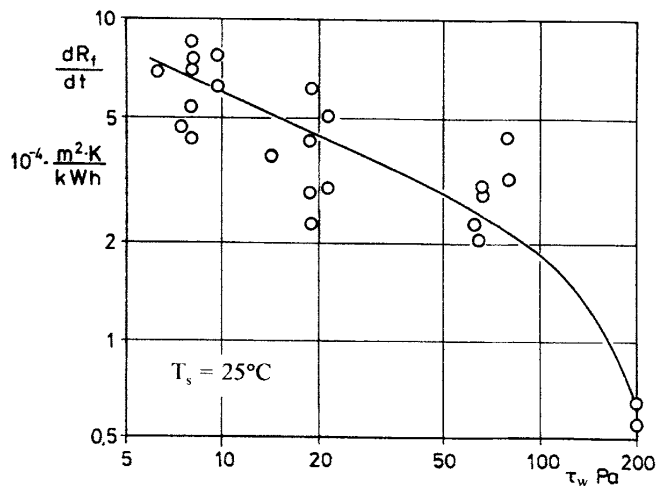


FIG. 46. Effect of flow velocity on biofouling in plate heat exchangers [79].

TABLE IX
FOULING RATES OF RHINE RIVER WATER FOR A SURFACE TEMPERATURE OF 25°C

Type	$u(\text{m/s})$	$\tau(\text{Pa})$	$dR_r/dt(10^4 \text{ m}^2 \cdot \text{K/kWh})$
Plate heat exchanger	0.13	6.7	7.4
Plate heat exchanger	0.19	14.5	4.3
Plate heat exchanger	0.77	190.0	0.6
Spiral plate exchanger	0.43	7.5	5.0

From Ref. 79.

1. Fouling usually increases linearly with increasing foulant concentration in the fluid bulk.
2. The fouling resistance nearly always decreases with increasing wall shear stress as a result of increased removal forces. As an average, it was found that the fouling resistance is inversely proportional to the flow velocity.
3. For all fouling mechanisms, the fouling resistance increases exponentially with increasing surface temperature according to an Arrhenius term [Eq. (6)]. For biological fouling, maximum fouling is observed for temperatures around 35°C.
4. Fouling was found to increase with increasing roughness of the heat transfer surface. Figure 48 shows measurements for fouling from Kraft black liquor in a standard pipe and in an electropolished pipe.

To date, not even these simple rules are used in the design of heat exchangers, even though they could significantly improve some heat-exchanger optimization procedures. This is demonstrated in Fig. 49, which shows the sum of annual operating and capital service costs of an arrangement of multiple double-pipe heat exchangers as a function of the flow velocity in the pipes. One curve has been calculated according to Martin [80] for a constant fouling resistance, the second is for the case where the fouling resistance is velocity-dependent, as suggested above. The optimum flow velocity shifts from 0.8 to about 1.3 m/s, and the total annual costs are reduced by 10% despite the higher friction losses.

The use of constant fouling resistances in the design of heat exchangers leads to initially oversized equipment. Heat duties in new or cleaned heat exchangers can, therefore, be considerably higher than the design specifications. In most chemical processes, however, product inlet and outlet temperatures, product flow rate, and cooling-water inlet temperature are specified. If this is the case, the heat exchanger is usually controlled via the flow rate of the cooling water. To reduce the heat duty, the water flow velocity must be reduced. Using the above approximate relationships, Fig. 50 shows that this procedure may cause a

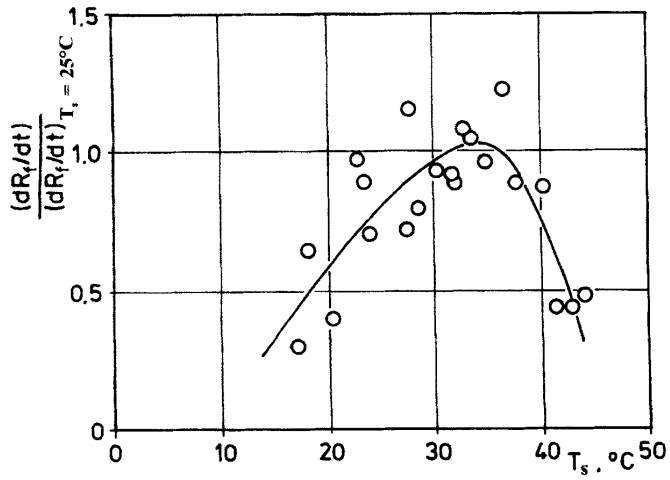


FIG. 47. Effect of surface temperature on biofouling in plate heat exchangers [79].

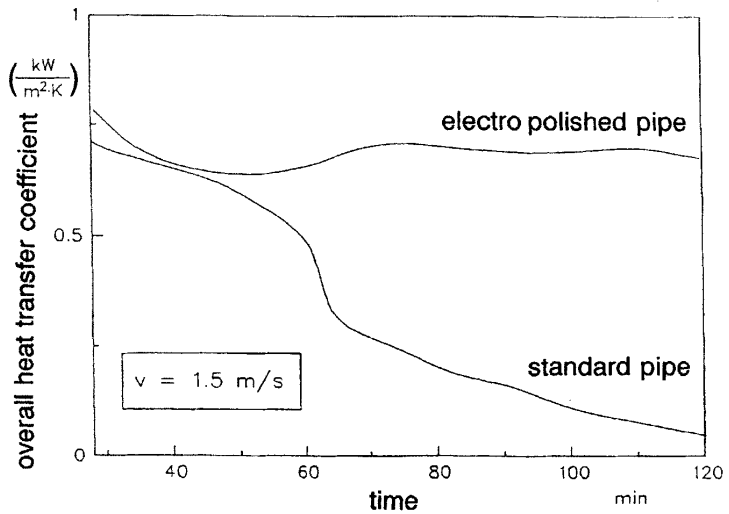


FIG. 48. Reduction of heat transfer coefficient during the evaporation of Kraft black liquor in standard pipes and electropolished pipes.

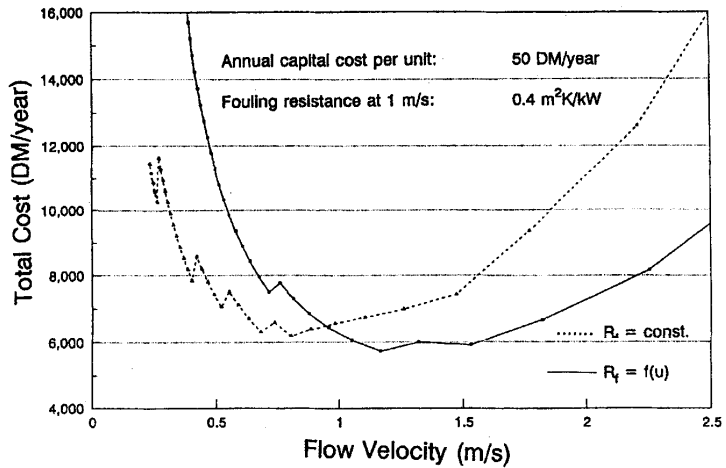


FIG. 49. Total annual costs of a double-pipe heat-exchanger arrangement as a function of the flow velocity.

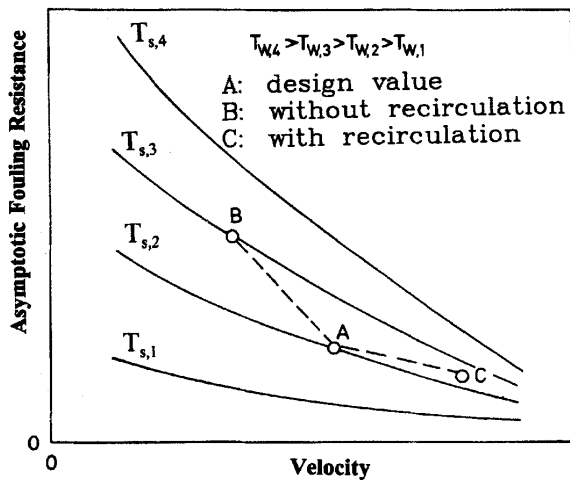


FIG. 50. Fouling resistance as a function of flow velocity and surface temperature during start-up of a new or cleaned heat exchanger.

TABLE X
COOLING-WATER FLOW VELOCITIES IN CONDENSERS

Pipe material	Recommended velocity	Minimum velocity
Arsenical copper	< 1.5 m/s	1.0 m/s
Admiralty	1.4–2.0 m/s	1.0 m/s
Aluminum brass	1.8–2.2 m/s	1.0 m/s
90/10 cupro-nickel	1.8–2.5 m/s	1.5 m/s
90/30 cupro/nickel	2.4–3.5 m/s	.8 m/s
Cu	1.5–2.0 m/s	1.0 m/s
Steel	2.0–4.0 m/s	1.0 m/s

From Ref. 81.

considerable increase in fouling as compared to fouling under design operating conditions.

Point A identifies the design values of flow velocity and heat transfer surface temperature. As the heat exchanger is initially overdesigned, the cooling-water flow velocity is throttled, which also causes an increase in the heat transfer surface temperature (point B). However, fouling at B is considerably worse, and deposits created during this part of the operation may not be removed completely, even if the flow velocity is increased later. Therefore, if high fouling resistances are specified, fouling may become a self-fulfilling prophecy. If part of the cooling water is recirculated, the flow velocity and the cooling-water inlet temperature can be increased to meet the required heat duty. The anticipated fouling at (C) will be similar to the design value (A), but a price will have to be paid to provide the higher flow velocity.

Fouling is heavier in regions with low flow velocity. While heat exchangers used to be designed for flow velocities around 1 m/s, Epstein [9] claims that modern design velocities are about 1.8 m/s. Table X shows optimum and minimum cooling water velocities for condensers of various pipe materials, according to Paikert [81].

IX. Mitigation of Cooling-Water Fouling

This section provides an overview of on-line chemical and on-line mechanical techniques used to mitigate water-side fouling.

A. CHEMICAL METHODS

Since about 1920, a number of companies have specialized in the mitigation of liquid-side fouling and corrosion, mainly for the flow of cooling water. These

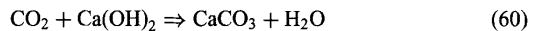
companies have gained considerable expertise and have developed a wide range of additives and equipment. Services include supplying chemicals as well as analyzing cooling water, estimating fouling and corrosion problems to be anticipated, and providing complete cooling-water treatment programs, including continuous monitoring of the system. In this subsection, several methods to reduce fouling by chemical means are discussed. For the final selection of the treatment as well as for the dosage of treatment chemicals, specialists should be consulted.

Because of environmental regulations and decreased water availability, the amount of water discharged from a cooling system has often been reduced. Therefore, systems are operating at higher concentration cycles, and any liquid discharge from the system is being eliminated if possible.

1. *Scaling*

Generally, there are three alternatives available to mitigate or to prevent scale deposition due to high concentration of scale-forming ions. These alternatives are removal of scaling species, pH control, and the use of scale inhibitors.

a. *Removal of Scaling Species* Scaling species may be removed by ion exchange and by chemical treatment. In the latter treatment, carbonic acid and calcium hardness are removed by the addition of chemicals. If the lime treatment is used,



During slow decarbonization (1 to 3 h reactor residence time), the calcium carbonate precipitates as silt; during fast decarbonization (5 to 10 min reactor residence time), it precipitates in the form of particles. Except for installations with high calcium hardness or large throughput, chemical removal of scaling species is not used anymore.

Instead, ion exchangers are used, in which the “harmful” scaling species in the fluid are replaced by “harmless” ions (for example, Ca^{++} or Mg^{++} by Na^+). Ion exchangers are usually manufactured from styrene-based polymers. The so-called cationic exchangers contain weak and strong acids; the anionic exchangers contain weak and strong alkaline groups. With these two variations, all cations and anions can be removed from the fluid. Ion exchangers have to be regenerated regularly with the appropriate salt solution. According to Dubbel [82], chemical decarbonization leaves a residual hardness of 17 to 30 ppm as CaCO_3 , whereas ion exchange can reduce the hardness to 2 ppm as CaCO_3 . Both methods of fluid treatment have high capital and operating costs.

b. pH Control The solubility of scale-forming constituents increases with decreasing pH. Many treatment programs, therefore, involve the addition of acid (usually H_2SO_4) to the system to maintain a pH in the region of 6.5 to 7.5. If the system contains corrosion-resistant materials, a pH at which no scaling will occur may be selected. The Langelier Saturation Index or the Ryznar Stability Index (see Sec. V.A) are commonly used to determine the value to which the pH is to be adjusted.

c. Scale Inhibitors Growth of crystals or the nucleation of crystals can be inhibited by the addition of scale inhibitors. Many proprietary compounds for scale control are available.

Chelating agents (for example, EDTA) complex strongly with the scaling cations and hence inhibit their deposition at the heat transfer surface. Inhibitor and scalant must be available in stoichiometric ratios.

Processes that are based on physical rather than on chemical reactions stabilize supersaturated solutions by adsorption at the crystal nuclei (for example, polyphosphates) or modify or weaken the crystalline structure (for example, polycarboxylic acid). The effect of various additives on the shape of CaCO_3 deposits is shown in Fig. 51 [83, 84]. Table XI, which has been adopted from Harris and Marshall [85], shows the ability of additives to maintain CaCO_3 in solution. Lists of additives to reduce crystallization from hard waters have been compiled by Harris and Marshall [85] and by Krisher [86].

2. Particulate Fouling

Particulate fouling is usually mitigated by the addition of surfactants or dispersants. If the surface tension is reduced, large particle agglomerates can

TABLE XI
ABILITY OF VARIOUS ADDITIVES TO MAINTAIN CaCO_3 IN
SOLUTION

Additive	% inhibition at dose level			
	2.5 ppm	5.0 ppm	7.5 ppm	10 ppm
Polyphosphate	98%	8%	99%	100%
Aminophosphonic acid	7%	96%	95%	94%
Acetodiphosphonic acid	83%	65%	84%	93%
Polyacrylate	30%	65%	84%	93%
Polymaleic acid	26%	35%	44%	56%
EDTA	15%	20%	20%	20%

From Ref. 85.

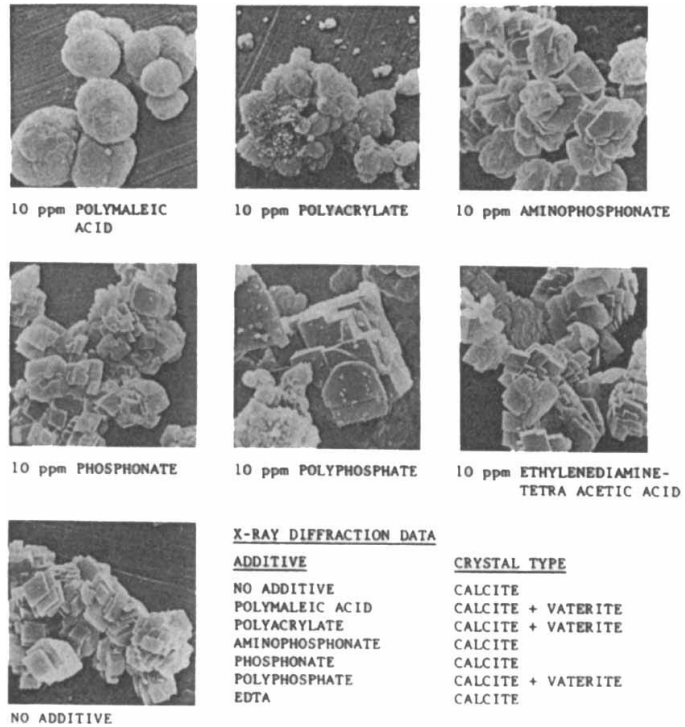


FIG. 51. Micrographs of CaCO_3 deposits in the presence of additives [83].

break down into smaller particles, which tend less to sedimentation. Dispersants impart like charges to both the heat transfer surface and the particles and reduce deposition. For cooling-water applications, polyacrylates or polysulfonates with molecular weights between 2000 and 3000 g/mol are used. According to Krisher [86], the addition of polyphosphates to reduce scaling may cause a slight reduction in the dispersion of particulates.

3. Corrosion Fouling

Generally, it is desirable to have a thin, passivating oxide layer on the heat transfer pipes. Epstein [9] mentions that this oxide layer is removed if the flow

velocity exceeds 30 m/s. Excessive corrosion can be controlled by the addition of corrosion inhibitors (chromate or polyphosphate based) or by control of the pH. Chromate is a highly efficient and cost-effective inhibitor. However, the toxicity of chromates in the environment has restricted their use. This also holds for zinc-based inhibitors.

4. Biofouling

Biological growth is usually controlled by the addition of biocides. In recent years, chlorine has been the most widely used. It reacts with water to form hydrochloric and hypochlorous acids:



Hypochlorous acid is an extremely powerful oxidant that easily diffuses through the cellular walls of microorganisms. Various studies [79, 87, 88] have assumed that HOCl oxidizes the active sites of certain enzyme sulfhydryl groups that constitute intermediate steps in the production of adenosine triphosphate (ATP). The system ATP-ADP allows conversion of carbohydrates and hence the energy supply for living organisms. Unlike other, nonoxidizing biocides, chlorination also weakens the biofilm matrix, allowing the removal of biofilms from the heat transfer surface by fluid shear forces. Continuous application of chlorine at concentrations between 0.1 and 0.5 ppm has been shown to be a reliable but costly method to avoid deposition. Cheaper but less effective is a dosage of 1 to 10 ppm for 15 min at intervals of 4 h. However, it was found that biofilm growth is accelerated after a shock chlorination (see Fig. 52).

For higher values of pH, HOCl dissociates as shown:



Grier and Christensen [89] found that OCl^- is considerably less effective as a biocide than HOCl. Biological fouling control with chlorination has the disadvantage that chlorine has to be added continuously, since it reacts not only with microbes but also with process contaminants such as H_2S or NH_3 . Chlorine concentrations exceeding 0.5 ppm in water may give rise to corrosion problems, especially for stainless steel equipment. Because of the biocidal action of chlorine, the effluent chlorine concentration is limited to 0.2 ppm. For these reasons, chlorine is increasingly being replaced by other chemicals such as methylene-thiocyanate or chlorophenols (see Waite and Fagan [90]).

Even though the addition of hypochlorous acid is effective in preventing the growth of a wide range of bacteria and algae, there are a number of species that can be controlled only by excessively high HOCl concentrations (>30 ppm). Since this may cause operational problems, compounds that can be added to the chlorinated water have been developed to eliminate these species. Grade and

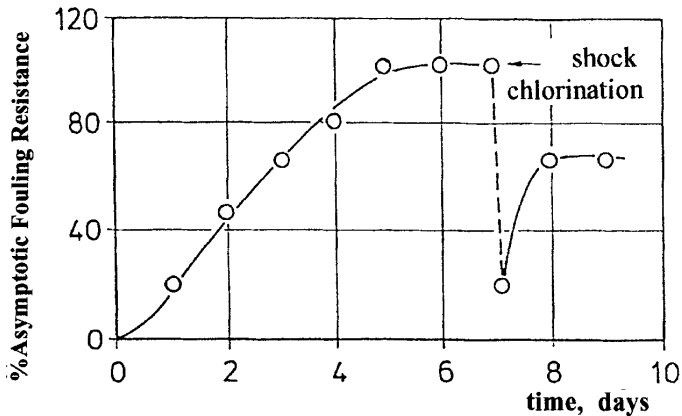


FIG. 52. Effect of shock chlorination on the growth of biological material at heat transfer surfaces [89].

Thomas [91] discuss treatment programs that are effective against bacteria and algae. Generally, it is recommended that biocide treatments be varied regularly to avoid immunization of microorganisms.

Because of the toxic effect of copper ions on biological matter, another method to reduce bacterial growth is the use of piping with a copper content over 60% or the addition of copper sulfate to the water. For potable water, the copper concentration must be below 1 ppm.

B. MECHANICAL METHODS

For most liquid-side applications, it is difficult to design mechanical on-line fouling mitigation methods. Nevertheless, a number of mitigation techniques have been developed. They are generally based on one of the following mechanisms:

1. Short-time overheating of the heat transfer surfaces. The different thermal expansivity of tubes and tube deposits may cause cracking of the deposit.
2. Mechanical vibration of heat transfer surfaces.
3. Acoustical vibration of heat transfer surfaces.
4. Increased shear stress at the fluid-deposit interface.
5. Reduced adhesion of deposits

Most of the commonly used liquid-side fouling mitigation techniques have been developed for the tube-side liquid.

1. Reversal of Flow Direction

Regular reversal of the flow direction in conjunction with a short-term increase in the flow velocity is sometimes used to mitigate the formation of weak deposits. Figure 53 shows that this procedure reduces the fouling resistance, but only for a short period of time. A much better result could be achieved by operating at a higher flow velocity.

2. Modification of Heat Transfer Surface

Surface roughness increases the contact surface area, and so the true contact area is much larger than the apparent surface area. As a result of this difference, a system that has a rough surface has a greater effective surface energy than the same system with a smooth surface. It then follows that stronger adhesion should occur on the rough surface, since it would undergo a greater decrease in effective surface energy. This is confirmed by measurements with Kraft black liquor in electropolished tubes (see Fig. 48).

The pipe material itself can also have an effect on cooling-water fouling. Biofouling is reduced in brass tubing, as shown in Fig. 54 for seawater fouling.

The stickability of deposits on heat transfer surfaces is directly related to the reduction in Gibbs energy of the surface. In general, maximum adhesion occurs in systems undergoing a maximum decrease in surface energy, such as occurs when a low-energy fluid spreads upon a high-energy surface [95, 96]. Although

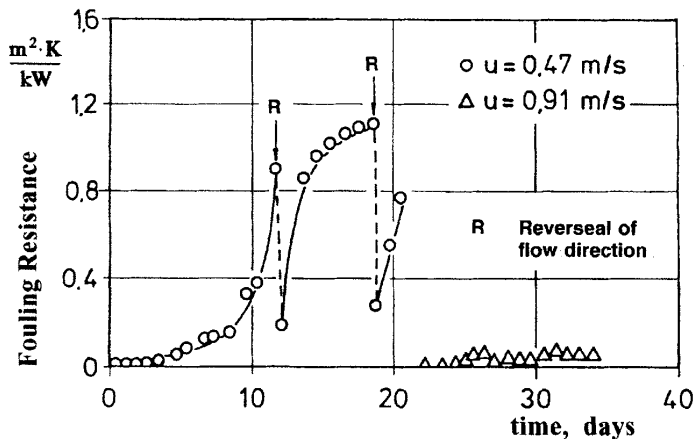


FIG. 53. Continuous cleaning by reversal of flow direction [93].

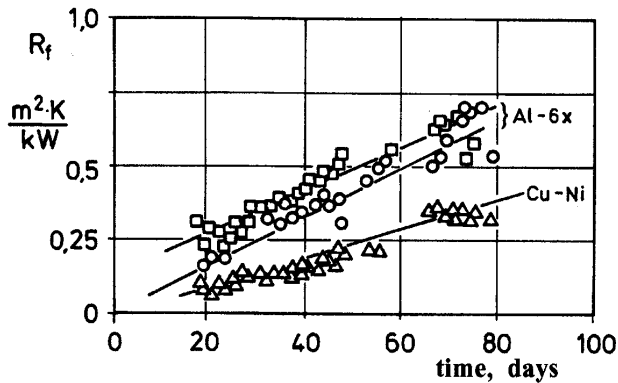


FIG. 54. Influence of pipe material on biofouling [92].

this principle was developed for liquids on solids, it should be applicable for solids upon solids, too. Therefore, poorest scale adherence should occur on materials that have very low surface energies. Hardly any information is available on surface energies of metallic surfaces in the solid phase. Of the conventional metallic heat transfer surface materials, it appears that the order of decreasing surface energies is probably as follows:

1. Stainless steel
2. Monel
3. 70–30 copper–nickel
4. 90–10 copper–nickel
5. Titanium

Surface energies for these materials range from 1800 to 1300 erg/cm². However, the surface crystallographic structure, the presence of small quantities of impurities in the metal, the roughness of the surface, and the presence of oxide on the surface may alter the effective surface energy. Organic materials such as Teflon have surface energies of less than 20 erg/cm² [96]. Teflon and Sākaphen surfaces have indeed been shown to reduce fouling from various fluids—for example, during seawater evaporation and heat transfer to Kraft black liquor (see Fig. 55) [97].

The main reason why such materials or coatings are not more widely used is that they are poor heat conductors and form an additional resistance to heat transfer that is comparable to the TEMA fouling resistance for cooling water. If

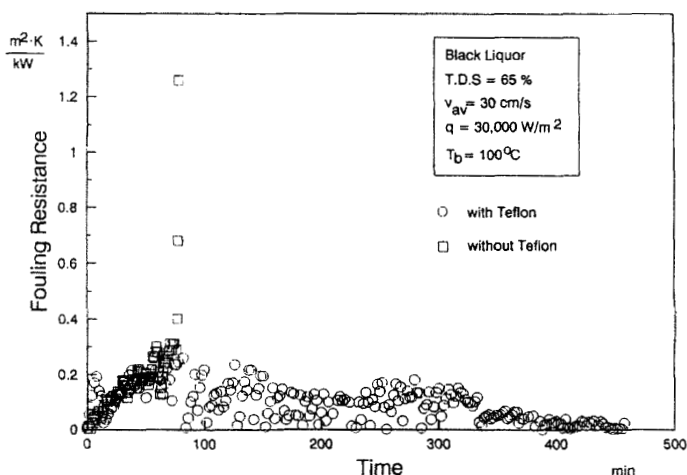


FIG. 55. Reduced fouling from Kraft black liquor on PTFE-coated surfaces [97].

very thin coatings were used, the resistance to erosion or other mechanical stress would be greatly diminished.

A novel method of producing low-energy heat transfer surfaces has been suggested by Zhao and Müller-Steinhagen [98–101]. Surface alloys or extremely thin ($<2\mu\text{m}$) and adherent coatings have been produced by ion bombardment using ion-beam implantation, magnetron sputtering, multiarc ion plating, filtered cathodic vacuum arc plating, and similar techniques. CaSO_4 scaling can be reduced by more than 70% for pool boiling (see Fig. 56) and almost eliminated for convective heat transfer (see Fig. 57).

The adhesion of bacteria is also reduced by the above surface treatment. Table XII shows that, depending on the treatment method, reduction of up to 99% has been observed for thermophilic streptococci.

The University of Surrey (U.K.) has been awarded a patent for the above findings [101].

3. Turbulence Promoters

According to Mascone [102], tube corrugations and tube inserts can increase the plain tube heat transfer coefficient by a factor of 2 to 15. This is achieved by reducing the average thermal boundary-layer thickness. As deposition rates for most fouling mechanisms are inversely dependent on fluid wall shear stress and

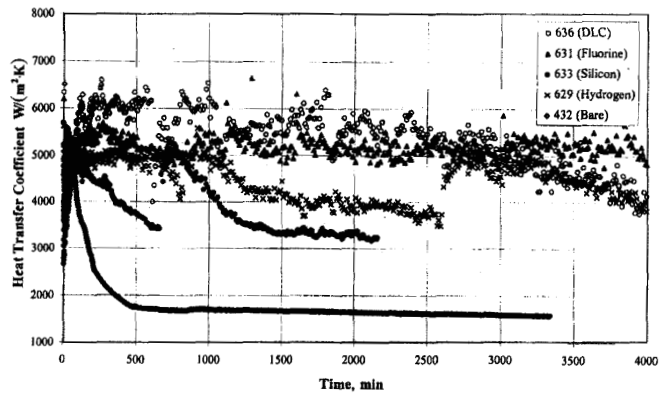


FIG. 56. CaSO_4 scaling during pool boiling on treated stainless steel heat transfer surfaces [98].

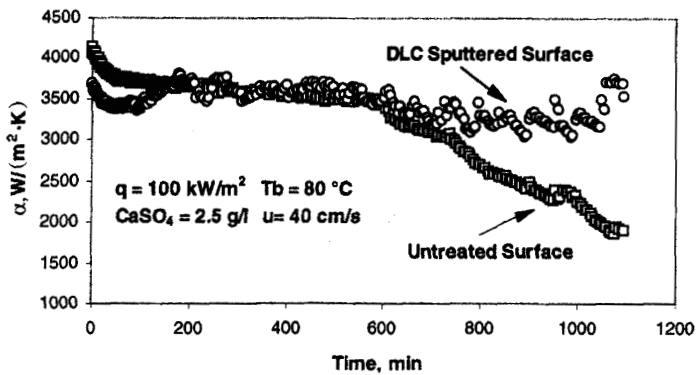


FIG. 57. CaSO_4 scaling during convective heat transfer on treated stainless steel heat transfer surface.

TABLE XII
ADHESION OF THERMOPHILIC STREPTOCOCCI

Sample no.	Ions sputtered/implanted	Number of cells (cells/cm ²)	Reduction (%)
0	None (Control)	501,187	—
1	DLC—sputtered	100,000	0
2	Cr_2O_3 —sputtered	100,000	80
3	CrN—sputtered	3,126	99
4	CrC—sputtered	12,589	97
5	Si—implanted	79,432	84

heat transfer surface temperature, reduction of the viscous and thermal sublayer thickness may also considerably reduce fouling. The increased thermal efficiency of pipes equipped with turbulence promoters is, however, accompanied by increased pressure drop. Additionally, major cleaning problems may occur if the pipes get fouled despite the inserts. Typical inserts are sphere matrices (see Fig. 58), coils (SPIRELF system, Fig. 59), and wire inserts (see Fig. 60).

Figure 61 shows the reduction of fouling from Chicago city water after installation of sphere matrix inserts [102].

Figure 62 shows the performance of a tar preheater for carbon black production, before and after installation of Cal Gavin radial mixing elements. In

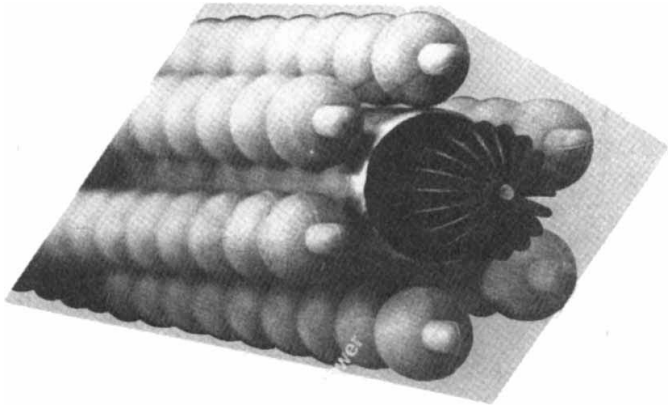


FIG. 58. Sphere matrix insert to increase heat transfer for flow in pipes [102].

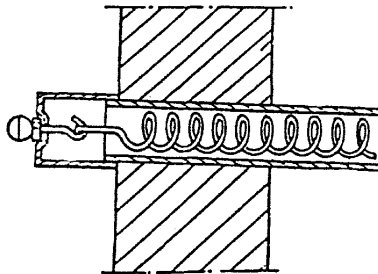


FIG. 59. Spiral insert (SPIRELF system).

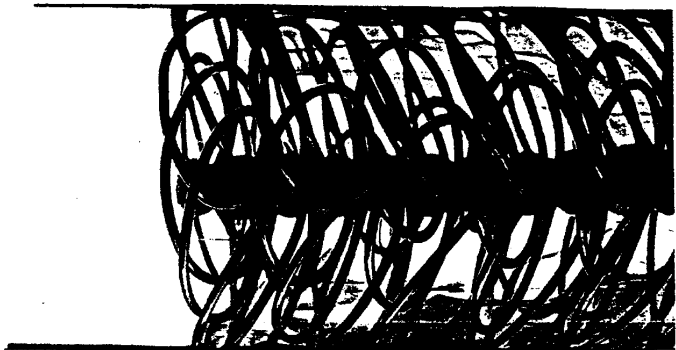


FIG. 60. Cal-Gavin wire insert to increase heat transfer for flow in pipes.

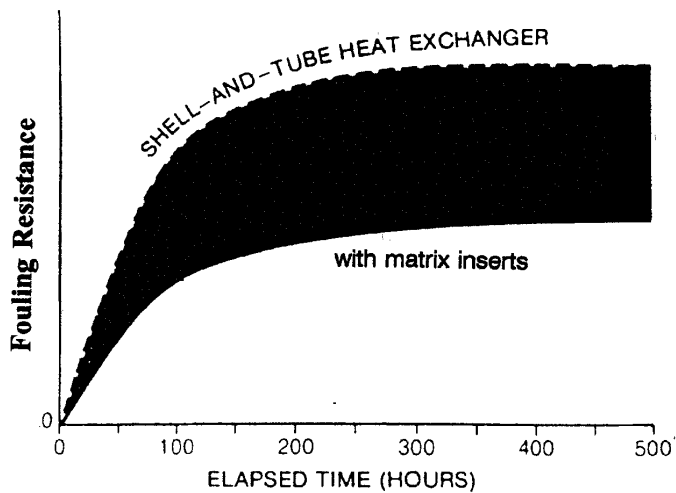


FIG. 61. Fouling mitigation with sphere matrix inserts [102].

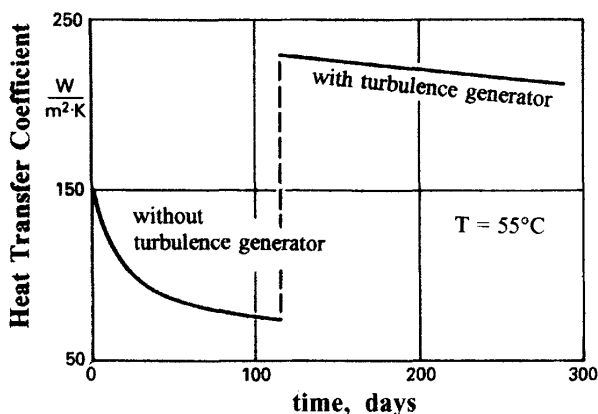


FIG. 62. Fouling mitigation in a tar preheater with Cal-Gavin inserts [105].

addition to the increased heat transfer coefficient, the almost complete absence of fouling should be noted [105]. Similar inserts have been installed in crude oil preheaters, water-cooled heat exchangers, and aromatics process heat exchangers. Preliminary investigations have indicated that the Cal Gavin inserts may also be used to reduce biofouling in heat-exchanger tubes [106].

The SPIRELF system, shown in Fig. 59, has been developed for crude oil preheating. A flexible stainless steel wire coil is inserted into each tube in a way that allows a self-driven rotation. The effect of these inserts is a claimed reduction in fouling resistance of between 75 and 95%. The payback time is only a few months.

4. Continuous Transport of Cleaning Devices through Tubes

These methods require major modifications of the flow system and are, therefore best implemented in the design stage. However, they have the advantage that exchangers may be kept clean over long periods of time. All systems work best if they are applied to an initially clean heat exchanger.

A number of companies (MAN, Water Services of America, KALVO [107], ATCS) have developed continuous tube-cleaning systems using small nylon brushes that are inserted into each tube (see Fig. 63). These brushes are pushed through the tubes by the fluid flow. For continuous operation and optimum cleaning efficiency, the flow direction has to be reversed about every 8 h. Life expectancy of the brushes is about 5 years. Typical applications for the nylon

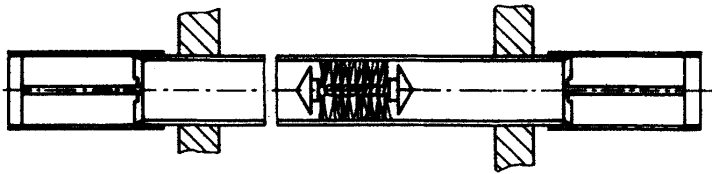


FIG. 63. Continuous cleaning with wire brush system.

brush system are cooling-water duties in condensers or chillers. The 1974 cost for installing the system in a 2-MW chiller was U.S.\$13,700. KALVO claims that the time for amortization is between 8 and 16 months. Water Services of America offers a modification of this system that can be applied for high-temperature process fluids.

Although there are many examples of successful application of the brush tube-cleaning system, a comparison of the performance of 52 power stations in the Federal Republic of Germany that were equipped with continuous tube-cleaning systems shows that the brush cleaning system may fail for hard deposits [108]. Better results were obtained with a system in which sponge balls with a rough surface were circulated through the heat exchangers (see Fig. 64).

The diameter of the sponge balls is slightly larger than the inside diameter of the tubes and the system is designed so that each tube sees a sponge ball every 5 to 10 min. Because the diameter of the sponge balls decreases with time and because of inevitable ball losses through the screening system, the sponge balls have to be replaced regularly. For hard and adherent deposits, carborundum-coated sponge balls can be used.

Presently, more than 90% of West-Germany's turbogenerators above 100MW output are equipped with the TAPROGGE sponge ball system. TAPROGGE [109, 110] claims that sponge ball systems are installed at more than 500 locations in the United States.

While the solids concentration in cooling-tower cycles can be tripled under normal circumstances, application of continuous sponge ball cleaning systems allows for concentration by a factor of 6 to 10. The application of sponge ball systems is limited to temperatures below 120°C. According to the manufacturers, application of sponge ball systems reduces the fouling resistance to close to $0 \text{ m}^2 \cdot \text{K/kW}$ (see Fig. 65). Eimer [110] claims that the time for amortization of the systems is about one year.

On-line cleaning systems may require considerable maintenance. They are not effective against stones, clamshells, etc., and upstream devices are needed to remove debris and macroscopic organic matter from the incoming water. If these systems are not maintained properly, excessive ball losses may occur. Strauss

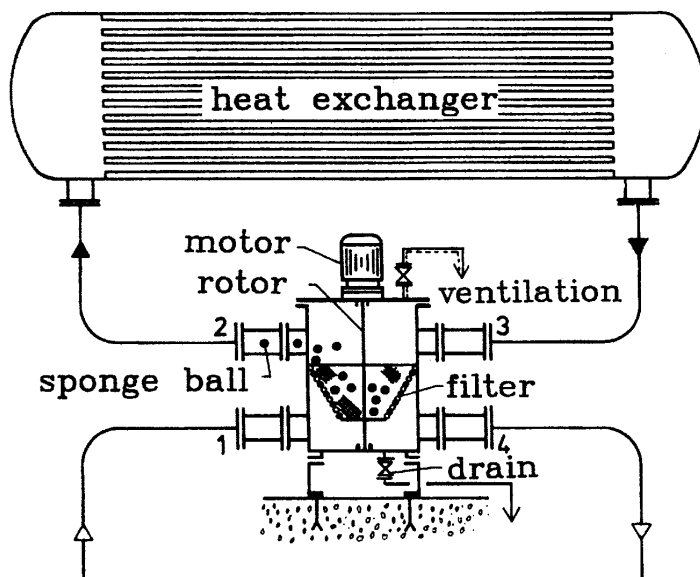


FIG. 64. Layout of Taprogge sponge ball cleaning system.

[111] states that about half of the on-line cleaning devices installed in U.S. utilities are out of commission at any one time.

5. Fluid-Bed Heat Exchangers

A fairly recent solution for heat transfer involving severely fouling liquids is the fluid-bed heat exchanger, which has been described by Klaren [112] and Kollbach *et al.* [113]. Small solid particles (glass, ceramic, metal) are fluidized inside parallel tubes by the upward flow of liquid. The solid particles regularly break through the viscous boundary layer, so that good heat transfer is achieved in spite of relatively low flow velocities [114]. More importantly, the solid particles have a slightly abrasive effect on the wall of the heat exchangers tubes, thus removing most deposits at an early stage. A schematic diagram of a fluid-bed heat exchanger is shown in Fig. 66.

Fluid-bed heat exchangers can also be operated as falling-film, rising-film, or circulating-bed heat exchangers. They have been installed in water treatment

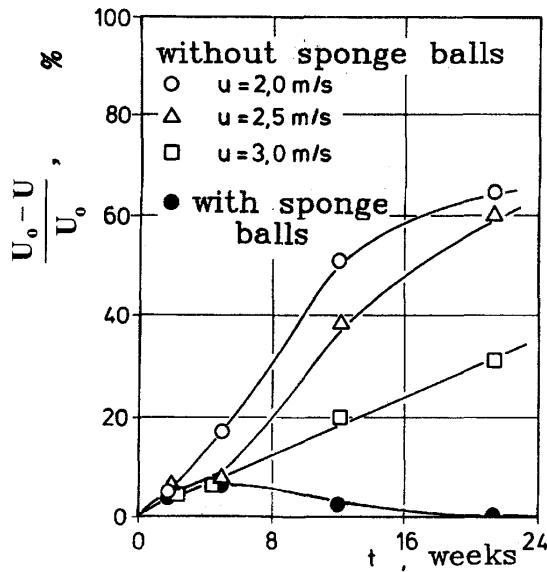


FIG. 65. Effect of sponge ball continuous cleaning [110].

plants, paper mills, food and dairy plants, geothermal plants, and various types of chemical plant. Figure 67 shows the heat transfer coefficient as a function of time for a fluidized-bed test heat exchanger on a side stream of an Alcoa of Australia Ltd. refinery [26].

To fluidize the 2.5-mm stainless steel particles, the flow velocity in the pipe was 0.32 m/s. It is important to realize the heat transfer coefficient for this low flow velocity is still about 10% higher than for purely convective heat transfer at 1.2 m/s, or 3.7 times higher than for a flow velocity of 0.3 m/s. Because of the hard and adherent nature of the silica deposit, scaling could not be eliminated completely, and the heat transfer coefficient decreased almost linearly with time. However, the fouling resistance after 12,000 min (8.3 days) is $0.2 \cdot 10^{-4} \text{ m}^2 \text{ K/W}$, which is only 15% of the fouling resistance observed without the fluidized bed over the same time span. The deposit was very smooth in those areas that were in contact with the fluidized particles and considerably rougher in unheated areas without particle contact. There was no deposition on the particles, even after 24 days of operation. Complete elimination of deposit formation has recently been achieved in confidential pilot measurements in a sulfuric acid concentration plant.

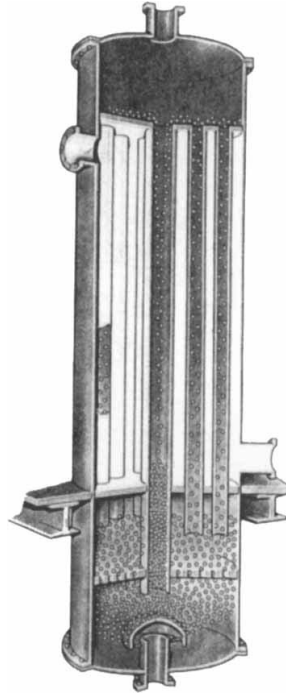


FIG. 66. Fluid-bed heat exchanger.

6. *Other Mitigation Techniques*

Wood pulp fibers are a cheap, renewable, biodegradable and nontoxic material. Recent investigations have shown that the addition of small amounts ($<0.01\%$) of these fibers to saturated CaSO_4 solutions can almost eliminate scale formation on heat transfer surfaces [115].

There are numerous devices on the market that claim to reduce fouling by magnetic, electric, radiation, or catalytic treatment. Parkinson and Price [116] and Donaldson and Grimes [117] report that the installation of magnets considerably reduced cooling-water fouling, whereas Hasson and Bramson [118] and Söhnle and Mullin [119] found no such effect. Industrial feedback on the installation of these devices covers the full spectrum from successful to no effect to made things worse. In the light of this evidence, it is difficult to come to any conclusion other

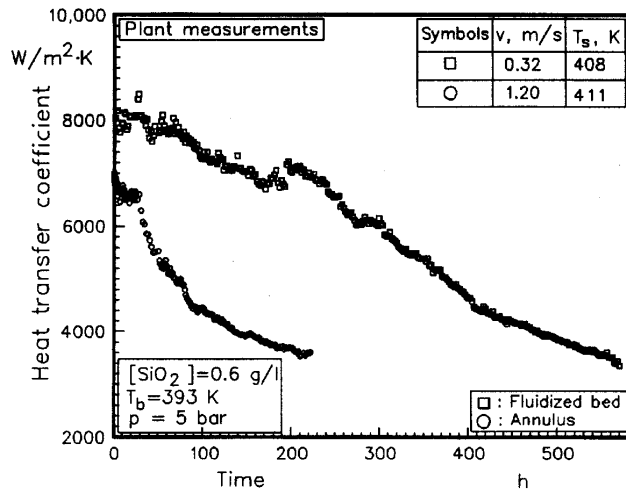


FIG. 67. Reduction of fouling from Bayer liquor in a fluidized bed [26].

than that the direct influence of a magnetic field on the process of crystallization is not very significant. Troup and Richardson [120] prepared a survey on the performance of various gadgets for cooling-water fouling mitigation and concluded that extensive experimental and field work is required to evaluate the efficiency of magnetic, electrical, and ultrasonic devices. Until this is done and any limitations that these apparatuses have are clearly established, their economic evaluation with respect to other available scale-prevention methods will not be realistically possible.

X. Conclusions

For (too) many years, cooling-water fouling has been considered to be an act of God—unavoidable and beyond human responsibility. Owing to the enormous costs associated with fouling and to the appearance of new processes, a variety of fouling mitigation techniques have been developed. Chemical methods have the lowest investment costs and can be installed in operating plants relatively easily. Sophisticated treatment programmes have been developed that, if carefully monitored can greatly extend the operation time of heat exchangers. However, long-term operating costs resulting from the need for a continuous supply of

chemicals and from environmental regulations may reduce this advantage. Mechanical fouling-mitigation techniques have high installation costs but relatively low operation costs. If fouling-mitigation techniques are implemented at the design stage, capital savings are possible, since no provisions have to be made for excess heat transfer surface area and excess weight. Prediction of the effect of operation conditions on fouling rates has improved, but is still far from being satisfactory and reliable. Research and development on fouling mitigation has developed in two ways:

1. Fundamental research using model fluids with constant flow and fouling parameters, to determine the mechanisms of deposition. Unfortunately, most results can hardly be translated into industrial applications because of experimental simplifications, such as small batch sizes, single-component fouling, and accelerated fouling conditions.
2. Applied research, mainly industrial development of additives and on-line/off-line cleaning strategies, to provide solutions to existing fouling problems. The results are highly empirical and cannot be extrapolated to other operating conditions and fluids.

Obviously, both approaches have serious limitations. Real progress in the mitigation of heat-exchanger fouling can be made only through close cooperation between industry and academia. This will provide relevant data that can be used to assess/curve-fit deposition models obtained under laboratory conditions. Fouling-mitigation strategies developed from these models or from laboratory experiments have to be verified by side-stream or plant measurements. While such projects will be time-consuming and relatively expensive, they can result in significant savings in capital and operating costs. They require, however, some commitment of the collaborating company in terms of staff time and access to equipment.

Heat exchangers are usually not stand-alone units. It may, therefore, be necessary to incorporate the non-steady-state behavior of the heat-exchanger performance into a plant model to determine optimum operating conditions and cleaning schedules.

Symbols

a	Constant defined in Eq. (24)	C_2	Constant in Eq. (24)
A	Constant defined in Eq. (10)	d	Tube diameter, m
A	Heat transfer surface area, m ²	d_h	Hydraulic diameter, m
b	Constant defined in Eq. (25)	D	Diffusivity, m ² /s
c	Constant defined in Eq. (26)	E	activation energy, J/mol
c	Concentration, g/L	f	Friction factor
C_1	Constant in Eq. (24)	I	Ionic strength of solution

k_r	Reaction rate constant, $\text{m}^4/\text{kg} \cdot \text{min}$	v_t	Thermophoretic velocity, m/s
K	Constant	W_1	Flow rate * specific heat of cooling water, W/K
K_1	First molar dissociation constant of carbonic acid, mol/m^3	z_i	Charge of component i
K_2	Second molar dissociation constant of carbonic acid, mol/m^3	α	Film heat transfer coefficient, $\text{W}/\text{m}^2 \cdot \text{K}$
K_{sp}	Molar solubility product for calcium carbonate, $(\text{mol}/\text{m}^3)^2$	β	Mass transfer coefficient, m/s
L	Length, m	γ	Activity coefficient
L.S.I.	Langelier Saturation Index	λ	Thermal conductivity, $\text{W}/\text{m} \cdot \text{K}$
m	Mass loading, kg/m^2	Λ	Electrical conductivity, μS
m_i	Molar concentration of component i	ν	Kinematic viscosity, m^2/s
$\dot{m}$	Mass flux, $\text{kg}/\text{m}^2 \cdot \text{s}$	Φ	Time constant, $1/\text{s}$
q	Heat flux, W/m^2	ρ	Density, kg/m^3
p	Pressure, Pa	τ_w	Wall shear stress, N/m^2
pK	$-\log(K)$		
pH_s	pH at saturation		
$\mathcal{R}$	Universal gas constant, $8.314 \text{ J}/\text{mol} \cdot \text{K}$	Subscripts	
R	Annulus outside diameter, m	a	Adhesion
R_A	Bulk reaction term in Eq. (28)	b	Bulk
Re	Reynolds number	c	Clean
R_f	Fouling resistance, $\text{m}^2 \cdot \text{K}/\text{W}$	d	Deposition
R_f^*	Asymptotic fouling resistance, $\text{m}^2 \cdot \text{K}/\text{W}$	dep	Deposit
R.S.I.	Ryznar Stability Index	eff	Effective
s	Deposit thickness, m	f	Fouled
Sc	Schmidt number	i	Inside
Sh	Sherwood number	i	Component i
S.I.	Saturation index	l	Liquid
t	Time, s	max	Maximum
t_c	Time constant, s	o	Outside
T	Temperature, K	r	Removal
T.A.	Total alkalinity	s	Heat transfer surface
T.D.S.	Total dissolved solids concentration, mg/l	scale	Due to scale formation
u	Flow velocity, m/s	t	Transport
U	Overall heat transfer coefficient, $\text{W}/\text{m}^2 \cdot \text{K}$	w	Wall
		*	Saturation
		0	Inside diameter

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