

FOULING: THE ULTIMATE CHALLENGE FOR HEAT EXCHANGER DESIGN

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ABSTRACT

The costs caused by the process related formation of deposits on heat transfer surfaces are in the order of 40,000 million dollar U.S. per year, for the total industrialized world! Despite increased attention during the past two decades, the allowance for fouling is still the most underestimated problem in the design of heat exchangers. Further refinement of heat exchanger design procedures will only be possible 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, 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. In this key-note paper, results of numerous experimental and theoretical investigations during the past 10 years are presented. These investigations have been performed with model or industrial fluids in laboratory or plant equipment. The state-of-the-art in fouling mitigation is reviewed and attention is drawn to areas where future academic and industrial research is required.

1. INTRODUCTION

The need for proper energy management has considerably increased the importance of heat exchanging equipment over the past 2-3 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, possible due to the improved knowledge of heat transfer fundamentals as well as the availability of larger and faster computers. Design calculations have shifted from overall heat transfer coefficients via 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 & tube, plate & frame, coiled bundles or plate-fin heat exchangers. The most widely known and accepted design software was developed by Heat Transfer And Fluid Flow Service (HTFS) in the U.K. and by Heat Transfer Research Incorporated (HTRI) in the U.S.

1.1 Description of Problem

The usefulness of the above design procedures is considerably reduced by the fact that they are only valid for clean fluids. In most practical processes, however, the heat exchanging fluids contain certain amounts of dissolved or suspended material or provide conditions favourable 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 1, their resistance to heat transfer may well be within the order of magnitude of that of the clean fluids. As a result, reduced

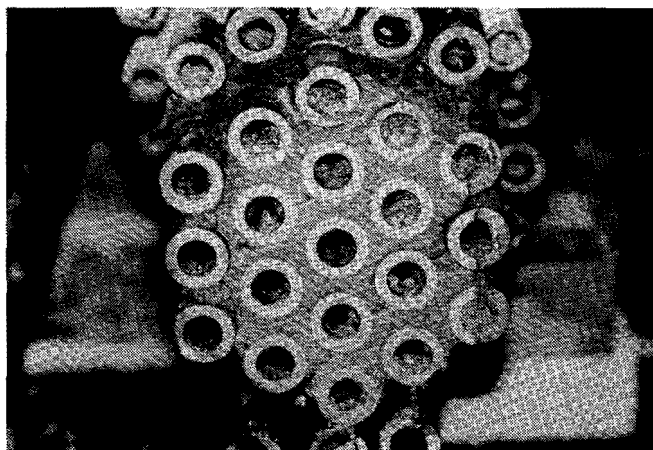


Fig 1 Deposition on the shell-side of an ammonia reactor effluent cooler

Table 1 Thermal Conductivity of Various Deposits

sodium aluminium silicate	0.2 - 0.4 W/mK
milk components	0.5 - 0.7 W/mK
hematite (boiler deposit)	0.6 W/mK
biofilm	0.7 W/mK
calcium sulphate (boiler)	0.8 - 2.2 W/mK
calcite (boiler deposit)	0.9 W/mK
serpentine (boiler deposit)	1.0 W/mK
gypsum (boiler deposit)	1.3 W/mK
calcium sulphate	2.3 W/mK
magnesium phosphate	2.3 W/mK
calcium phosphate	2.6 W/mK
calcium carbonate	2.9 W/mK
magnetite iron oxide	2.9 W/mK

heat exchanger performance is common and regular cleaning may be required. This is shown in Figure 2 for the live steam heater in a bauxite refinery, where sodium aluminium silicate deposits necessitate acid washes in intervals between 5 and 15 days.

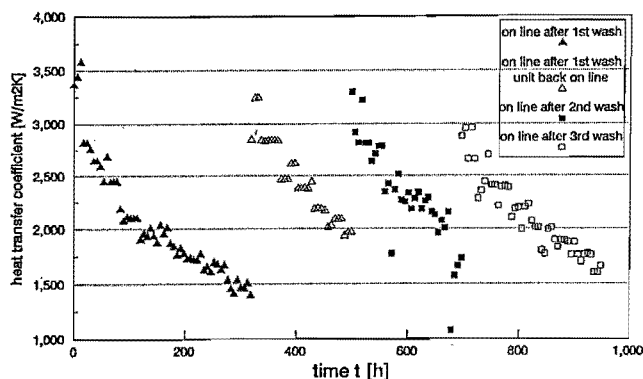


Fig. 2 Operation cycles of the live steam heater of an Alcoa of Australia Ltd. bauxite refinery

A mail survey covering 3,000 heat exchangers from about 1,100 New Zealand companies⁽¹⁾ showed that more than 90% of heat exchangers had some fouling problems, see Fig. 3.

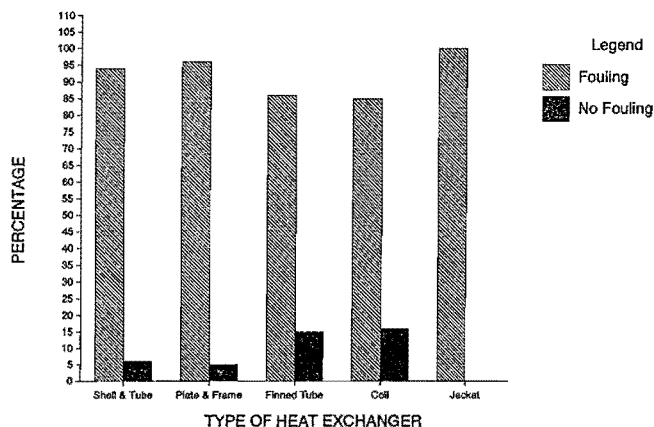


Fig. 3 Fouling problems in various heat exchanger types

1.2 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

$$\frac{1}{U} = \left(\frac{1}{\alpha_1} + R_{f1} \right) \frac{d_2}{d_1} + \frac{d_2 \ln(d_2/d_1)}{2 \lambda_w} + \frac{1}{\alpha_2} + R_{f2} \quad (1)$$

The almost exclusive source of fouling resistances are the approximately 100 values suggested by the Tubular Exchanger Manufacturers Association (TEMA)⁽²⁾, which have not been significantly updated since their first publication in 1947. Most major companies have a number of 'proprietary fouling resistances' but closer inspection usually reveals that these values are all but identical to the TEMA resistances. Since the TEMA values are only for a very limited number of fluids (mainly water and hydrocarbon processing streams) and do not have any correction for the effects of operating parameters (dirt concentration, surface temperature, flow velocity), selection of the appropriate fouling resistance is more or less 'guess-work' or 'gut-feeling'. In many cases, the user of heat exchangers over-specifies 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 required heat transfer surface area is inversely proportional to the overall heat transfer coefficient. Table 2 shows the excess heat transfer surface required for several heat exchanger types, if a typical TEMA fouling resistance of 0.18 m²K/kW is used for each of the two heat exchanging fluids.

Table 2 Excess Surface Area for Various Heat Exchanger Application, $R_f = 0.36 \text{ m}^2\text{K/kW}$

Application	Clean Overall Coefficient	Excess Area
gas/gas shell & tube heat exchanger	50 W/m²K	1.8%
liquid/gas shell & tube heat exchanger	150 W/m²K	5.4%
liquid/liquid shell & tube heat exchanger	1000 W/m²K	36%
liquid/liquid plate & frame heat exchanger	3000 W/m²K	108%
water-cooled shell & tube steam condenser	4500 W/m²K	162%

Obviously, the percentage 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. Most manufacturers of plate and frame heat exchangers recommend that the excess surface should not exceed 25% of the heat transfer surface area calculated for the clean duty. However, for all heat exchanger types, the percentage over-design should routinely be calculated to re-evaluate the usefulness of the design or of the selected fouling resistance.

For 5 major chemical plants built between 1970 and 1990 with more than 100 heat exchangers per plant, the

following average excess surface was found for all heat exchangers.

Table 3 Excess Surface Area Installed in Several Chemical Plants

Plant	Required heat transfer surface	Add. heat transfer surface for fouling	Percentage
A	14,181 m ²	5,533 m ²	39%
B	10,159 m ²	2,856 m ²	28%
C	31,045 m ²	9,413 m ²	30%
D	21,831 m ²	11,024 m ²	50%
E	21,345 m ²	9,349 m ²	44%

This confirms the findings of Garrett-Price et al.⁽³⁾ that heat exchangers are in average over-designed by 35% to allow for fouling.

1.3 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 due to fouling and to attribute these costs to the various aspects of heat exchanger design and operation. However, reliable knowledge of fouling economics is desirable to evaluate the cost efficiency of various mitigation strategies. The total fouling related costs consist of

Capital Expenditure. This includes excess surface area, costs for stronger foundations, provisions for extra space, increased transport and installation costs. There are additional capital costs for antifouling equipment, such as the installation of on-line cleaning devices, pre-treatment plants and cleaning-in-place equipment.

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.

Maintenance Costs. Maintenance costs are costs for removing fouling deposits and costs for chemicals or other operating costs of antifouling devices. According to Pritchard⁽⁴⁾ and Thackery⁽⁵⁾, about 15% of the maintenance costs of process plant can be attributed to heat exchangers and boilers and of that 50% is probably due to fouling.

Costs Due To Production Loss. Because of planned and unplanned plant shut-downs due to fouling in heat exchangers, large production losses are possible. These costs 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 in refs 1 and 3. Surprisingly good agreement was found in refs 1, 3 and 5 for the total costs due to fouling. While refs 3 and 5 found that total fouling costs for highly industrialized countries such as the U.S. and the U.K. are about 0.25% of the countries' Gross National Product (GNP), the fouling costs for New Zealand, which is less industrialized, are 0.15%

of the New Zealand GNP⁽¹⁾. Using these percentages, Table 4 lists total fouling related costs for various countries, based on 1984.

Table 4 Total Fouling Costs for Several Countries

Country	Fouling costs in million \$US	GNP 1984 billion \$US	Fouling as % of GNP
UK (1978)	700-930	285	0.20-0.33
USA (1982)	3,860-7,000 8,000-10,000	3,634	0.12-0.22 0.28-0.35
NZ	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

2. MECHANISMS OF DEPOSITION

Because of the great variety of fouling mechanisms it is useful to divide fouling according to the key physical/chemical processes into five major categories⁽⁶⁾, namely:

- Crystallization Fouling
- Particulate Fouling
- Chemical Reaction Fouling
- Corrosion Fouling
- Biological Fouling

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

The above fouling mechanisms generally occur in five consecutive steps:

Initiation period or delay period. When the new or cleaned heat exchanger has been taken into operation, the initially high heat transfer coefficients may remain unchanged for a certain time. During this time, nuclei for crystallization are formed or nutrients for biological growth are deposited. This delay period may last anytime from few seconds to several days. No delay period occurs for particulate fouling⁽⁷⁾. For crystallization fouling and for chemical reaction fouling, the initiation period decreases with increasing surface temperature, as supersaturation and/or reaction rate increase⁽⁶⁾. Generally, it is reported that the delay time before deposition starts decreases with increasing roughness of the heat transfer surface.

Mass transport. To form a deposit at the heat transfer surface it is necessary that at least one key component is

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.

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 (for example CaCO_3). If the total matter available for deposition sticks to the heat transfer surface, the concentration at the wall, c_w , becomes zero. The deposition is then controlled by the diffusion to the heat transfer surface. If c_w is only slightly less than c_b , fouling is reaction or sticking controlled.

Removal or auto-retardation. Depending on the strength of the deposit, erosion occurs immediately after the first deposit has been laid down. Furthermore, several mechanisms exist which cause auto-retardation 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 heat transfer surface and fluid.

Aging. Every deposit is subjected to aging. Aging may increase the strength of the deposit by polymerization, recrystallization, de-hydration etc. Biological deposits get 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 modelling attempts.

Depending on the process parameters and the dominant fouling mechanism, the fouling rate can be either constant or decreasing with time, see Fig. 4. For hard, adherent deposits such as silicates or some polymerization products, steps iv) and v) 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.

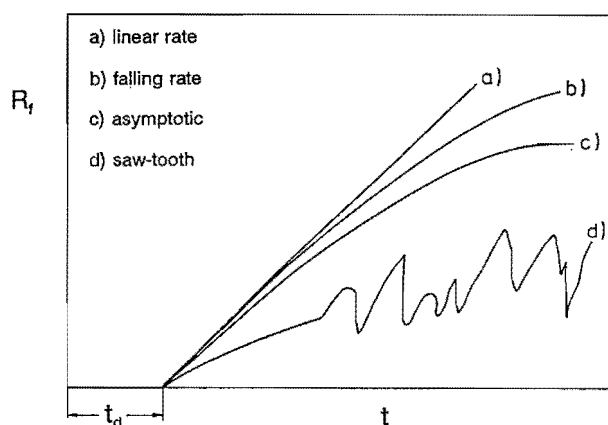


Fig. 4 Possible fouling resistance vs. time curves

3. EFFECT OF PROCESS PARAMETERS

If the effect of process parameters such as flow

velocity, bulk and surface temperature, dirt concentration etc. on the deposition rate is known, selection, design and operation of heat transfer equipment may be adjusted to minimize fouling.

3.1 Approximate Approach

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

- Fouling usually increases linearly with increasing foulant concentration in the fluid bulk. While several ideal reactions may be of different order, this assumption holds for most industrial cases.
- The fouling resistance nearly always decreases with increasing wall shear stress due to increased removal forces. As an average, it was found

$$R_f \sim u^{-1.5} \quad (2)$$

- For all fouling mechanisms, the fouling resistance increases exponentially with increasing surface temperature according to an Arrhenius term.

$$R_f \sim e^{\frac{-B}{RT}} \quad (3)$$

For biological fouling, maximum fouling is observed for temperatures around 35°C .

- Fouling was found to increase with increasing roughness of the heat transfer surface.

To date, not even these simple results are used in the design of heat exchangers, even though they could significantly improve some heat exchanger optimization procedures. This is demonstrated in Figure 5, 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 in ref 8 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 m/s to about 1.3 m/s and the total annual costs are reduced by 10% despite the higher friction losses.

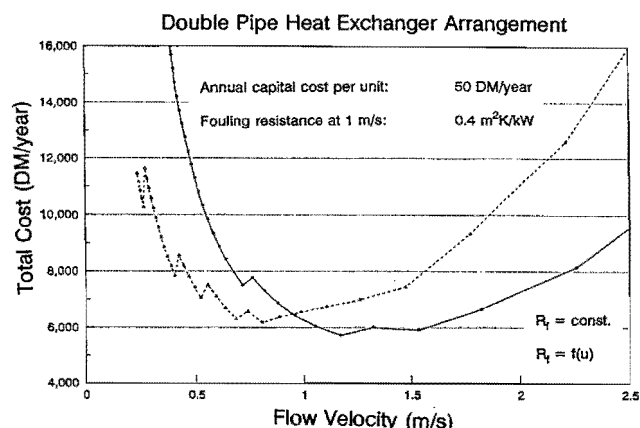


Fig. 5 Total annual costs of a double pipe heat exchanger arrangement as a function of the flow velocity

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 or heating medium inlet temperature are specified. If this is the case, the heat exchanger is usually controlled via the flow rate of the cooling or heating medium. To reduce the heat duty, the service fluid flow velocity must be reduced. Using the above, approximate relationships, Figure 6 shows that this procedure may cause a considerable increase of fouling as compared to fouling under design operating conditions. Point "A" refers to the design values of flow velocity and heat transfer surface temperature. As the heat exchanger is initially over-designed, the cooling water flow velocity is throttled which also causes an increase of the heat transfer surface temperature, see 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, by specifying high fouling resistances, fouling may become a self-fulfilling prophecy. If part of the cooling water is re-circulated, the flow velocity and the cooling water inlet temperature can be increased to meet the required heat duty. The anticipated fouling ("C") will be similar to the design value ("A"), but a price is to be paid to provide the higher flow velocity.

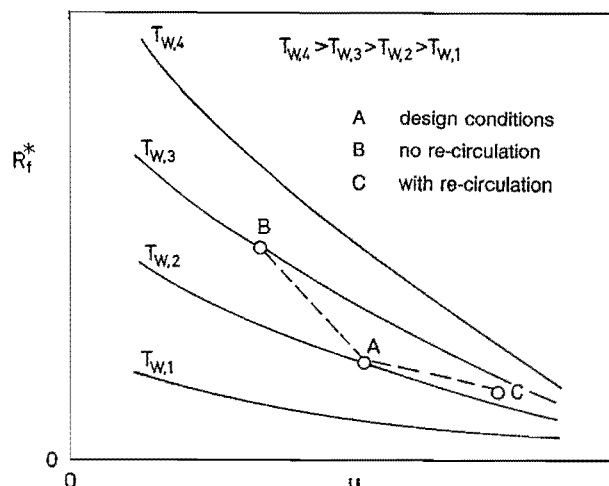


Fig. 6 Fouling resistance as a function of low velocity and surface temperature during start-up of a new or cleaned heat exchanger

3.2 Modelling Approach

Since the actual fouling rates will be different for each situation, a more sophisticated approach is desirable which considers the effects of both, operating conditions and fluid composition.

Most fouling models are based on the assumption 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⁽⁹⁾.

$$\frac{dm}{dt} = \dot{m}_d - \dot{m}_r = \lambda_d \rho_d \frac{dR_f}{dt} \quad (4)$$

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 modelled as a diffusion process

$$\dot{m}_t = \beta (c_b - c_s) \quad (5)$$

where 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 forces or thermophoretic forces.

The adhesion step can be described by

$$\dot{m}_a = k_r (c_s - c^*)^n \quad (6)$$

with

$$k_r = A e^{\frac{-E}{RT}} \quad (7)$$

If the adhesion rate is considerably higher than the diffusion rate, the deposition is transport-controlled, $c_w = c^*$ and $\dot{m}_d = \dot{m}_t$. If the adhesion rate is considerably slower than the transport rate, $c_w = c_b$ and $\dot{m}_d = \dot{m}_a$. If both are of the same order of magnitude, c_w is eliminated by equating eqs. (5) and (6).

In most models, the removal flux is proportional to the fluid shear stress and to the thickness of the deposit:

$$\dot{m}_r = K \tau_w s \quad (8)$$

Sometimes, the constant K is assumed to be proportional to a 'deposit strength', which is even more elusive to measure.

Integration of eq. (4) yields

$$R_f(t) = R_f^* (1 - e^{-t/t_c}) \quad (9)$$

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

In the following, expressions based on the above fundamental approach are given for the fouling resistances of several ideal and industrial fluids.

CaCO₃ scaling. CaCO₃ deposits are common in heated and unheated water pipelines, and are caused by super-saturation owing to changes in temperature, pH and pressure. The following mechanisms are responsible:

- + transport
- + reaction at heat transfer surface
- removal (moderate flow rates)

From the combination of transport and reaction rate equations, Hasson's Ionic Diffusion model⁽¹⁰⁾ can be derived

$$\dot{m}_d = \frac{\beta [Ca^{++}] ((1 + 4ac/b^2)^{0.5} - 1)b}{2a} \quad (10)$$

with

$$a = 1 - \frac{4K_2 k_r [Ca^{++}]}{\beta K_1} \quad (11)$$

$$b = \frac{[\text{CO}_2]}{[\text{Ca}^{++}]} + \frac{4K_2k_t[\text{HCO}_3^-]}{\beta K_1} + \frac{K_{sp}k_t}{\beta[\text{Ca}^{++}]} \quad (12)$$

$$c = \frac{K_2k_t[\text{HCO}_3^-]^2}{K_1\beta[\text{Ca}^{++}]} - \frac{K_{sp}k_t[\text{CO}_2]}{\beta[\text{Ca}^{++}]} \quad (13)$$

Equations for the equilibrium constant, the reaction rate constant and the dissociation constants of aqueous CaCO_3 solutions are given, for example, in ref 11.

Eq. (10) has been used in conjunction with the Schlünder-Gaddis⁽¹²⁾ cell model to determine temperature distribution in and effectiveness of shell and tube heat exchangers, see Fig. 7. Configurations G1-G4 refer to the position of inlet and outlet nozzles of cold and hot stream. Without fouling ($t = 0$), configurations G2 and G3 are advantageous. However, if CaCO_3 scaling may be expected, only option G3 can be recommended.

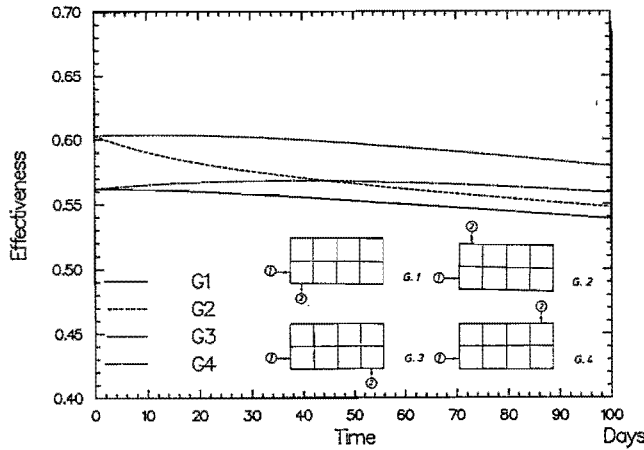


Fig. 7 Effectiveness of various shell and tube heat exchanger configurations as a function of time

Particulate fouling. Particulate fouling is the accumulation of suspended solids (sand, silt, clay, iron oxide) onto the heat transfer surface. Generally, particulate fouling starts without initiation period and shows asymptotic behaviour. The following mechanisms have been identified and been included in a correlation for the asymptotic fouling resistances.

- + transport of particles (diffusion, inertia, thermophoresis)
- + adhesion (Van-der-Waals forces, electrical double layer)
- + removal of particles

$$R_f^* = \frac{K_3(\beta - v_t/2)c_b e^{-\frac{E}{RT}}}{\lambda_d \rho_d (K_4 e^{-\frac{E}{RT}} + \beta + v_t/2) v_1 \xi u^2 d_p} \quad (14)$$

Fig. 8 compares the measured effect of particles size on the asymptotic fouling resistances with the predictions of the original Kern-Seaton model⁽⁹⁾, the Watkinson-Epstein model⁽¹³⁾ and eq. (14). The measurements have been performed with suspensions of Al_2O_3 particles in a solvent mixture X-2 with properties similar to heptane.

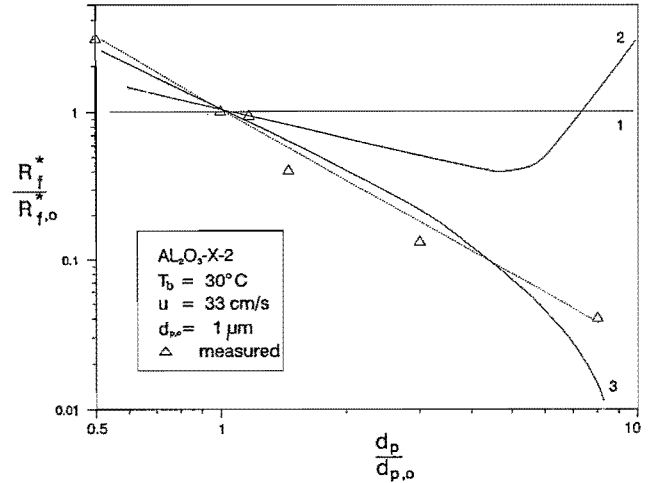


Fig. 8 Effect of particle size on the asymptotic fouling resistance 1 = ref 9, 2 = ref 13, 3 = eq (14)

While adding excess heat transfer surface in the design of shell and tube heat exchangers does not necessarily reduce the flow velocity, this is usually the case if additional plates are added to a plate and frame heat exchanger. Eq. (14) was used to design a plate heat exchanger built from plates with a very small surface area. Fig. 9 shows the heat duty as a function of the heat transfer area which is proportional to the number of parallel plates. If a constant fouling resistances is used, the duty increases with increasing heat exchangers size, while an optimum design exists if eq. (14) is used for the fouling resistance. The latter results is in accordance with practical experience.

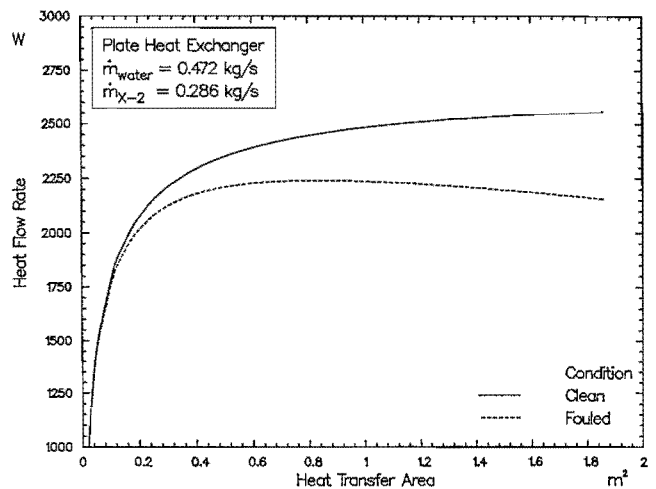


Fig. 9 Effect of plate number on the heat duty of a plate and frame heat exchangers

Fouling under boiling conditions. Heat transfer to a boiling liquid occurs by three mechanisms, namely by natural convection to the liquid, by conduction through and evaporation of a liquid microlayer between bubble and heat transfer surface and by heat transfer from the superheated liquid boundary layer to the vapour bubble. From these mechanisms, only the first two may cause the formation of deposits on the heat transfer surface. The micro-layer between heat transfer surface and bubble is supplied with liquid from the bulk liquid. Due to evaporation of the carrier liquid, the concentration of components with lower vapour pressure can increase in the micro-layer by up to four orders of magnitude. If this concentration effect causes supersaturation, rings of deposit are formed around the nucleation sites. In this case, the growth rate of deposit is strongly depending on the number of active nucleation sites and, hence, on the heat flux. Fig. 10 shows the effect of heat flux on the relative drop of heat transfer coefficient during boiling of a saturated CaSO_4 solution.

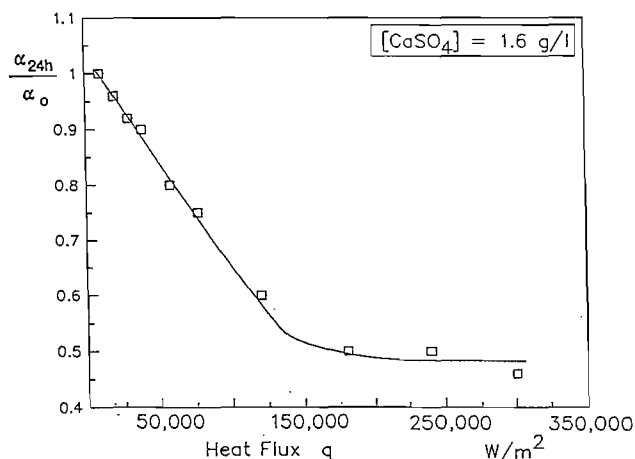


Fig. 10 Effect of heat flux on the relative drop of heat transfer coefficient during boiling

Fouling during boiling in evaporator tubes is strongly depending on the flow pattern. After the transition to droplet flow, fouling increases considerably due to the total evaporation of all droplets contacting the heat transfer surface⁽¹⁴⁾. The presence of a second phase during boiling significantly increases the level of turbulence, even during pool boiling. Fig. 11 shows that if three heaters are aligned vertically, the CaSO_4 deposition decreases from the bottom to the top heater, because of the shear forces generated by the ascending bubbles. It is, therefore, impossible to extrapolate from single tube data to tube bundle applications.

Fouling from Kraft black liquor. The Kraft pulping process is extensively used in the pulp and paper industry. Components in the wood such as lignin, starches, sugars, resins and fatty acids are leached from the wood chips by a solution consisting of a dissolved mixture of sodium hydroxide, sodium carbonate and sodium sulphide, leaving a suspension of wood pulp fibres in the spent pulping chemicals, known as black liquor. To make the process economically viable, the weak black liquor (15% solids) is concentrated in multiple effect evaporators up to 70% solids and incinerated in the recovery boiler from which the expensive process chemicals are recovered as the smelt.

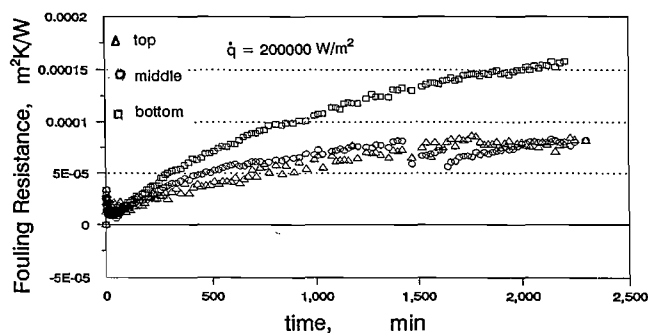


Fig. 11 CaSO_4 deposition on a boiler tube arrangement

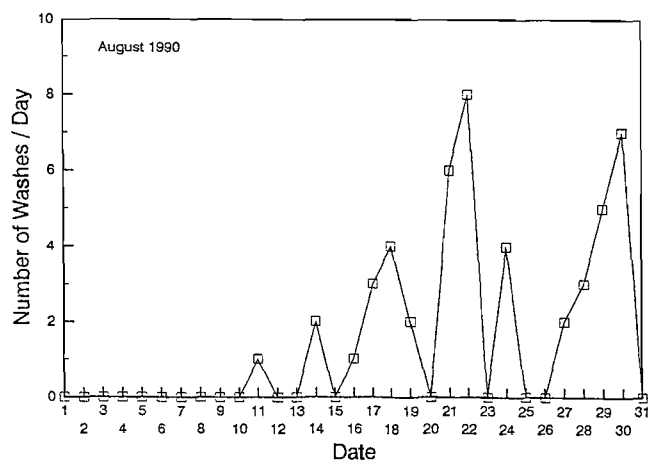


Fig. 12 Number of daily black liquor washes of the #5 evaporator high concentration effect in the NZFP Kinleith mill

However, significant fouling problems occur in the high concentration effects requiring washing with dilute liquor as often as every 2-3 hours, see Fig. 12. To optimize the performance of black liquor evaporators, extensive measurements have been performed with liquor from the NZ Forest Products Kinleith mill, which resulted in the following correlation:

- transport (supersaturated)
- + reaction (different for convective and nucleate boiling conditions)
- removal (insignificant for $u < 1 \text{ m/s}$)

$$\frac{\partial R_f}{\partial t} = \frac{k_r k_c}{\rho_d \lambda_d} (T_w - T_b)^n \quad (15)$$

$$k_r = K e^{\frac{-E}{RT}} \quad (16)$$

$$\frac{k_c K}{\rho_d \lambda_d} = 2.4 \cdot 10^{-3} \frac{\text{m}^2}{\text{kW m}^2} \quad (17)$$

$$n = 1 \quad (18)$$

$$E = 3,431.0 \text{ J/mole} \quad (19)$$

$$n = 1 \quad (20)$$

This equation was used to find an optimum operation time between washings, in conjunction with a solution by Epstein⁽¹⁵⁾.

$$t_{\text{evap}}^{\text{opt}} = t_{\text{clean}} + \sqrt{\frac{2t_{\text{clean}}}{\alpha \frac{\partial R_f}{\partial t}(0)}} \quad (21)$$

For a typical washing time of $t_{\text{clean}} = 120$ minutes and a heat transfer coefficient $\alpha = 2000 \text{ W/m}^2\text{K}$, the optimum evaporation time was calculated as 220 minutes, which is very close to the actual operating value found by trial and error.

Fouling from Bayer liquor. To extract the alumina from bauxite in the low temperature Bayer process the bauxite is digested in a strong caustic soda solution between 150°C and 180°C . The dissolved Al_2O_3 is then precipitated by cooling the liquor. The cool, dilute and alumina-depleted spent solution is usually supersaturated with respect to silica. When the spent liquor from the precipitation building is re-heated and concentrated to the digestion process conditions, the silica forms hard and tenacious sodium aluminium silicate deposits on the heat exchangers walls. As shown in Fig. 2, cleaning frequencies as high as every 5-10 days are required at the hot end of the process. The deposition process can be modelled by

- + transport (only for $[\text{SiO}_2] > 1.6 \text{ g/l}$)
- + reaction
- + removal (only for $[\text{SiO}_2] > 1.6 \text{ g/l}$)

For typical process liquors with $[\text{SiO}_2] \approx 0.7 \text{ g/l}$, the precipitation of scale on heat transfer surfaces is controlled by a second order reaction at the heat transfer surface. For constant heat flux ($\approx$ constant surface temperature), a homogeneous deposit and a moderate deposit thickness:

$$R_f(t) = \frac{k_r(c_b - c^*)^2}{\rho_d \lambda_d} t \quad (22)$$

The kinetic constant k_r is a function of temperature and was obtained by curve-fitting the experimental data

$$k_r = \exp\left(20.37288 - \frac{12387.622}{T}\right) \quad (23)$$

Obviously, scale formation increases with increasing heat transfer surface temperature. The highest temperatures in the Bayer process occur in the live steam heat exchanger, where steam from the boiler house is used to heat the liquor up to a temperature of about 155°C . A numerical model has been written for local heat transfer in the live steam heat exchanger. If this model is combined with eq. (22), the local deposition rates can be determined. Fig. 13 shows the effect of steam temperature on the drop in bulk silica concentration

between liquor inlet and liquor outlet in a shell and tube heat exchanger with 4 tube-side passes. A considerable reduction of the de-silication can be reached by operating at lower steam temperatures.

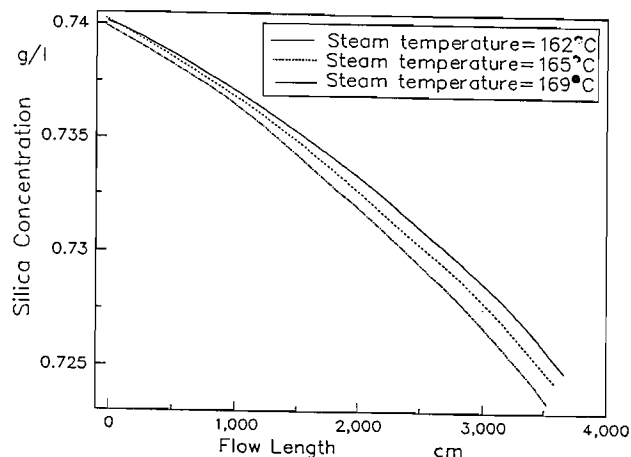


Fig. 13 Effect of steam temperature on silica concentration profile

4. MITIGATION OF HEAT EXCHANGER FOULING

Even with the best thermal and hydraulic design, formation of deposits will occur if the fluid becomes supersaturated or if particulate matter is present. Since fouling does not only reduce heat transfer, but also increases friction losses and promotes corrosion, attempts have to be made to either reduce the potential of a fluid to form deposits or to remove the deposits during operation.

4.1 Chemical Mitigation

Fouling generally increases with increasing foulant concentration in the fluid bulk. Particulate matter can be removed by filtration, flocculation and sedimentation. Reduction of attractive forces by dispersants and surfactants is also possible.

Scaling species may be removed by ion exchange or by chemical treatment. Growth and/or nucleation of crystals can be inhibited by the addition of scale inhibitors. Chelating agents (for example EDTA) complex strongly with the scaling cations and hence inhibit their deposition at the heat transfer surface by reducing the driving concentration difference. The effect of EDTA addition on CaSO_4 deposition during pool boiling is shown in Fig. 14. Processes which are based on physical rather than on chemical reactions are those that stabilize supersaturated solutions by adsorption at the crystal nuclei (for example polyphosphates) or that modify or weaken the crystalline structure (for example polycarboxylic acid).

Lists of additives to reduce crystallization from hard waters have been compiled by Harris and Marshall⁽¹⁶⁾ and by Krisher⁽¹⁷⁾. Polymerization at the heat transfer surface can often be reduced by the addition of oxygen scavengers (for example 3-butyl-catechol) to the fluid. The most common method to reduce biological fouling is the chlorination of water.

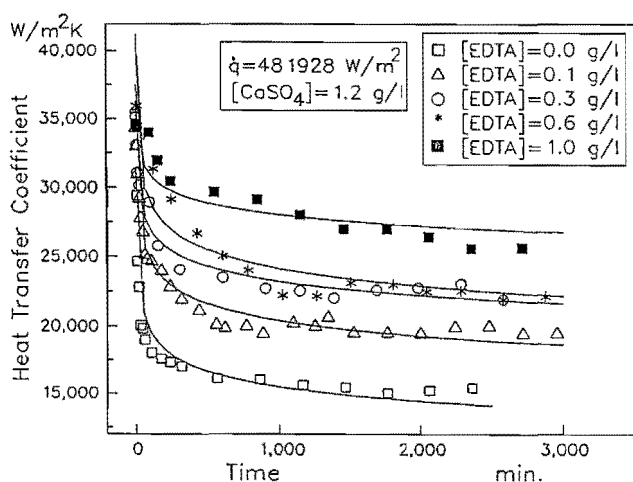


Fig. 14 Effect of EDTA concentration on deposition during boiling of a saturated CaSO_4 solution

Selection and dosage of chemical additives is highly proprietary and a number of national and international companies have specialized in this field. If possible, trial runs in pilot units are recommended, because the effect of additives on complex process fluids can not be reliably predicted. In ref 18 it has been reported that from two additives recommended for Kraft black liquor, one required a tenfold dosage to be effective, whereas the other actually increased the fouling rate.

4.2 Mechanical Mitigation

While avoiding the continuous addition of chemicals, mechanical mitigation methods involve considerable investment costs. A number of mitigation techniques have been developed which generally are based on one of the following mechanisms:

- Short-time overheating of the heat transfer surfaces. The different thermal expansivity of tubes and tube deposits may cause cracking of the deposit.
- Mechanical vibration of heat transfer surfaces.
- Acoustical vibration of heat transfer surfaces.
- Increased shear stress at fluid/deposit interface.
- Reduced stickiness of the heat transfer surface

For gas-side heat transfer soot blowers are the most popular method. Some other techniques such as scrapers, rappers and chains work well in special applications but are not as readily available. The two common types of soot blowers are the jet soot blowers and the sonic soot blowers. The jet type of soot blower operates by emitting pulses of steam or air at programmed intervals directed at the tubes and/or down tube lanes to dislodge the deposits and re-entrain them in the gas stream. These soot blowers work best if used frequently, thus avoiding the build-up of material. When build-up occurs, it insulates the surface from the coolant, allowing a temperature rise that can produce a glassy deposit. Glassy deposits are much harder to dislodge.

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

Increasing the flow velocity. An effective method to reduce fouling is to increase the wall shear stress via the increase of flow velocity. This procedure is only effective as long as the design velocity is maintained during operation, see Section 3. Zones with low flow velocity should be avoided as much as possible. Temporarily increasing the flow velocity or 'air rumbling' are effective and inexpensive mitigation techniques, if deposits are not too hard and adherent. The influence of flow velocity and water quality on cooling water fouling is shown in Figure 15.

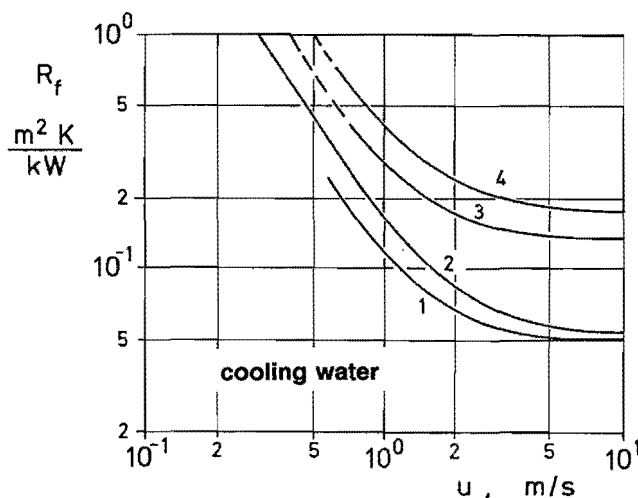


Fig. 15 Effect of flow velocity and water quality on the fouling resistance⁽¹⁹⁾

Reversal of flow direction. Regular reversal of the flow direction in conjunction with a short-time increase of the flow velocity is sometimes used as a method to remove weak deposits. Fig. 16 shows that this procedure reduces the fouling resistance, but only for a short period of time. A much better performance could be achieved by operating at a higher flow velocity.

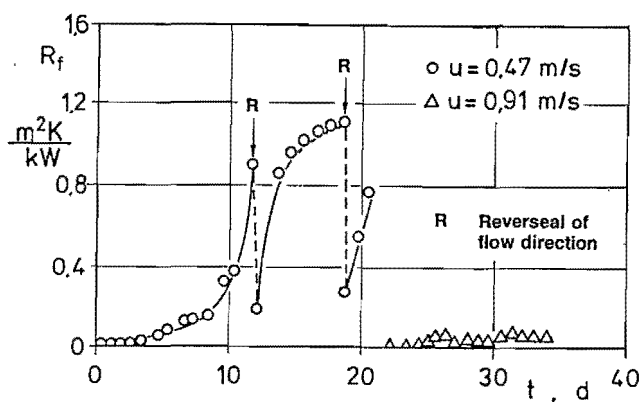


Fig. 16 Effect of flow reversal on the formation of deposits⁽²⁰⁾

Heat transfer surface roughness. Surface roughness provides ideal nucleation sites for the growth of deposit and increases the stickiness of the fouling layers. The use of electro-polished stainless steel evaporator tubes in pulp and paper industry, chemical industry and food processing industry consistently allowed for longer operation periods between cleaning. If the heat transfer resistance of the tube material is not a major problem, PTFE or polypropylene tubing, or glass lined tubing may be used. This is demonstrated in Fig. 17 for deposition from Kraft black liquor. The beneficial aspects of smooth heat transfer surfaces is lost once the original surface is covered with a continuous dirt coating. If heat exchangers are cleaned it is important to ensure that maximum surface smoothness is achieved.

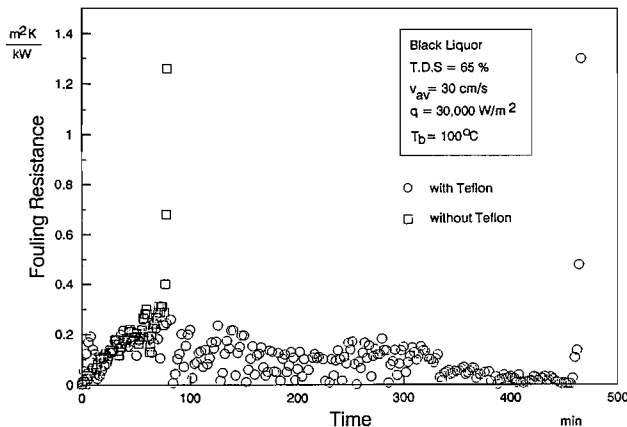


Fig. 17 Fouling from Kraft black liquor on a stainless steel surface and on a PTFE coated surface

Turbulence promoters. Tube corrugations and tube inserts such as those shown in Fig. 18 can increase the heat transfer coefficient by a factor of 2 to 15 by reducing the average thermal boundary layer thickness. As deposition rates for most fouling mechanisms are inversely dependent on fluid wall shear stress and heat transfer surface temperature, reduction of the viscous and thermal sub-layer thickness may also considerably reduce fouling. This will, however, only be effective for crystallization or chemical reaction fouling. If particulate or fibrous material already exists in the fluid, the presence of turbulence promoters may actually increase the deposition rate. 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.

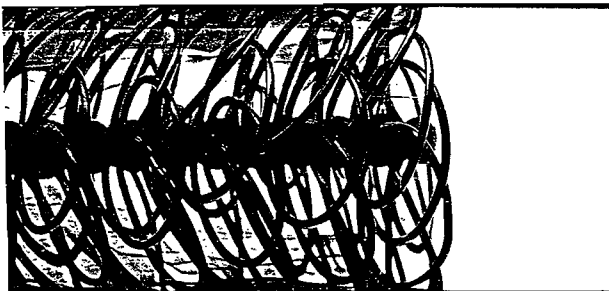


Fig. 18 Cal-Gavin wire matrix insert⁽²¹⁾

Instead of artificially increasing the turbulence level in pipe flow, heat exchangers with an inherently higher turbulence level, such as plate and frame heat exchangers can be used. It has been reported⁽²²⁾ that the fouling rates in plate and frame heat exchangers are significantly lower than those in shell and tube heat exchangers. For a 65% solids Kraft black liquor which caused severe fouling in a double pipe heat exchanger after only 2 hours, no drop in heat transfer was observed in a plate heat exchanger over a period of 2 days⁽²³⁾.

Continuous transport of cleaning devices through tubes. These methods need 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 applied to an initially clean heat exchanger.

A number of companies (MAN, Water Services of America, KALVO, ATCS) have developed a continuous tube cleaning system using small nylon or stainless steel brushes which are inserted into each tube, see Fig. 18. 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 hours. Life expectancy of the brushes is about 5 years. Typical applications for the nylon brush system are cooling water duties in condensers or chillers. The time for amortization is claimed to be between 8 and 16 months.

Although there are many examples for the successful application of the brush tube cleaning system, it was found that it may fail for hard deposits. Better results were obtained with a system where sponge balls with a rough surface are circulated through the heat exchanger, see Fig. 19.

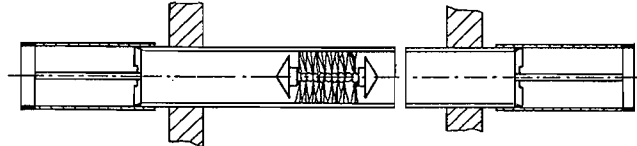


Fig. 19 Continuous cleaning with wire brush system

The diameter of the sponge balls is slightly larger than the inside diameter of the tubes and the system is designed such that each tube sees a sponge ball every 5 to 10 minutes. Since 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 turbo-generators above 100 MW output are equipped with the TAPROGGE sponge ball system, see Fig. 20. 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 6-10. The application of sponge ball systems is limited to temperatures below 120°C. According to the manufacturers, application of sponge ball systems almost eliminates fouling and has an amortisation time of about one year.

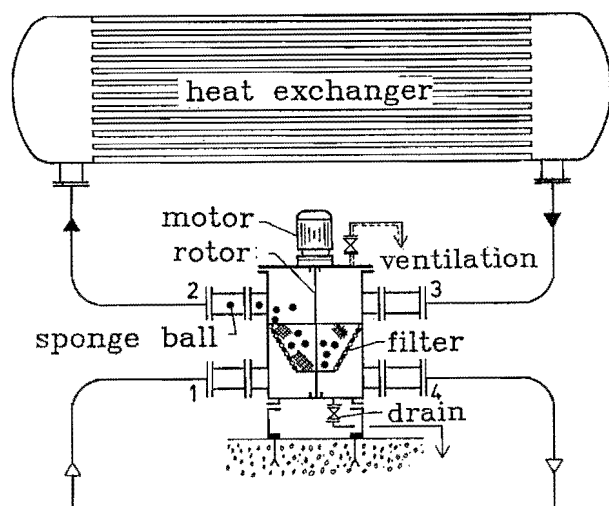


Fig. 20 Lay-out of TAPPROGGE sponge ball cleaning system

On-line cleaning systems may require considerable maintenance. They are not effective against stones, clamshells etc. and need up-stream devices to remove debris and macroscopic organic matter from the incoming water. If not maintained properly, excessive ball losses may occur. Strauss⁽²⁴⁾ states that about half of the on-line cleaning devices installed in U.S. utilities are out of commission at any time.

Fluid bed heat exchangers. A fairly recent solution for heat transfer involving severely fouling liquids is the fluid bed heat exchanger^(25,26). 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⁽²⁷⁾. More importantly, the solid particles have a slightly abrasive effect on the wall of the heat exchanger tubes, thus removing most deposits at an early stage.

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 plants, paper mills, food and dairy plants, geothermal plants and in various chemical plants. Fig. 21 shows the heat transfer coefficient as a function of time for measurements with a fluidized bed test heat exchanger on a side-stream of an Alcoa of Australia Ltd. refinery. 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 that 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 minutes (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 which 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.

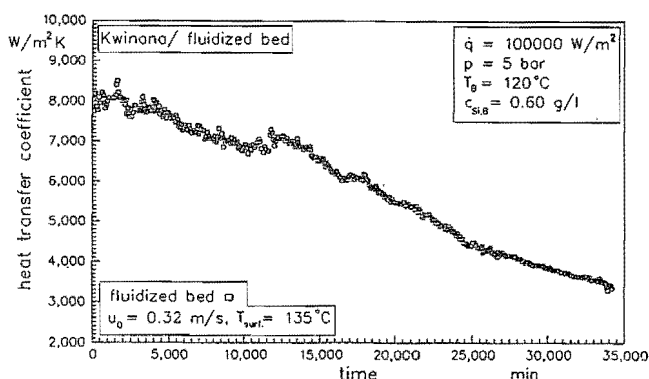


Fig 21 Heat transfer coefficient in the fluidized bed test section as a function of time for heating of spent Bayer liquor

4.3 Other Methods

Currently, there are several devices on the market which claim to reduce fouling by magnetic, electric, radiative or catalytic treatment. Parkinson and Price⁽²⁸⁾ and Donaldson and Grimes⁽²⁹⁾ report that the installation of magnets considerably reduced cooling water fouling, whereas Hasson and Bramson⁽³⁰⁾ and Söhnel and Mullin⁽³¹⁾ found no effect of the water conditioner. Plant measurements in a NZ dairy factory did not show any beneficial effects. According to Karabelas⁽³²⁾, fouling increased in pipe-lines for geothermal brine after magnetic mitigation devices had been installed. In the light of this evidence, it is difficult to come to any other conclusion than that the direct influence of a magnetic field on the process of crystallization is not very significant.

For severe fouling duties scraped surface heat exchangers may provide the shear stress necessary to reduce the adhesion of material.

6. CONCLUSIONS

For many years, fouling has been considered unavoidable and outside human responsibility. Heat exchanger design has been performed with unrelated fouling resistances which more often than not have been artificially increased to cover all design uncertainties. It is absolutely essential for heat exchanger users and designers to feed-back information to ascertain if there is any correlation between design fouling resistances and practical fouling problems.

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 lowest investment costs and can relatively easily be installed in operating plants. However, long-time operating costs caused by the need for continuous supply of chemicals and by 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 already at the design stage of the heat exchanger, considerable savings are possible since no provisions have to be made for excess heat transfer surface area and excess weight.

Numerous experimental and theoretical investigations have been performed in the past decade to increase the understanding of fouling mechanisms and to provide correlations for the effects of process parameters on the

fouling rates. Attempts should be made to implement the results of these studies into heat exchanger design procedures. Fouling is a transient process and future plant design will have to take this into consideration.

NOMENCLATURE

a	defined by eq. (11)
A	constant in eq. (7)
b	defined by eq. (12)
c	concentration, kg/m ³
c	defined by eq. (13)
d	pipe diameter, m
d _p	particle diameter, m
E	activation energy, J/mole
k _c	constant in eq. (15)
k _r	reaction rate constant
K	constant
K ₁	first molar dissociation constant of carbonic acid, mole/l
K ₂	second molar dissociation constant of carbonic acid, mole/l
K ₃	constant in eq. (14)
K ₄	constant in eq. (14)
K _{sp}	molar solubility product of calcium carbonate, (mole/l) ²
m	mass per area, kg/m ²
m	mass flux, kg/m ² s
n	reaction order
R	universal gas constant, 8314 J/mole • K
R _f	fouling resistance, m ² K/W
s	deposit thickness, m
t	time, s or min or h
t _c	time constant, min
T	temperature, °C or K
u	flow velocity, m/s
U	overall heat transfer coefficient, W/m ² K
v _t	thermophoretic velocity, m/s
α	film heat transfer coefficient, W/m ² K
β	mass transfer coefficient, m/s
λ	thermal conductivity, W/mK
ν	kinematic viscosity, m ² /s
ρ	density, kg/m ³
τ	wall shear stress, N/m ²
ξ	friction factor

subscripts - superscripts

a	adhesion
b	bulk
d	deposit
l	liquid
opt	optimum
r	removal
s	surface
t	transport
w	pipe wall
1	tube inside
2	tube out side
*	at saturation
*	asymptotic

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