TUBE MATERIAL AND AUGMENTED SURFACE EFFECTS IN HEAT EXCHANGER SCALING

By

ROYA SHEIKHOLESLAMI

B.Sc., The University of Kansas, 1980

A THESIS IN PARTIAL FULFILLMENT OF THE REQUIREMENTS FOR THE DEGREE OF MASTER OF APPLIED SCIENCE

in

THE FACULTY OF GRADUATE STUDIES (Department of Chemical Engineering)

We accept this thesis as conforming to the required standard

THE UNIVERSITY OF BRITISH COLUMBIA

April 1984

(c)Roya Sheikholeslami, 1984

In presenting this thesis in partial fulfilment of the requirements for an advanced degree at the University of British Columbia, I agree that the Library shall make it freely available for reference and study. I further agree that permission for extensive copying of this thesis for scholarly purposes may be granted by the head of my department or by his or her representatives. It is understood that copying or publication of this thesis for financial gain shall not be allowed without my written permission.

Department of <u>Chemical Engineening</u>

The University of British Columbia 1956 Main Mall Vancouver, Canada V6T 1Y3

Date <u>APR, 27, 1984</u>

DE-6 (3/81)

ABSTRACT

The performance of stainless steel, copper and mild steel plain heat exchanger tubes and an externally-finned mild steel tube was studied under calcium carbonate scaling conditions at constant heat flux. Artificial hard waters made up with sodium bicarbonate and calcium salts to alkalinities mg $CaCO_3/1$ and calcium hardness to 650 mg $CaCO_3/1$ 350 up to were recirculated through a steam heated annular test section for periods of 70 hours. The effect of velocity on the rate of heat flow, the overall heat transfer coefficient, the fouling resistance and the fouling rate has been determined. Fouling resistance increased with time in a linear manner in the majority of runs although falling rate and asymptotic behaviour were also Initial scaling rates were compared with the predictions of observed. Hasson's ionic diffusion model.

Generally, scaling decreased in extent as the tube material was changed from mild steel to copper to stainless steel, although there were some operating conditions where this trend was not followed. No firm conclusion could be drawn concerning initial fouling rates for these experiments which were done primarily at low levels of water alkalinity.

The longitudinally finned mild steel tube, having fin and total surface efficiencies of 33_{Δ} and 53_{Δ} , respectively, was examined under the same operating conditions as for the plain mild steel tube. In addition to higher values of heat flow rate, the former had higher values of both

the clean and the dirty overall heat transfer coefficients along with the lower values of fouling resistance, all based on the nominal (bare-tube) outside area. The deposits were thicker on the prime surface. The fouling process appeared to be more gradual on the finned tube than on the plain tube. However, the model predictions suggested a slightly higher rate for the finned tube at the same velocity.

For all tubes, the clean and the dirty overall heat transfer coefficients and the predicted fouling rates increased with velocity. Generally, the fouling resistances decreased with increasing velocity except for the copper tube at high alkalinity. No generalization could be made regarding the relationship between the experimental values of fouling rate and the velocity for either of non-corroding plain tubes. However, for corroding tubes, as the velocity increased, the fouling rate decreased.

The finned tube appears to be the most suitable choice in the presence of hard water scaling. In the velocity region tested, the model can be safely applied to predict the scaling rate of the copper and both mild steel tubes at alkalinities of about 350 since it over-predicted the experimental values; however it does not predict the effect of velocity.

TABLE OF CONTENTS

			Page
Abst List List Ackr	tract t of 1 t of F nowled	Tables	ii vii viii viii xii
ı.	INTRO	DDUCTION	٦
2.	THEOR	RY	5
	2.1	Types of Fouling	5
	2.2	Fouling Behaviour	6
	2.3	Precipitation Fouling	8
	2.4	Calcium Carbonate Precipitation	11
	2.5	Models Predicting the Rate of CaCO $_3$ Precipitation	17
	2.6	Hasson's Ionic Diffusion Model	23
	2.7	Factors Effecting Precipitation Fouling	30
		2.7.1Velocity2.7.2Heating Surface2.7.3Temperature2.7.4Water Chemistry	30 34 35 37
	2.8	Heat Transfer from Plain Surfaces	43
	2.9	Heat Transfer from Extended Surfaces	48
	2.10	Comparison Between an Enhanced and a Plain Heat Transfer Surface	55
3.	EXPE	RIMENTAL APPARATUS	59
	3.1	Water Flow Loop	59
	3.2	Supply Tank	61
	3.3	Steam System	61
	3.4	Tube Material and Geometry	63
	3.5	Temperature Measurements	67

- iv -

			Page
4.	EXPE	RIMENTAL PROCEDURES	70
	4.1	General Approach	70
	4.2	Solution Preparation	71
	4.3	pH Measurements	72
	4.4	Alkalinity Measurements	72
	4.5	Hardness Measurements	73
	4.6	Determination of Total Dissolved Solids	74
	4.7	Procedure for a Scaling Run	76
	4.8	Cleaning	80
5.	RESU	ILTS AND DISCUSSION	82
	5.1	General Outlook	82
		5.1.1 Low Concentration Runs	82 8 5
	5.2	Effect of Reynolds Number on Overall Heat Transfer Coefficient	96
		5.2.1 Copper and Stainless Steel Tubes	96 100 104
	5.3	Fouling Resistance With Respect to the Time	10 7
	5.4	Effect of Reynolds Number on Fouling Resistance	120
		5.4.1 Copper and Stainless Steel Tubes	120 122 124
	5.5	Determination of the Fouling Rate	126
		5.5.1 Measured Fouling Rate	126 126

- v -

													Page
	5.6	Effect	; of Rey	nolds Nur	nber or	n the	Measu	red I	Fouli	ng Ra	te.	•	128
		5.6.1 5.6.2 5.6.3	Coppe Coppe Plain	and Sta and Pla and Finne	inless in Milc ed Milc	Steel Stee Stee	Tube: 1 Tube 1 Tube 1 Tube	s. es. es.	•••	•••	•••	•	128 128 132
	5.7	Effect Foulir	; of Rey ng Rates	vnolds Nur	mber or	the	Predi	cted		••	• •	•	134
	5.8	Compar Foulir	rison Be ng Rates	etween the	e Predi	cted	and M	easu •••	red ••••	••	• •	-	140
6.	CONCL	USION	• • •	• • • •		• • •	• •	••		••	••	•	145
7.	NOME	NCLATUR	ε			•••	••	••	• • •	••	• •	•	149
8.	REFE	RENCES	• • •	••••	• • •	• • •	••	••		••	••	•	156
APPI	ENDIX	Ι.	CALIBR	TION OF	THERMOO	COUPLE	ES AND	ROT	AMETE	RS .	• •	•	165
APPI	ENDIX	II.	SAMPLE	CALCULAT	IONS .	• • •	••	••	• • •	••	••	•	169
			II.1 II.2	Determin Determin	ation c ation c	of Hea of Hea	at Flo at Tra	w Ra nsfe	te r Are	a Tompo	•••	•	169 170
			II.3 II.4	Differen	ce ation (of Ove	erall	Heat	Tran	sfer	•••	•	170
			II.5 II.6	Coefficie Determin Determin	ent ation c ation c	of the	e Foul Wall	ing Tem	 Resis perat	tance ure.	•••	• •	171 171 174 174
			II.8	Determin	ation (of the	e Reyn	olds	Numb Numb	er.	••	•	176
			II.10 II.11	Paramete Numerica Numerica Deposit	rs 1 Examp 1 Deter Thickne	ole Us rminat ess.	sing R tion o	un 2 f Fi	6 n Eff	icien	icy a	• • nd	177 178 182
APPI	ENDIX	III.	COMPUTI	ER PROGRA	MS		•••	••		• •	• •	•	184
APP	ENDIX	I۷.	FOULIN	G CURVES.	• • •	• • •	• • •	••	• • •	••	• •	•	193

LIST OF TABLES

Table		Page
2-1	Some Fouling Rate Models for Precipitation Fouling	18
2-2	Effect of Temperature and Velocity on Fouling Factor	49
3-1	Summary of Properties of Tubes	66
5-1	Summary of Results for Runs at Low Concentration of Chemicals	8 6
5-2	Summary of Results for Runs at High Concentration of Chemicals	91
5-3	Summary of Results for Stainless Steel Tube	98
5-4	Summary of Results for the Copper Tube at Low Concentration of Chemicals	99
5-5	Summary of Results for the Copper Tube at High Concentration of Chemicals	102
5-6	Summary of Results for the Plain Mild Steel Tube	103
5-7	Summary of Results for the Finned Mild Steel Tube	105
5-8	Summary of Model Calculations	127
5-9	Comparison of Measured Fouling Rates on Copper and Stainless Steel	129
5-10	Comparison of Predicted Fouling Rates on Copper and Stainless Steel	136
5-11	Comparison of Predicted and Measured Fouling Rates on Copper	141
5-12	Comparison of Predicted and Measured Fouling Rates on Stainless Steel	141
5-13	Comparison of Predicted and Measured Fouling Rates for High Concentration Runs	142
APPENDIX	I	
I-1	Calibration Table for Thermocouples	165
I-2	Constants Corresponding to the Calibration Equation for Thermocouples	166

.

.

- vii -

- viii -

•

LIST OF FIGURES

Figure		Page
2-1	Characteristic Fouling Curves	9
2-2	Equilibrium Distribution Fractions of Total Carbon at Various pH Levels	16
2-3	Solubilities of Hydroxides and Calcium Carbonate With Respect to pH	38
2-4	Sketch and Nomenclature of a Rectangular Fin	52
3-1	Flow Diagram of Apparatus	60
3-2	Photograph of Apparatus from the Back	62
3-3	Sketch of a Twelve Longitudinally Finned Tube	64
3-4	Horizontal (a) and Vertical (b) Cross Sections of a Finned Tube	65
3-5	Circuit Diagram for Datalogger Connection	68
3-6	Photograph of Apparatus from the Front	69
5-1	Run 5 Typical Output	84
5-2	Run 16 Typical Output	88
5-3	Run 24 Typical Output	90
5-4	Photograph of Disassembled Fouled Copper Tube	93
5-5	Photograph of Disassembled Fouled Plain Mild Steel Tube	94
5-6	Photograph of Disassembled Fouled Finned Mild Steel Tube .	95
5-7	Effect of Reynolds Number on Clean and Dirty Overall Heat Transfer Coefficients of Copper and Stainless Steel Tubes	97
5-8	Effect of Reynolds Number on Clean and Dirty Overall Heat Transfer Coefficients of Copper and Plain Mild Steel Tubes	101

Figure		Page
5-9	Effect of Reynolds Number on Clean and Dirty Overall Heat Transfer Coefficients of Plain and Finned Mild Steel Tubes	106
5-10	Run 2 Fouling Resistance Versus Time	109
5-11	Run 4A Fouling Resistance Versus Time	110
5-12	Run 20 Fouling Resistance Versus Time	111
5-13	Run 26 Fouling Resistance Versus Time	112
5-14	Run 36 Fouling Resistance Versus Time	113
5-15	Run 1 Fouling Resistance Versus Time	114 .
5-16	Run 6 Fouling Resistance Versus Time	115
5-17	Run 23 Fouling Resistance Versus Time	116
5-18	Run 27 Fouling Resistance Versus Time	117
5-19	Run 31 Fouling Resistance Versus Time	118
5-20	Run 38 Fouling Resistance Versus Time	119
5-21	Effect of Reynolds Number on Fouling Resistance for Copper and Stainless Steel Tubes	121
5-22	Effect of Reynolds Number on Fouling Resistance for Copper and Plain Mild Steel Tubes	123
5-23	Effect of Both Velocity and Reynolds Number on Fouling Resistance for Plain and Finned Mild Steel Tubes	125
5-24	Effect of Reynolds Number on Measured Fouling Rate for Copper and Plain Mild Steel Tubes	131
5-25	Effect of Both Velocity and Reynolds Number on Measured Fouling Rate for Plain and Finned Mild Steel Tubes	133
5-26	Effect of Reynolds Number on Predicted Fouling Rate for Copper and Stainless Steel Tubes	135

	- x -	
Figure		Page
5-27	Effect of Reynolds Number on Predicted Fouling Rate for Copper and Plain Mild Steel Tubes	138
5-28	Effect of Both Velocity and Reynolds Number on Predicted Fouling Rate for Plain and Finned Mild Steel Tubes	139
5-29	Comparison of Measured Scaling Rate With Rate Predicted by the Ionic Diffusion Model	144
APPENDIX	I	
I-1	Calibration Curve for Large Rotameter	167
I-2	Calibration Curve for Small Rotameter	168
APPENDIX	III	
111-1	Program to Evaluate Overall Heat Transfer Coefficients and Fouling Resistances	185
III-2	Program to Linearly Fit Fouling Resistances	186
III-3	Program to Asymptotically Fit Fouling Resistances	187
III-4	Program to Fit Fouling Resistances to a Polynomial Function	188
III-5	Program to Plot the Fouling Resistance Data and the Best Fit	189
III-6	Program to Determine the Rates Predicted by the Hasson's Ionic Diffusion Model	191
APPENDIX	IV	
IV-1	Run 1A Fouling Resistance Versus Time	193
IV-2	Run 2A Fouling Resistance Versus Time	194
IV-3	Run 3A Fouling Resistance Versus Time	195
IV-4	Run 3 Fouling Resistance Versus Time	196

Figure

APPENDIX IV

IV-5	Run	5A	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	197
IV-6	Run	5 F	Fouling F	Resistance N	lersus 1	lime .	•	•	•	•	•	•	•	•	•	•	198
IV-7	Run	6A	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	199
IV-8	Run	21	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	200
IV-9	Run	22	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	201
IV-10	Run	24	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	202
IV-11	Run	25	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	203
IV-12	Run	28	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	204
IV-13	Run	29	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	205
IV-14	Run	30	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	206
IV-15	Run	32	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	207
IV-16	Run	33	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	208
IV-17	Run	37	Fouling	Resistance	Versus	Time.	•	•	•	•	•	•	•	•	•	•	209

•

Page

ACKNOWLEDGEMENTS

I would like to thank Dr. A.P. Watkinson for his patience and conscientious supervision during the course of this study.

I am also very grateful to the members of my family for their unrelenting support.

My special thanks go to Dr. B.D. Bowen for his valuable technical advice in computer graphic plotting of data.

My thanks also go to the personnel of the workshop, electronic shop and stores, and especially to Mr. H. Lam for his assistance in the acquisition of parts and chemicals.

1. INTRODUCTION

During heat transfer operation with most liquids and some gases, an undesirable film gradually builds up at phase interfaces. The accumulation of this solid material can occur both on fluid-solid and fluid-liquid interfaces. However, this thesis will only deal with the fluid-solid interfaces.

In cooling systems, the deposition process can be categorized as either scaling or fouling. Scaling refers to the crystallization and precipitation of dissolved salts on the heat exchanger surfaces. Fouling is a more general term, and is used here to include deposition of scale or non-scale forming substances.

Every salt has a solubility limit at a given temperature. If this limit is exceeded, precipitation will occur given a favourable site. Most salts have solubility curves for which solubility increases as temperature rises. However, some salts show "inverse" solubility characteristics. For these salts, solubility decreases with increasing temperature over some range. If water containing such a scale-forming salt enters a heat exchanger and is subsequently heated, frequently the solubility limit will be exceeded and deposition of the scaling material will take place on the hot heat exchanger surface. In other words, precipitation fouling is primarily due to presence of supersaturated salts under the process condition. The thermal conductivity of the solid deposit residing on the heat exchanger surface is usually much less than the metal wall. Therefore, this scaling process would increase the total thermal resistance, decrease the overall heat transfer coefficient, and hence the performance of the heat exchanger will deteriorate.

The thermal resistance of the accumulated scale is called the fouling factor or fouling resistance (R_f) . Fouling factors for tubular heat exchangers can be found in the Tubular Exchanger Manufacturers Association (TEMA) standards (1). Since fouling resistance increases with time, a period of one year is usually chosen as a basis for reporting these numerical values.

Even though the scaling process is an unsteady state operation, the fouling factor is added indiscriminately to the steady state heat transfer resistances to calculate the total resistance of the unit. The following formula is the design equation for a fouled plain heat exchanger surface.

$$q = \frac{1}{(R_{h_{o}} + R_{w} + R_{h_{i}} + R_{f_{i}} + R_{f_{o}})} A\Delta T_{m}$$
(1-1)

At a given q and ΔT_m additional surface area is added to compensate for the increased thermal resistance due to fouling.

Fouling is a costly phenomenon if all operating, capital, maintenance and production expenses are considered. Due to the insulating effect of the deposits, more fuel energy may be needed to supply the required

- 2 -

heat transfer rate. In addition, increased pumping costs due to the reduced cross-sectional area or roughness would increase the operating expenses. There is also a rise in capital costs in case of additional heat exchange surface provided in anticipation of fouling. Manual or even chemical cleaning of the equipment results in high maintenance costs. Furthermore, there are production losses during the plant downtime.

Despite the difficulty in obtaining the accurate measurements regarding fouling expenses, Thackery (2) has suggested, in his recent study at AERE Harwell, England, that the fouling cost is about \$0.73 - \$1.20 billion per year in the United Kingdom. He has estimated that about 41% of this amount is due to the increased operating cost as the result of energy losses. About 21% of the fouling cost is due to increased capital expenses. About 17% and 21% of the fouling expenses are attributed to maintenance costs and production losses, respectively.

Since precipitation fouling is a major problem in industrial systems, the ability to predict the rate and extent of thermal scaling processes can be of a great help in evaluation of the performance of a heat exchanger. In cooling water systems, CaCO₃ is one of the major scale-forming salts. The deposition rate of CaCO₃ based on the radial diffusion of Ca⁺⁺ and $C\bar{O}_3^-$ ions from the bulk of the fluid toward the hot heat exchange interface has been predicted by Hasson et al. (3). In this thesis, one objective is to compare experimental scaling rates of CaCO₃ on different tubes with the predictions of the ionic diffusion model of Hasson. To see

- 3 -

if the data for plain surfaces could be used to predict what would happen on finned surfaces, an attempt is also made to study the effect of extended surfaces on both the scaling rate and the scaling resistance. In addition, the effect of fluid velocity on the fouling process is examined. 2. THEORY

2.1 Types of Fouling

Fouling has been classified into six different categories by Epstein (4). This classification has been carried out according to the "immediate cause of the fouling".

- Scaling or Precipitation Fouling: the precipitation of inverse solubility salts which are supersaturated under the process condition.
- 2. Particulate Fouling: the deposition of suspended solids present in a fluid onto a heat exchange interface. This process could be gravity controlled and called "Sedimentation Fouling" in case of a horizontal heat transfer surface.
- 3. Chemical Reaction Fouling: solid materials are formed as a result of chemical reactions which occur at the heat transfer surface. The surface material usually acts as a catalyst rather than as a reactant. This type of fouling is often encountered in petroleum refineries.
- 4. Corrosion Fouling: deposits are made of the corrosion products resulting from chemical reaction of the heat transfer surface and the process fluid. The fouled layer may assist the deposition of other fouling material.

- 5. Biofouling: the accumulation of biological organisms and their generated slimes onto a heat transfer surface.
- 6. Freezing Fouling: deposits are formed due to the solidification of a pure liquid or components of a liquid solution as a result of sub-cooled heat transfer surface.

The first five types of fouling may be synergistic. In the case of categories 1-5, the fouling process is enhanced with increasing the temperature. However, freezing fouling can only be accomplished by a temperature drop in the process fluid.

2.2 Fouling Behaviour

Fouling proceeds by a sequence of six fundamental mechanisms:

- Initiation which can be referred to as nucleation, induction, incubation or surface conditioning depending upon the type of fouling involved. This step may produce a delay period before thermal effects are manifest.
- Transport of the foulant from the bulk of process fluid toward the heat transfer surface.
- Adsorption, attachment or adhesion of the fouling material to the surface.

- 4. Accumulation of the deposits at the heat transfer interface.
- 5. Detachment, re-entrainment, scouring, erosion or sloughing-off of the deposits from the surface back into the bulk fluid.
- Aging such as crystal dehydration or chemical degradation of the deposits.

The fouling behaviour is usually represented by the fouling resistance (R_f) which is a function of time. The most common type of fouling curves follow linear, asymptotic or falling rate behaviour. The net rate of fouling should be defined first in order to understand any kind of fouling behaviour. Since fouling occurs due to two physico-chemical processes called accumulation and re-entrainment of the deposits, the net fouling rate (\hbar) can be specified by the following equation (5, 6).

$$\frac{\mathrm{d}\mathbf{m}}{\mathrm{d}\mathbf{\Theta}} = \mathbf{m} = \mathbf{m}_{\mathrm{d}} - \mathbf{m}_{\mathrm{r}} \tag{2-1}$$

where \dot{m}_d and \dot{m}_r are the gross rates of deposition and removal of the foulant, respectively.

The linear curve exhibits the fouling behaviour in which there is no re-entrainment process or the rate of deposition exceeds the rate of removal. In case of asymptotic fouling, the fouling resistance increases up to a point at which the net fouling rate equals zero due to an increase in the removal rate or a decrease in the deposition rate. Asymptotic

- 7 -

behaviour might result in indefinite operation of the unit with no need for cleaning. The third mode of behaviour (falling rate) is a transition between the asymptotic and linear modes. The three different types of fouling curves are shown in Figure 2-1.

2.3 Precipitation Fouling

As was mentioned earlier, precipitation fouling occurs due to supersaturation of sparingly soluble salts under the process condition. Supersaturation can be achieved as the result of one of the following processes:

- Evaporation of a solution beyond the solubility limit of a dissolved salt.
- 2. Mixing of two different streams can cause supersaturation as in the case of phosphoric acid preparation. 41.5 - 44.5%phosphoric acid saturated with $CaSO_4$ is produced by the addition of phosphate rock to the sulfuric acid. The solution is concentrated in the evaporators, operated at $70^{\circ}C$, to get 70 - 74% H₃PO₄. Since the solubility of $CaSO_4$ changes at different acid concentrations (decreases in this case), bulk precipitation of the salt takes place.



Figure 2-1. CHARACTERISTIC FOULING CURVES

3. Cooling down or heating up normal or inverse solubility salts, respectively. Supersaturation can occur during the heat transfer operation with respect to the heat transfer surface, the bulk of the process fluid or both.

Following the creation of supersaturation, the ultimate accumulation of the deposits occurs according to the six fundamental mechanisms mentioned earlier. In precipitation fouling, the delay time (e_n) is due to nucleation phenomena. Growth of the crystals follows the nucleation step. Since an increase in temperature would lead to a higher degree of supersaturation, it is obvious that the delay time would decrease with increasing temperature. This effect was statistically proved by Banchero and Gordon (7). Diffusion of the ionic species and particulate solids, in case of their presence in the bulk fluid, onto the heat exchanger wall will succeed the crystal growth. The deposits will be absorbed by and accumulate at the solid-layer interface. According to Hasson (8), a removal process would take place simultaneously with the deposition process, due to the fluid shear stress. Aging processes can cause the material already deposited on the heat exchange surface to recrystallize and harden due to a temperature increase. Also, as the result of the shedding-off mechanism caused by stresses, the deposits can be weakened with the passage of time.

2.4 Calcium Carbonate Precipitation

The common ion effect determines the saturation concentration of sparingly soluble salts such as $CaCO_3$ according to the following formula:

$$\begin{bmatrix} C_{a}^{++} \end{bmatrix} \begin{bmatrix} C_{0}^{-} \end{bmatrix} = K_{sp}$$
(2-2)

where K_{sp} is the temperature dependent concentration solubility product. The bracketed parameters are the saturation concentrations, and the activity coefficient is unity. Therefore, supersaturation occurs when the concentration of the ion increases beyond its saturation concentration with respect to that salt:

$$[C_{a}^{++}] > [C_{a}^{++}]_{s} = \frac{K_{sp}}{[C\bar{0}_{3}]_{s}}$$
 (2-3)

or

$$[c\bar{0}_{3}] > [c\bar{0}_{3}]_{s} = \frac{K_{sp}}{[c_{a}^{++}]_{s}}$$
 (2-4)

 K_{sp} is significantly affected by temperature. Larson and Buswall (9) have used the experimental data of other investigators such as Frear and Johnston (1929) to predict the solubility product of CaCO₃ (10). The following relationship represents the temperature dependency of K_{sp} between 0°C and 80°C for unit activity coefficient:

$$pK_{sp} = 0.01183T + 8.03$$
 (2-5)

where T is in degree centigrade. Considering the above-mentioned equation, it is obvious that K_{sp} decreases with increasing temperature. Hence, precipitation of $CaCO_3$ in a solution will take place due to either a decrease in K_{sp} as a result of temperature increase or addition of a common ion $(C_a^{++} \text{ or } C\bar{O}_3^{-})$.

When a soluble salt of calcium is added to a bicarbonate solution, the following reactions would take place:

$$H_2^0 \longrightarrow H^+ + 0\bar{H}$$
 (2-6a)

$$HC\bar{O}_3 + H^+ \longrightarrow H_2CO_3 = CO_2 + H_2O$$
 (2-6b)

$$HC\bar{O}_3 \longrightarrow C\bar{O}_3^- + H^+$$
 (2-6c)

$$C\bar{O}_{3}^{-} + Ca^{++} - CaCO_{3}$$
 (2-6d)

Thus, the total carbon concentration (C_T) is given by:

$$C_{T} = [HC\bar{O}_{3}] + [H_{2}CO_{3}] + [C\bar{O}_{3}]$$
 (2-7)

To determine the carbonate solubility, the chemistry of carbonic acid solution should be first understood. Since dissociation of strong acids and bases in pure water is complete, the hydrolysis effect of water is negligible and pH can be calculated using the molar mass of the substance added. However, in the case of weak acids or bases, the dissociation is not complete and the pH cannot be determined by the above-mentioned manner. Therefore, a reference pH is defined as the equivalence point which is the pH of an equivalent solution established by adding, say, x moles of a weak acid or base to a litre of pure water. This pH or the equivalence point depends on the amount of substance added to the water.

The concept of equivalence point becomes important when mixing of a weak acid or its salts with a strong base or strong acid, respectively, takes place. Calcium carbonate is the salt of a diprotic weak acid (H_2CO_3) . Carbonic acid has three different equivalence points with respect to H_2CO_3 , $HC\bar{O}_3$ and $C\bar{O}_3^-$. In a water sample, if the pH of solution is above the H_2CO_3 , $HC\bar{O}_3$ or $C\bar{O}_3^-$ equivalence points, the concentration of the monoprotic strong base which establishes that observed pH is called total alkalinity, phenolphthalein alkalinity and caustic alkalinity, respectively.

Therefore, based on the electroneutrality condition through the proton balance equation, the total alkalinity (T.A.) of a water with total carbon species (C_T) would be:

$$[T.A.] = [B^{+}] = [HC\bar{O}_{3}] + 2[C\bar{O}_{3}] + [O\bar{H}] - [H^{+}]$$
(2-8)

The concentration of a diprotic strong base which satisfies the same condition is:

$$[T.A.] = [B^+] = \frac{1}{2} [C_a^{++}]$$
(2-9)

In a solution containing $CaCO_3$, the distribution of the carbon species is significantly affected by the pH. The effect of pH can be investigated considering the following relationships obtained from the equilibria of carbonic acid.

$$\kappa_{1} = \frac{[H^{+}] [HC\bar{0}_{3}]}{[CO_{2}]}$$
(2-10)
$$\kappa_{2} = \frac{[H^{+}] [C\bar{0}_{3}]}{[HC\bar{0}_{3}]}$$
(2-11)

where K_1 and K_2 are, respectively, the first and second molar dissociation constants of carbonic acid. Solving three independent equations 2-8, 2-10 and 2-11, for three different unknowns would yield:

$$\begin{bmatrix} C\bar{0}_{3}^{-} \end{bmatrix} = \frac{\begin{bmatrix} T.A. \end{bmatrix} + \begin{bmatrix} H^{+} \end{bmatrix} - \begin{bmatrix} O\bar{H} \end{bmatrix}}{2 (1 + \frac{\begin{bmatrix} H^{+} \end{bmatrix}}{2K_{2}})}$$
(2-12)

$$[HCO_{3}^{-}] = \frac{[T.A.] + [H^{+}] - [O\overline{H}]}{(1 + \frac{2K_{2}}{[H^{+}]})}$$
(2-13)
$$[CO_{2}^{-}] = \frac{[T.A.] + [H^{+}] - [O\overline{H}]}{\frac{K_{1}}{[H^{+}]} (1 + \frac{2K_{2}}{[H^{+}]})}$$
(2-14)

And the distribution fractions of carbon species (i.e. $\alpha_{C\bar{0}_{3}} = \frac{[C\bar{0}_{3}]}{C_{T}}$) would be:

$${}^{\alpha}C\bar{0}_{3}^{-} = (1 + ([H^{+}]/K_{2}) + ([H^{+}]/K_{1}K_{2}))^{-1}$$
 (2-15)

$${}^{\alpha}\text{HCO}_{3} = (1 + ([H^{+}]/K_{1}) + (K_{2}/[H^{+}]))^{-1}$$
 (2-16)

$${}^{\alpha}CO_{2} = (1 + (K_{1}/[H^{+}]) + (K_{1}K_{2}/[H^{+}]^{2}))^{-1}$$
(2-17)

Thus, it is obvious that the distribution of carbon species in water is very pH dependent as is shown in Figure 2-2. With regard to equation 2-3, it is clear that the degree of supersaturation is also a function of both the total carbon concentration (C_T) and the pH.



Figure 2-2. EQUILIBRIUM DISTRIBUTION FRACTIONS OF TOTAL CARBON AT VARIOUS pH LEVELS (8)

2.5 Models Predicting the Rate of CaCO₂ Precipitation

The ability to predict the fouling rate is of great importance regarding both the design and the performance of a heat exchanger. Several predictive models have been presented with respect to both type of fouling and mode of operation. Some of these correlations involving precipitation fouling are shown in Table 2-1 and will be briefly discussed below.

Apparently, the first analytical study of fouling was carried out by McCabe and Robinson (11). Their experimental work in evaporators under constant ΔT (two isothermal fluids) was in a good agreement with their proposed model (equation 2-19) which predicted a falling rate behaviour. The model was based on the proportionality of the quantity of scale deposited with both the amount of liquid evaporated and the total heat transferred for accomplishing this process. Removal processes were not taken into consideration by these authors.

In 1959, Kern and Seaton (5, 6) presented their model (equation 2-20) considering both the deposition and removal processes in determining the net rate of fouling. Their mathematical correlation was based on the asymptotic behaviour of fouling in industrial heat exchangers. The following equation was proposed to approximate the fouling resistance.

$$R_{a} = R^{\star}(1-\bar{e}^{B\Theta})$$
 (2-18)

				•
Year	Authors	Fouling rate = (<u>dm</u>) = m de		Systems Described
1924	McCabe a Robinson (11)	£t = a ₀ = a ₁ U	(2-19)	In evaporators with constant ∆T
1959	Kern a Seaton (5, 6)	ћ = а ₂ С _Б V - Ъ ₁ тх _f	(2-20)	Particulate and other fouling
1962	Hasson (12)	mt = a ₃ u ^f	(2-21)	Non isothermal fluid
		$\mathbf{m} = \mathbf{k}_{0}(\mathbf{c_b} - \mathbf{c_s})$	(2-22)	CaCO ₃ precipitation
1964	Reitzer (15)	m ≖ K(c _b -c _s) ⁿ	(2-23a)	Precipitation of inverse solubility salts
		mt = a ₄ ∪ ⁿ	(2-23b)	With constant ΔT
		n = a ₅	(2-23c)	With constant heat flux
1968	Hasson et al. (17)	$\hat{m} = k_{BF}[(C_{a}^{++})_{b}^{-}(C_{a}^{++})_{1}]$	(2-24a)	For CaCO ₃ scaling with
		m = a ₆ (Re) ^{0.716} fflm	(2-24b)	constant heat flux.
1972	Taborek et al. (18, 19)	$\hat{\mathbf{m}} = \mathbf{a}_7 \mathbf{P}_d^9 \Omega \exp\left(\frac{(\mathbf{-E})}{\mathbf{R}_g \mathbf{I}_s} - \mathbf{b}_2 \frac{\tau}{\Psi} \mathbf{x}_f^r\right)$	(2-25)	Cooling water
1975	Watkinson a Martinez (20)	f = K(c _b -c _s) ⁿ -b ₃ τx _f	(2-26)	CaCO, scaling at
		= K $\left[a_{8} \frac{(T_{w}-T_{b})}{1+(hx_{f}/k_{f})}^{n}\right] = b_{3} \tau x_{f}$		constant steam temperature

Table 2-1. SOME FOULING RATE MODELS FOR PRECIPITATION FOULING

where R^* is the asymptotic value of the fouling resistance and B is a constant in time⁻¹. The gross deposition rate was considered to be a function of fluid flow (V) and the concentration of foulant in bulk (c_b).

```
deposition rate = a_2 c_b V
```

The removal rate was proposed to be a function of fluid shear stress, τ , and thickness of the scale, $x_{\rm f}.$

The mathematical model (equation 2-21) expressed by Hasson (12) involved sensible heating of cooling water. The model results in an infinite fouling resistance buildup due to the lack of removal term. His experimental analysis for tube side fouling at low velocities (v<1 m/sec) confirmed the predictive model using a value of f=2.5. However, at higher velocities, the correlation does not hold since the removal term becomes effective. In the same paper, Hasson has also formulated a growth rate (equation 2-22) for CaCO3 precipitation based on the mass transfer principles (13). The overall mass transfer coefficient K. was considered to be a function of the diffusion of HCO₃ and Ca⁺⁺ ions through the boundary layer and the formation of a crystal lattice.

$$\frac{1}{K_{o}} = \frac{1}{K_{D}} + \frac{1}{K_{R}}$$
(2-27)

where K_D and K_R are, respectively, the mass transfer coefficient for diffusion and the rate coefficient for surface reaction. K_D can be estimated using the Sherwood equation for turbulent flow inside pipes (14):

$$\frac{K_D D}{D_{AB}} = 0.023 (Re) \frac{0.83}{film} (Sc) \frac{0.33}{film}$$
(2-28)

where D is the hydraulic diameter and D_{AB} is the diffusivity of $HC\bar{D}_3$ or Ca^{++} ions. K_R would be evaluated through the Arrhenius equation as follows:

$$\ln K_{\rm R} = J - \frac{E}{R_{\rm q} T_{\rm s}}$$
(2-29)

where R_g is the gas constant, T_s is the surface temperature, E is the activation energy and J is the constant in the Arrhenius equation.

Reitzer (15) proposed his model (equation 2-23a) for precipitation of inverse solubility salts based on the degree of supersaturation (16) considering a linear inverted solubility curve for a relatively small range of temperature. His model predicted an increasing, non-linear, fouling resistance build up for constant operating conditions according to the equation 2-23b. At a constant heat flux, a linear deposition rate was predicted as is shown in equation 2-23c. No removal term was involved in formulating this model. Hasson et al. (17) have also studied the $CaCO_3$ scaling using an annular constant heat flux heat exchanger. He suggested that in a diffusion controlled process, when the chemical reaction and surface crystallization rates are high, the overall growth rate is a function of Ca^{++} ion driving force for diffusion (equation 2-24a). Using the J-factor analogy between mass and heat transfer, he derived the net fouling rate as a function of Reynolds number (equation 2-24b). His experimental measurements based on the mass transfer data for waters having Reynolds numbers between 13000 and 42000 (24.8 < v < 82 cm/sec) provided the following rate equation:

$$\dot{m} = a_9(\text{Re})_{film}^{0.68}$$
(2-30)

which was in a close agreement with the model presented. However, the practical value of the model might be doubtful at higher velocities due to the absence of the removal term. It should be mentioned that a more recent model containing both the removal and the deposition terms was formulated by Hasson in 1978. This so-called "Ionic Diffusion Model" has taken into account the effect of water chemistry on the diffusion process and will be discussed separately.

The model (equation 2-25) proposed by Taborek et al. (18, 19) involved both deposition and removal processes. In this study, the effect of water chemistry on the fouling rate has been considered. The gross deposition term was assumed to be reaction controlled and hence a function of crystallization process. The effect of water chemistry was taken into account using the Langelier saturation index and represented by the, Ω ,

"water characterization factor". The effect of residence time of the foulant in the reaction zone which is an inverse function of the velocity, was taken into consideration through the, P_d , "probability function of velocity".

$$\hat{\mathbf{m}}_{d} = \mathbf{a}_{7} \mathbf{P}_{d}^{\Omega} \exp(-\mathbf{E}/\mathbf{R}_{g} \mathbf{T}_{s})$$

The gross removal rate was defined in terms of fluid shear stress, τ , and the deposit bond resistance, R_b · R_b was expressed as a function of both the deposition thickness, x_f , and the deposition structure, Ψ .

$$\mathfrak{m}_{r} = \mathfrak{b}_{2} \frac{\tau}{\Psi} (\mathfrak{x}_{f})^{r}$$

In 1975, Watkinson and Martinez (20) presented their model (equation 2-26) for scaling of $CaCO_3$ inside copper tubes. They used the Kern-Seaton concept of deposition and removal with the Reitzer deposition term, as the basis of their gross deposition rate. Considering a linear inverted solubility curve for the range of temperature involved, the gross rate of scale build up can be determined by:

 $m_d = K(c_b - c_s)^n = Ka_8^n(T_s - T_b)^n$

writing the scale surface temperature, T_s , in terms of the wall temperature, T_w , they derived:

$$\dot{m}_{d} = K \left[\frac{a_{8}(T_{w} - T_{b})}{1 + (hx_{f}/k_{f})} \right]$$

This model differs from those expressed by other investigators in that the gross deposition term is not constant but varies both with time and the scale thickness. Assuming an Arrhenius function for the rate of crystal growth, K, it shows that the gross rate of deposition is strongly temperature dependent. The removal term is considered to be proportional to both the deposit thickness and the fluid shear stress which is, in turn, a function of friction factor, fluid velocity and density ($\tau = \frac{fv^2\rho}{2}$).

2.6 Hasson's Ionic Diffusion Model

Even though water chemistry is an important factor in precipitation of $CaCO_3$, it hadn't been taken into consideration in the predictive models proposed before 1972. However, the model presented by Taborek et al. (19) has taken this factor into account. They have used the Langelier saturation index (LSI) as a parameter relating the water chemistry to the fouling rate. Since the index is rather a qualitative than quantitative measure of the tendency of water to either dissolve or precipitate the $CaCO_3$, Hasson has (8) criticized the use of the index and has formulated the ionic diffusion model.
His model (13) is based on the radial diffusion of C_a^{++} and $C\bar{O}_3^-$ (or $HC\bar{O}_3$) ions from the bulk of the fluid toward the hot heat transfer surface. He has disregarded the deposition by precipitated $CaCO_3$ present in the bulk fluid and has used the experimental data published by Morse and Knudsen (21) to verify the accuracy of his model. The rate of crystallization and gross deposition of $CaCO_3$ per unit area at the interface was expressed, based on the surface controlled mechanism (22, 23), by the following equation:

$$w = K_{R} \left(\left[Ca^{++} \right]_{i} \left[CO_{3}^{-} \right]_{i} - K_{sp} \right)$$
(2-31)

where bracketed parameters are the interfacial concentrations and K_{sp} is the solubility product of CaCO₃ at the scale water interface. The rate of crystal growth, K_{R} , is temperature dependent and assumed to follow the Arrhenius law. As was experimentally shown by Gazit and Hasson (24), it can be evaluated from the following relationship:

$$1nK_{R} = 38.74 - \frac{20700}{R_{g}T_{s}}$$
(2-32)

where R_g is the gas constant in cal/mole °K, T_s is the absolute temperature of the scale surface and K_R is in cm/sec/gr CaCO₃/cm³.

The distribution of various carbon species $(C\bar{O}_3, HC\bar{O}_3, CO_2)$ in a carbonate solution is pH dependent as was shown previously. Since the diffusion process depends on the concentration of all the diffusing species present in the water, first it should be determined which one of the species is predominant. Looking at equations 2-12, 2-13 and 2-14, it is obvious that for a given solution, most of the carbon is in the form of $HC\bar{O}_3$ ion at low pH values. However, the tendency toward formation of $C\bar{O}_3$ ions increases with increasing pH. Therefore, in the case of high pH values, the rate of diffusion will be controlled by $C\bar{O}_3$ ions concentration as follows:

$$w = K_{D}[(C_{a}^{++}) - (C_{a}^{++})_{i}] = K_{D}[(C_{3}^{-}) - (C_{3}^{-})_{i}]$$
(2-33)

The controlling species are either CO_2 or HCO_3 ions at low pH values giving rise to the following rate equation:

$$w = K_{D}[(C_{a}^{++}) - (C_{a}^{++})_{i}] = K_{D}[(CO_{2}) - (CO_{2})_{i}] = \frac{K_{D}}{2}[(HCO_{3}) - (HCO_{3})_{i}] \qquad (2-34)$$

where K_D represents the convective diffusion coefficient and can be considered the same for all above-mentioned carbon species due to their close diffusivity values. The numerical value of this parameter, K_D , can be approximated using the following relationship (14):

$$\frac{K_D}{v} \operatorname{Sc}^{2/3} = 0.023 \operatorname{Re}^{-0.17}$$
(2-35)

where v is the velocity and Sc and Re are Schmidt and Reynolds numbers, respectively.

Having eliminated the interfacial concentrations from equations 2-31 and 2-33 he obtained the following gross deposition rate for $CaCO_3$ scaling at high pH values.

$$w = \frac{K_{D}[C_{a}^{++}]}{2} (1 + \frac{[C_{0}^{-}]}{[C_{a}^{++}]} + \frac{K_{D}}{K_{R}[C_{a}^{++}]}) (1 - \sqrt{\frac{4[C_{0}^{-}]}{[C_{a}^{++}]} (1 - \frac{K_{sp}}{[C_{a}^{++}][C_{0}^{-}]})} (2 - 36) (1 - \sqrt{\frac{1 - \frac{K_{sp}}{[C_{a}^{++}]} (1 - \frac{K_{sp}}{[C_{a}^{++}]})} (1 - \frac{K_{sp}}{[C_{a}^{++}]}) (1 - \frac{K_{sp}}{[C_{a}^{++$$

If the $C\bar{0}_3^-$ concentration is much lower than the $C\bar{a}^+$ concentration, as is often the case, the following simplified equation would express the gross deposition rate.

$$w = K_{D}[C\bar{0}_{3}^{-}] = \frac{1 - \frac{K_{sp}}{[C_{a}^{++}] [C\bar{0}_{3}^{-}]}}{1 + \frac{K_{D}}{K_{R}[C_{a}^{++}]} + \frac{[C\bar{0}_{3}^{-}]}{[C_{a}^{++}]}}$$
(2-37)

At low pH values, equations 2-13, 2-14, 2-31 and 2-34 were used for the gross rate of scale build up resulting in the following relationship:

$$\frac{\kappa_{1}}{4\kappa_{2}}\left(\frac{w}{\kappa_{R}[c_{a}^{++}]^{2}} + \frac{\kappa_{sp}}{[c_{a}^{++}]^{2}}\right)\left(\frac{w}{\kappa_{D}[c_{a}^{++}]} + \frac{[c_{0}_{2}]}{[c_{a}^{++}]}\right) = \left(1 - \frac{w}{\kappa_{D}[c_{a}^{++}]}\right)\left(\frac{[Hc\bar{0}_{3}]/2}{[c_{a}^{++}]} - \frac{w}{\kappa_{D}[c_{a}^{++}]}\right)$$

$$(2-38)$$

2

Having the above-mentioned condition, $C_a^{++} >> C\bar{D}_3^-$, the equation will be reduced to:

$$w = \frac{K_{D}[c_{a}^{++}]}{2} \cdot \frac{b}{a} \cdot (\sqrt{1 + \frac{4ac}{b^{2}}} - 1)$$
 (2-39)

where:

$$a = 1 - \frac{4K_2}{K_1} \cdot \frac{K_R[C_a^{++}]}{K_D}$$
 (2-40)

$$b = \frac{[CO_2]}{[C_a^{++}]} + \frac{4K_2}{K_1} \cdot \frac{K_R}{K_D} \cdot [HC\bar{O}_3] + \frac{K_{sp}K_R}{K_D[C_a^{++}]}$$
(2-41)

$$c = \frac{K_2}{K_1} \cdot \frac{K_R [HC\bar{O}_3]^2}{K_D [C_a^{++}]} - \frac{K_{sp} [CO_2] K_R}{[C_a^{++}]^2 K_D}$$
(2-42)

Under the condition where the term in the square root is approximately 1, a further simplification results:

$$w = K_{D}^{[C\bar{0}_{3}]} \cdot \frac{1 - \frac{K_{sp}}{[C_{a}^{++}] [C\bar{0}_{3}]}}{\frac{K_{D}}{K_{R}^{[C_{a}^{++}]}} + \frac{4[C\bar{0}_{3}]}{[HC\bar{0}_{3}]} + \frac{K_{sp}}{[C_{a}^{++}] [C\bar{0}_{2}]}}$$
(2-43)

Hasson has used the Kern-Seaton type model for the overall rate of the scale build up. The removal term was expressed as a function of the fluid shear stress, the deposit thickness and mechanical strength of the adhering material. Therefore, the net growth velocity of the layer can be correlated by the following equation:

$$(dx_{f}/de) = (w/\rho_{f}) - (x_{f}\tau/M)$$
 (2-44)

where τ is the fluid shear stress, x is the deposit thickness and M is a constant representing the mechanical strength of the deposit.

Since the fouling resistance is the difference in the overall resistance of the clean and the fouled heat exchanger, it can be expressed by:

$$R_{f} = R - R_{o} = \frac{x_{f}}{k_{f}}$$
 (2-45)

where k_f is the thermal conductivity of the foulant. Differentiation of this equation yields:

$$\frac{dR_{f}}{d\Theta} = \frac{dR}{d\Theta} = \frac{1}{k_{f}} \frac{dx_{f}}{d\Theta} = \frac{w}{\rho_{f}k_{f}} - \frac{x_{f}\tau}{Mk_{f}} = \frac{w}{\rho_{f}k_{f}} - R_{f}\frac{\tau}{M}$$
(2-46)

Integration of the above-mentioned equation with the initial condition e=0, $R_f=0$ results in an asymptotic fouling behaviour represented by:

$$R_{f} = R_{f}^{*} [1 - \exp(\theta/\theta_{c})]$$
 (2-47)

where the adherent term (the time constant, $\boldsymbol{\theta}_{C})$ is given by:

$$e_{c} = M/\tau$$
 (2-48)

and the asymptotic fouling resistance is expressed by:

$$R_{f}^{*} = \frac{W}{\rho_{f}k_{f}} \cdot \Theta_{c}$$
 (2-49)

substituting equations 2-48 and 2-49 in equation 2-46 yields:

يد.

$$\frac{dR_{f}}{d\theta} = \frac{R_{f}}{\theta_{c}} - \frac{R_{f}}{\theta_{c}}$$
(2-50)

when the fluid shear stress is low or generally in case of very adherent deposits, $\boldsymbol{\Theta}_{c}$ becomes very large and the removal term diminishes resulting in a linear fouling behaviour as follows:

$$\frac{dR_{f}}{d\Theta} = \frac{R_{f}}{\Theta_{c}} = \frac{W}{\rho_{f}k_{f}}$$
(2-51)

2.7 Factors Affecting Precipitation Fouling

The fouling process is best defined by either the time dependence of the thickness of the fouled layer or the thermal fouling rate. Therefore, the effect of parameters such as velocity, surface material, temperature and water chemistry on the fouling rate will be discussed below.

2.7.1 Velocity

A decrease in the fouling rate with increases in velocity was usually considered to be in effect in design of heat exchangers (25, 26). As was mentioned earlier, the net rate of fouling is a function of both the deposition and the removal rates. The importance of velocity effects in the removal process has been widely recognized. It has been suggested that the re-entrainment process takes place due to either the fluid shear stress (5,6) which is a function of velocity, friction factor and the flow configuration or to turbulent bursts (27, 28). According to Cleaver and Yates, removal will occur if the minimum or critical friction velocity (v^*) is exceeded. The critical friction velocity is described by:

$$v^{*} = \sqrt{\frac{\tau}{\rho}}$$
 (2-52)

where τ and ρ are the fluid shear stress at the surface and the fluid density, respectively. For the occurrence of a removal process, Hasson et al. (17) have reported the minimum velocity value of 2.6 cm/sec while Watkinson and Martinez (26) have suggested a minimum value of 2.9 cm/sec for v* in CaCO₃ scaling.

Nucleation, diffusion and chemical reaction rate processes will determine the gross rate of deposition. In precipitation of $CaCO_3$, the first scale layer is deposited on a metal surface followed by subsequent addition of the deposits on the scaled surface. In the case of $CaCO_3$ the rate of nucleation is considered to be constant and independent of the surface material (17); therefore, gross deposition rate is defined only in terms of significant operating parameters (diffusion and chemical reaction).

Precipitation of $CaCO_3$ is initiated by diffusion of Ca^{++} and $HC\bar{O}_3$ ions toward the hot heat transfer surface and followed by the following chemical reaction:

$$2HC\bar{O}_3 \longrightarrow C\bar{O}_3 + CO_2 + H_2O$$
 (2-53)

The subsequent crystallization of $CaCO_3$ on the hot surface would take place and be represented by:

$$C_{a}^{++} + C_{3}^{-} \xrightarrow{K_{R}} C_{a}C_{3}^{-} (s)$$
 (2-54)

The precipitation process would be completed if the lattice binding forces of the $CaCO_3$ crystal exceeds the shear force exerted by the liquid flow. It should also be mentioned that the dissolved CO_2 produced, according to equation 2-53, would either migrate towards the bulk or be released in gaseous form at the surface.

Since the rate coefficient (k') is relatively high, the rate of diffusion is controlled by the migration of C_a^{++} and $HC\bar{O}_3$ ions toward the surface. In addition, assuming that the generated CO_2 either escapes rapidly in gaseous form or diffuses toward the water bulk, the gross rate of deposition, \dot{m}_d , is governed by both diffusion and crystallization processes.

Hasson has experimentally shown (12) that diffusion is the effective factor at low velocities while the effect of surface reaction becomes important at relatively higher velocities. When the process is diffusion controlled, velocity enters the gross deposition term via the convective mass transfer coefficient (equation 2-28). Since there is not any appreciable removal process at lower velocity values, the net rate of scale build up would be constant with no asymptote and can be approximated through equations 2-22 and 2-27 by the following relationship for the overall mass transfer coefficient:

$$K_{0} \propto K_{D} \propto v^{0.83} \propto Re^{0.83}$$
 (2-55)

thus,

$$m^{\alpha} \text{Re}^{0.83}$$
 (2-56)

where \hat{m} is the net rate of scaling. Since the fouling process is specific to the operating conditions, the above-mentioned equation holds for a constant heat flow system where the scale surface temperature and hence K_R are invariant with respect to time.

Another theoretical model (equation 2-24b) suggested by Hasson et al. (17) represents the increase in the fouling rate with the fluid velocity. Their experimental results (equation 2-30) were in close agreement with their predictive model for relatively low velocities ($v \leq 0.82$ m/sec) and hence the absence of removal term.

The experimental results presented by Watkinson and Martinez (20) showed the positive effect of velocity on the asymptotic fouling resistance for Reynolds numbers below 12000 (v<0.5 m/sec). However, a decrease in the

asymptotic fouling resistance with the velocity was both observed (equation 2-57) and predicted (equation 2-58) for Reynolds numbers above 12000.

$$R_{f}^{*} \approx v^{-1.33}$$
 (2-57)

$$R_{f}^{*} + \frac{2}{h}R_{f}^{*2} + \frac{R_{f}^{*}}{h^{2}} - \frac{\beta_{2}}{(hv)^{2}} = 0$$
 (2-58)

where β_2 is temperature dependent and can be approximated for bulk temperatures below 350°k through:

$$\beta_2 = 0.000271T_b - 0.08489$$
 (2-59)

It should be mentioned that the system was operating under constant steam temperature.

2.7.2 Heating Surface

In $CaCO_3$ scaling, the heat transfer surface is neither a reactant nor a catalyst. The major influence of the heating surface on this kind of fouling process occurs during the induction period. Parameters such as average roughness, surface free energy, wettability and heat of immersion are used to characterize the nature of the solid surface. Corrosion of the surface material can also affect the fouling process. Some metallic surfaces may undergo corrosion and provide roughened areas as nucleation sites for deposition to initiate (29) or increase the fouling factor by the deposition of corrosion products (30).

Nucleation of $CaCO_3$ occurs at the metallic surface. Even though subsequent deposition of $CaCO_3$ will take place on $CaCO_3$ crystals, the solid interface continues to effect fouling process through the adhesion force of the fouled layer. However, experiments have shown (17) that, in the case of $CaCO_3$, the rate of nucleation on a metallic interface is almost equal to the rate of growth on a fouled $CaCO_3$ layer.

2.7.3 Temperature

Bulk temperature would effect the physical properties of the process fluid as well as all the reactions involved in the fouling phenomena. For an inverse solubility salt the temperature-solubility relationship can be approximated by a linear relationship

 $(c_{b}-c_{s}) = b'(T_{s}-T_{b})$ (2-60)

where c is the concentration, T is the temperature and subscripts b and s are denoted to the bulk and the scale surface, respectively. The suggested deposition models based on the concentration driving force (equations 2-22, 2-23a, 2-24a and 2-26) show that an increase in the bulk temperature would lower the diffusional driving force, at constant scale surface temperature, due to a decrease in the bulk solubility limit. However, this condition may give rise to bulk precipitation and hence to particulate fouling. The negative effect of the temperature rise on the diffusion process is also clear in equations 2-20 and 2-25 through bulk concentration and water characterization factor, respectively. The convective mass transfer coefficient (K_D) , for which the fluid properties will be measured at the film temperature, would also be affected by the bulk temperature.

As the scale surface temperature rises, the gross deposition rate increases through an increase in the concentration gradient (equation 2-60). The effect of T_s on both the convective mass transfer coefficient (K_D) due to the change in film temperature and the surface rate coefficient (K_R) is obvious according to equation 2-28 and the Arrhenius law (equation 2-29), respectively. The positive effect of raising the surface temperature is experimentally shown in the work of Hasson et al. (17) on CaCO₃ scaling.

Physical properties of the fouled layer such as adhesion to the solid surface and friability are greatly affected by the surface temperature. Under a constant heat flux condition where the scale surface temperature is constant, the wall temperature rises as the tube scales. The internal temperature of the fouled layer increases with increasing temperature and recrystallization of the deposited material is likely to occur as the time passes. This is the so called aging process and in this case would tend to strengthen the deposit and weaken the removal process.

- 36 -

2.7.4 Water Chemistry

Water properties such as carbon dioxide content, pH, hardness and total alkalinity play an important role in the deposition of calcium carbonate. The solubility of $CaCO_3$ in water, in the absence of CO_2 , is 14 ppm at 25°C. Addition of carbon dioxide would increase the $CaCO_3$ solubility to 300-400 ppm (31) via the following reaction:

$$CaCO_3 + CO_2 + H_2O \longrightarrow Ca(HCO_3)_2$$
 (2-61)

Solubility of CaCO₃ also changes with respect to pH of the liquid. At high pH values most of the carbon is in the form of $C\overline{O}_3^-$ ion (equation 2-12). Therefore, for a solution having the total carbon concentration of C_T and temperature of T, the degree of supersaturation increases with increasing pH (equation 2-3). Figure 2-3 shows the decrease in the solubility of CaCO₃ with increasing pH. The reduction of the asymptotic fouling factor due to a decrease in pH is experimentally shown in Knudsen and Roy's study of CaCO₃ precipitation on a 90-10 copper-nickel surface (30).

The presence of calcium and magnesium salts are usually considered as hardness. Equation 2-3 represents the positive effect of the hardness, due to C_a^{++} ions, on the degree of supersaturation and hence on the fouling process. Alkalinity is a function of $HC\bar{O}_3$, $C\bar{O}_3^-$ and $O\bar{H}$ ions. Looking at both equations 2-8 and 2-4, it is obvious that the supersaturation of $CaCO_3$



Figure 2-3. SOLUBILITIES OF HYDROXIDES AND CALCIUM CARBONATE WITH RESPECT TO pH (.8)

occurs more readily with increasing total alkalinity. The positive effect of alkalinity on the fouling resistance is experimentally shown by Morse and Knudsen (21).

Some methods including the effect of hardness, alkalinity and pH were suggested to relate the water chemistry with the fouling process. Langelier (32, 33) has developed a qualitative formula by which the tendency of water to either precipitate or dissolve calcium carbonate can be predicted. The equation is based on the pH, total alkalinity, hardness and temperature which are effective in solubility of calcium carbonate in water. The basis of his prediction involved in the reversible scaling process is:

$$C_{a}C_{3}(s) + H^{+} = C_{a}^{++} + HC\bar{O}_{3}$$
 (2-62)

The solubility constant, k, for the above-mentioned reaction can be expressed by:

$$k = \frac{[C_{a}^{++}] [HC\bar{D}_{3}]}{[H^{+}]}$$
(2-63)

Therefore, the saturation or equilibrium pH can be written:

$$pH_{s} = pC_{a}^{++} + pHC\bar{D}_{3} - pk$$
 (2-64)

considering both the second solubility constant, K_2 , for carbonic acid and the CaCO₃ solubility product, K_{sp} , yields:

$$pk = pK_2 - pK_{SD}$$
 (2-65)

using equation 2-13, PHCO₃ can be formulated in terms of pH and total alkalinity:

$$pHC\bar{O}_3 = p (T.A. + [H^+] - [O\bar{H}]) + log (1 + \frac{2K_2}{[H^+]}) (2-66)$$

substituting equations 2-65 and 2-66 into equation 2-64 yields:

 pH_s corresponds to the pH of a solution, with a given calcium carbonate content and total alkalinity, at which water is in equilibrium with CaCO₃. Hence, the Langelier Saturation Index, which is defined as:

 $LSI = pH-pH_{S}$ (2-68)

can be a qualitative measure of $CaCO_3$ solubility in water. If the index is positive, $CaCO_3$ tends to deposit. Negative values of the index represent the tendency of water to dissolve $CaCO_3$. The water is at equilibrium with $CaCO_3$ when the index is zero. Watkinson (34, 35) has experimentally shown that the asymptotic fouling resistance increases with increasing saturation index.

Ryznar (36) has criticized the use of saturation index as a qualitative measure of $CaCO_3$ solubility in the water. He has empirically developed an index which supposedly makes it possible to predict whether a water is scale forming or corrosive. This parameter was called Ryzner Stability Index and can be given by:

$$RSI = 2pH_{pH}$$
(2-69)

He has experimentally shown that a water having a stability index of approximately 6.0 or less is scale forming while a water having an index above 7 tends to be corrosive.

The presence of other chemical substances also changes the water chemistry and may in turn influence both the fouling process and the constituents of the scale. It has been shown (21) that the scale strength is generally proportional to the scale purity; therefore, the higher degree of purity accelerates the fouling rate. The experiments carried out by Heat Transfer Research Corporation, HTRI (37), have supported the purity effect, showing that in a solution of single salt, the crystalline structures are strongly adhered to each other.

The effect of magnesium on the fouling process has very briefly been studied. Thurston (38) reported that the presence of Mg^{+2} at concentrations higher than 0.2 Ca⁺⁺, in waters low in silica, would weaken the structure of CaCO₃ deposit by production of calcite in the suspension which depresses the supersaturation. Watkinson's experimental work (34) on the waters having $Mg^{2+}>0.2$ Ca²⁺ showed a decrease in both the fouling rate and asymptotic fouling resistance and resulted in a deposit containing less than 0.5% Mg^{++} content. However, in his further study, Watkinson (39) has reported that as the Mg^{++}/Ca^{++} increases, both the fouling resistance and the scaling rate would decrease and go through a minimum at Mg^{++}/Ca^{++} about 0.2. The Mg^{++} content of the deposits was about 1.6% by weight in the latter.

The presence of any impurity in a solution may also alter the precipitation process. Peters and Stevens (40) have reported that the existence of iron would decrease the crystal growth rate of $CaCO_3 - Mg(OH)_2$ precipitate while enhancing the nucleation rate. Since it has been observed that the crystallization of $CaCO_3$ and $Mg(OH)_2$ are independent of each other (41), it can be deduced that the same effect is applicable to the $CaCO_3$ precipitation. The increase in the solubility of $CaCO_3$ in the presence of strontium was also reported (42).

Knudsen and Roy (30) have experimentally shown that in the absence of Mg^{+2} the SiO₂/CuO ratio in the scale is an influencing parameter in the magnitude of the fouling resistance. It has been observed that when this ratio is about 3.0, the fouling factor is high while it decreases as the ratio becomes unity or less.

The presence of dissolved oxygen and chloride ion might also affect the fouling process since they both enhance the corrosion of some heat transfer surfaces (43, 44, 45).

2.8 Heat Transfer From Plain Surfaces

In a tubular heat exchanger, the transfer of heat between the two streams in the tube and in the annulus would take place due to: the conduction through the tube wall; the parallel mechanisms of convection and radiation between the streams and the inner and outer surfaces of the tube. Therefore, the total rate of heat transfer under the steady state condition in the absence of radiation, which is negligible, would be (46):

$$q = (q_c)_i = (q_c)_0 = q_k$$
 (2-70)

where q is the total rate of heat flow and subscripts c, k, i and o are denoted to convection, conduction, inner surface and outer surface, respectively. The rates of heat flow for three different modes of heat transmission are represented by the following equations:

$$q_{c_{i}} = \bar{h}_{c_{i}} A_{i} (T_{in} T_{s_{i}})$$
 (2-71)

$$q_{c_0} = \bar{h}_{c_0} A_0 (T_{s_0} - T_0)$$
 (2-72)

$$q_k = 2k_w \frac{\bar{A}}{D_o - D_i} (T_{s_i} - T_{s_o})$$
 (2-73)

where:

 \bar{h}_c = the average unit thermal conductance for convection k_w = the average unit thermal conductance for conduction D_i , D_o = the inside and outside tube diameters T_{in} , T_o = the temperatures of the inner and outer streams T_{s_i} , T_{s_o} = the inside and outside surface temperatures

 A_i , A_o = the inner and outer surface areas

$$\bar{A}$$
 = the logarithmic mean area = $\frac{A_0 - A_1}{\ln \frac{A_0}{A_1}}$

substituting equations 2-71 to 2-73 in equation 2-70 yields:

$$q = A_i \bar{h}_{c_i} (T_{in} - T_{s_i}) = A_o \bar{h}_{c_o} (T_{s_o} - T_o) = 2k_w \frac{A}{D_o - D_i} (T_{s_i} - T_{s_i})$$
 (2-74)

which can also be expressed by:

$$q = \frac{T_{in} - T_{o}}{\frac{1}{A_{o}\bar{h}_{c}} + \frac{D_{o} - D_{i}}{2k_{w}\bar{A}} + \frac{1}{A_{i}\bar{h}_{c}}}$$
(2-75)

The denominator of the above equation represents the overall resistance of the exchanger. The overall unit thermal resistance, $\frac{1}{U}$, of the exchanger based on the outside area is given by:

$$\frac{1}{U_0} = \frac{1}{h_{c_0}} + \frac{(D_0 - D_i)A_0}{2k_w A} + \frac{A_0}{h_{c_i}A_i}$$
(2-76)

thus,

$$q = U_0 A_0 (T_{in} - T_0)$$
 (2-77)

Since the streams and the wall temperatures are not constant along the path of the exchanger, equation 2-77 holds for an elemental area, dA_0 , and can more properly be represented by:

$$dq = U_0 dA_0 (T_{in} - T_0) = U_0 dA_0 \Delta T$$
 (2-78)

dividing both sides of the equation 2-78 by $d\Delta T$ would yield:

$$\frac{\mathrm{dq}}{\mathrm{d\Delta}T} = \frac{\mathrm{U}_{0}\mathrm{dA}_{0}\mathrm{\Delta}T}{\mathrm{d\Delta}T}$$
(2-79)

Now considering the heat balance equations in the absence of phase change:

$$dq = W_{in} c_{pin} dT_{in} = W_0 C_{po} dT_0$$
(2-80)

where W and c_p are the mass flow and the specific heat of the fluid, respectively. The above relationships represent that q is linear both in T_{in} and T_{o} ; therefore, its linearity in their difference is evident and can be given over the length of the exchanger, L, by:

$$\frac{dq}{d\Delta T} = \frac{q}{(\Delta T)_{x=L} - (\Delta T)_{x=0}}$$
(2-81)

substituting equation 2-81 in equation 2-79 yields:

$$\frac{q}{(\Delta T)_{x=L} - (\Delta T)_{x=0}} = \frac{\bigcup_{o} dA_{o} \Delta T}{d\Delta T}$$
(2-82)

For a constant overall unit thermal conductance, U_0 , the integration of the above-mentioned equation along the path of the exchanger yields:

$$q = U_0 A_0 \Delta T_m$$
 (2-83)

where ${}_{\Delta}T_{m}$ is the logarithmic mean temperature difference, LMTD, given by:

$$\Delta T_{m} = \frac{(\Delta T)_{x=L} - (\Delta T)_{x=0}}{(\Delta T)_{x=0}}$$
(2-84)
$$\ln \frac{(\Delta T)_{x=L}}{(\Delta T)_{x=0}}$$

The above-mentioned procedure can be carried out to find the overall unit thermal conductance based on the inner surface area:

$$q = U_i A_i \Delta T_m$$
 (2-85)

Equation 2-76 represents the overall unit thermal resistance of a clean tube. To calculate the overall unit thermal resistance of a fouled exchanger, the unsteady state fouling resistances are indiscriminately added to the steady state heat transfer resistances as follows:

$$(\frac{1}{U_{o}})_{d} = \frac{1}{h_{c_{o}}} + \frac{(D_{o} - D_{i})A_{o}}{2k_{w}A} + \frac{A_{o}}{h_{c_{o}}A_{i}} + {}^{R}f_{o} + {}^{R}f_{i}\frac{A_{o}}{A_{i}}$$
(2-86)

where R_f is the fouling factor and the subscript d is denoted to the fouled or dirty exchanger. Using the same analogy as in equation 2-70, the performance of a heat exchanger can be defined by (47):

1

$$q = \frac{1}{R_{f_0}} A_0 \Delta T_{f_0} = \frac{1}{R_{f_1}} A_i \Delta T_{f_1}$$
(2-87)

where ΔT_{f} is the temperature drop through the fouling film. Equation 2-87 shows that fouling factor, R_{f} , is not the only parameter controlling the fouling effect; ΔT_{f} and the heat flux, $\frac{q}{A}$, are equally important. Equation 2-76 implies that the overall unit heat flow conductance of an exchanger depends on: the geometry; the flow velocity; the specific heat; the thermal conductivity of both the surface material and the stream; and the viscosity. It increases with the first four while it is an inverse function of the last parameter. The rate of heat flow and, in turn, the fouling process are also affected by these factors. Table 2-2 shows how fouling factor is related to temperature and velocity (1). It increases with increasing the temperature and decreasing the velocity.

2.9 Heat Transfer From Extended Surfaces

To increase the rate of heat transfer per unit volume of the exchanger, extended surfaces as fins attached to the heat transfer wall have a wide industrial application. They compensate for the poor rate of heat flow in the fluid with lower heat transfer coefficient by exposing more surface to it. Therefore, the total rate of heat flow would be:

$$q_t = q_p + q_f \tag{2-88}$$

where t, P and f are subscripts denoted to total, prime and finned areas.

Knowing that any extended surface is less effective per unit area than a bare one due to the distance from the heat source, the fin efficiency (n_f) is defined (47, 48) by: Table 2-2. EFFECT OF TEMPERATURE AND VELOCITY ON FOULING FACTOR (47)

Temperature of heating medium	Up to 240°F		240-400°F†	
Temperature of water Water	125°F or less		Over 125°F	
	Water velocity, ft/sec		Water velocity, ft/sec	
	3 ft and less	Over 3 fi	3 ft and less	Over 3 fi
Sea water	0.0005	0.0005	0.001	0.001
Brackish water	0.002	0.001	0.003	0.002
Cooling tower and artificial spray pond:	•			
Treated make-up	0.001	0.001	0.002	0.002
Untreated	0.003	0.003	0.005	0.004
City or well water (such as Great Lakes)	0.001	0.001	0.002	0.002
Great Lakes	0.001	0.001	0.002	0.002
River water:				
Minimum	0.002	0.001	0.003	0.002
Mississippi	0.003	0.002	0.004	0.003
Delaware, Schuylkill	0.003	0.002	0.004	0.003
East River and New York Bay	0.003	0.002	0.004	0.003
Chicago sanitary canal	0.008	0.006	0.010	0.008
Muddy or silty	0.003	0.002	0.004	0.003
Hard (over 15 grains/gal)	0.003	0.003	0.005	0.005
Engine jacket	·0.001	0.001	0.001	0.001
Distilled	0.0005	0.0005	0.0005	0.0005
Treated boiler feedwater	0.001	0.0005	0.001	0.0Ò1
Boiler blowdown	0.002	0.002	0.002	0.002

FOULING FACTORS.

• Standards of Tubular Exchanger Manufacturers Association, 5th ed., New York, 1968.

.

† Ratings in the last two columns are based on a temperature of the heating medium of 240 to 400°F. If the heating medium temperature is over 400°F, and the cooling medium is known to scale, these ratings should be modified accordingly.

.

$$n_{f} = \frac{actual heat transferred by fin}{heat which would have been} = \frac{q}{max}$$
(2-89)
transferred if entire fin were
at the base temperature

where

$$q_{max} = h_c A_f (T_s - T_b)$$
(2-90)

 T_s = prime surface temperature or the temperature at the base of the fin

$$T_b = bulk temperature$$

Substituting equation 2-89 and 2-90 in the equation 2-88 and applying the same analogy used for the finned area, equation 2-89, to the total area yields:

$${}^{n}t\bar{h}c_{t}^{A}t^{(T}s^{-T}b) = \bar{h}c_{p}^{(A}t^{-A}f^{)}(Ts^{-T}b) + {}^{n}f\bar{h}c_{f}^{A}f^{(T}s^{-T}b)$$
(2-91)

assuming that ${\rm \tilde{h}}_{\rm C}$ is invariant over the total surface would simplify the equation 2-91 to:

$${}^{\eta} t^{A} t = {}^{(A} t^{-A} f) + {}^{\eta} f^{A} f$$
 (2-92)

where $n_t A_t$ is the effective area (A_{eff}).

Fin efficiency can be estimated making a heat balance for a small element of the fin under steady state condition

with regard to Figure 2-4. The above-mentioned equation becomes:

$$q_x = q_{x+dx} + 2h_c(b+t)dx(T-T_b)$$
 (2-93)

or for uniform k_{fin}:

$$-k_{fin}bt \frac{dT}{dx} = -k_{fin}bt \frac{d}{dx} (T + \frac{dT}{dx}dx) + 2\bar{h}_{c}(b+t)dx(T - T_{b}) \qquad (2-94)$$

$$\frac{2}{dx^2} = \frac{2h_c(b+t)}{k_{fin}bt} (T-T_b)$$
(2-95)



Figure 2-4. SKETCH AND NOMENCLATURE OF A RECTANGULAR FIN

The above second order differential equation can be solved based on the following assumptions:

- Steady state and no heat generation by internal sources.
- Transverse temperature gradient is very small that T=T(x).
- Uniform thermal conductivity for both fin and the tube.
- Uniform convective heat transfer coefficient, \overline{h}_{c} , between surface and fluid.
- Uniform fluid temperature.
- Heat loss from the end of the fin is negligible.

Therefore, the actual heat transferred by fin would be given by:

$$q_{f} = -k_{fin}A_{f} \frac{dT}{dX} = k_{fin}A_{f}m(T_{s}-T_{b})tanh(m1) \qquad (2-96)$$

where

 A_f = fin cross section = bt

P = perimeter = 2(b+t)

$$m = (\bar{h}_{c}P/k_{fin}A_{f})^{0.5}$$
 (2-97)

Fin efficiency can be estimated by substituting equations 2-90 and 2-96 in equation 2-89:

$$n_{f} = \frac{Tanh(m1)}{m1}$$
(2-98)

Knowing that (m1) grows more rapidly than Tanh (m1) with increasing (m1), it can be seen that nf and therefore nt are reciprocally related to \bar{h}_c . Thus, fins are more effective when they are placed on the stream side having a lower convective heat transfer coefficient, \bar{h}_c .

As in equation 2-76, the overall unit thermal resistance of a clean externally finned tubular exchanger, based on the total outside area, can be expressed by:

$$\left(\frac{1}{U_{o}}\right) = \frac{1}{n_{t_{o}}h_{c_{o}}} + \frac{A_{t_{o}}(D_{o}-D_{i})}{2k_{w}A} + \frac{A_{t_{o}}}{h_{c_{i}}A_{i}}$$
(2-99)

,

,

Adding the fouling resistances results in the overall unit thermal resistance of a dirty exchanger as follows:

$$\left(\frac{1}{U_{o}}\right)_{o} = \frac{1}{\pi_{t_{o}}\bar{h}_{c_{o}}} + \frac{A_{t_{o}}(D_{o}-D_{i})}{2k_{w}\bar{A}} + \frac{A_{t_{o}}}{\bar{h}_{c_{i}}A_{i}} + \frac{A_{t_{o}}}{\pi_{t_{o}}} + \frac{R_{f}}{\pi_{t_{o}}} + \frac{A_{t_{o}}}{A_{i}}$$
(2-100)

When the outer surface is finned, particular consideration should be given to the fouling resistance of the inside of the tube. When the inside fouling resistance, R_{f_i} is very large, there would be no use in adding external fins and increasing the outer surface which would, in turn, increase the total fouling resistance.

2.10 Comparison Between an Enhanced and a Plain Heat Transfer Surface

In a tubular heat exchanger, fins can be attached either to the inside or to the outside surface area of the tube. There are different fin designs which are generally classified into two major categories: longitudinal and radial. Both the existence and the design of the fins affect the hydrodynamic flow condition and might, in turn, influence the fouling process which is a function of the dynamics of the fluid.

In a turbulent annular flow, the presence of fins would result in the formation of eddies and subsequent disturbance of the laminar sublayer as was shown by Knudsen and Katz (49). These investigators showed that the number of eddies and therefore the degree of turbulence increases with increasing the fin height, 1, to the fin spacing, S_f , ratio. The study of Knudsen and McCluer (50) on transverse finned tubes confirmed the results obtained by the previous investigators. They showed that, due to the existence of the transverse fins, the rate of calcium sulfate scaling decreases with increasing values of the $1/S_f$ ratio, resulting in a high turbulence. This gave rise to the notion that the fouling factors are less for finned tubes than for the plain tubes. The experimental data reported both by Katz et al. (51) and by Webber (52) also indicated a less pronounced fouling process on the finned tubes than on the plain tubes. However, Knudsen and Roy (30) studied the CaCO₃ scaling and reported the same fouling factor, based on the projected outside area, for both smooth and finned tubes.

The effect of fins on the asymptotic scaling of $CaCO_3$ under the constant steam temperature was studied by Watkinson et al. (53). Plain, inner fin and spirally indented heat exchanger tubes were tested. The experimental results showed substantial advantage in heat transfer coefficient of the enhanced tube over that of the plain tube after fouling. However, the asymptotic fouling resistance was higher for the inner-fin tube than for the plain tube in the velocity region (2 < v < 6 ft/sec) tested. For the spirally indented tubes at velocities below 3 ft/sec, the asymptotic fouling resistance was above that for the plain tube while it became less than that of the plain tube at velocities above 3 ft/sec.

To investigate the performance of an enhanced heat transfer surface, it is important to know the effect of fouling on the total efficiency of the surface, n_t . Epstein and Sandhu (54) presented two mathematical models

to predict the effect of uniform fouling on the total efficiency of extended heat transfer surfaces. In the first model, they treated the fouling deposit on the fin as a thermal resistance in series with the adjacent fluid, as it is normally considered to be. In the second one, the dirt on the extended surface was treated as a thermal conductor in parallel with the fin. Dirty fin efficiency, η_{fd} , was expressed in terms of parameter m_d for a dirty surface in the same manner shown in equation 2-98 for a clean surface. The models suggested that in both cases, the dirty fin efficiency would decrease due to an increase in m_d . Since m_d in either case is proportional to m of a clean surface (equation 2-97), any decrease in m would result in an enhancement of the fouled fin efficiency. In the series model, a decrease in $\mathbf{m}_{\mathbf{d}}$ and a subsequent increase in the dirty fin efficiency is accompanied by a reduced heat transfer due to the insulating effect of the dirt and a lower effective convective heat transfer coefficient. However, heat transfer and dirty fin efficiency are simultaneously increased due to a higher effective conductivity in the parallel model. The ratio of the total efficiency of the clean enhanced surface to that of the dirty one, n_t/n_{td} , was chosen as a measure to see the actual effect of fouling on Holding all the other parameters constant (for an extended surface. $0 < A_f/A_+ < 1$), this ratio increases with increasing m1 for the parallel model while it goes through a minimum of ml²2 for the series model. In both cases at $A_f/A_t=0$, n_t/n_{td} is constant and has a higher value than those with $A_f/A_t>0$. Therefore, both mathematical models show that an increase in thermal resistance caused by a uniform fouling deposit on an enhanced heat transfer surface is less pronounced than that on a plain surface having the same dirt thickness. With the other parameters constant,

at higher dirty fin efficiency, n_{fd} , due to the lower m_d the resistance factor, n_t/n_{td} , of an exchanger is enhanced according to the series model while reduced with regard to the parallel model. However, the negative effect of the dirt on the prime surface in series with the fluid in contact would reduce the performance of the exchanger by increasing n_t/n_{td} , in the parallel model. The models predict similar results for the n_t/n_{td} ratio at lower values of A_f/A_t , higher values of ml (lower fin efficiency) and lower values of h/h_d . The series model gives rise to more conservative results. It provides a higher value for n_t/n_{td} ratio than the parallel model even under the above-mentioned conditions.

3. EXPERIMENTAL APPARATUS

3.1 Water Flow Loop

A flow diagram of the apparatus is shown in Figure 3-1. The artificially hardened water leaves the supply tank via a 2 inch pipe and is pumped around the closed loop using a Cole-Parmer model K-7084-20 centrifugal pump. The pump is driven by a totally enclosed fan cooled (TEFC) 3/4 HP motor.

The fluid enters and exits the pump via 1-1/2 in. copper pipes. The flowrate of the recirculating water is controlled manually and measured by a Brooks Model R-12M-25-4 rotameter (see Appendix I for the calibration curve). The test fluid flows upwardly into the shell side of the heat exchanger which consists of a 1.15 meter long glass tube with an inside diameter of 37 mm. The shell is flanged from both ends to brass headers and sealed with rubber gaskets.

Hard water is heated by the steam which passes downwards through the central tube. The test fluid leaving the shell side of the heat exchanger is cooled down by passing through a finned helical heat exchanger. Cooling water is supplied via the building mains and controlled by a R-9M-25-2 Brooks rotameter (see Appendix I for the calibration curve).


Figure 3-1. FLOW DIAGRAM OF APPARATUS

- 60 -

Recirculating water which exits the finned helical heat exchanger can either enter the supply tank directly or first pass through an in-line filter. This filter is equipped with an AMF Cuno Mikro-Klean 50 micron cartridge which removes the particulates without affecting the water chemistry (Figure 3-2).

3.2 Supply Tank

Water is recirculated from a 220-liter stainless steel tank. Two holes are drilled in the bottom and on the side of the tank and tapped to bulk head fittings with 3/4 in. NPT (National Pipe Tapered) and 2 in. NPT, respectively. The 3/4 in. NPT hole is connected to a 3/4 in. steel pipe equipped with a valve which is used as a drain. The 2 in. NPT hole is connected to a 2 in. steel pipe which can either direct the test fluid via a 1-1/2 in. copper pipe to the pump or to the drain via a valve. Water samples are taken using this outlet. The tank is equipped with a steam heating coil to heat the test fluid up to the initial operating temperature. It is also covered with a piece of wood having an opening for introducing the chemicals.

3.3 Steam System

The steam is supplied by the main steam line at a regulated pressure measured by a 100 psig Marsh steam pressure gauge. This regulated steam is introduced either to the steam heating coil in the tank or to the tube side of the shell and tube heat exchanger via a 1/2 in. steel pipe



Figure 3-2. PHOTOGRAPH OF APPARATUS FROM THE BACK

(Figure 3-2). In either case, the line pressure is controlled by a valve and measured by a 30 psi Marsh steam pressure gauge. The condensate produced by a downward passage of the steam through the test exchanger is removed by a Clark No. 60 steam trap located at the base of the heat exchanger. The steam and generated condensate from the coil are directly discarded into the drain.

3.4 Tube Material and Geometry

On the basis of geometry, the tubes used can be divided into two groups: finned and plain. The 1.5 meter mild steel finned tube having twelve longitudinal fins welded to its outside surface area, is supplied by Bas-Tex Corporation (Figure 3-3). Figure 3-4a and Figure 3-4b show horizontal and vertical cross-sections of a finned tube. The three different 1.5 meter long plain tubes used can be categorized into two groups: first, the non-corroding stainless steel and copper tubes, and second, the corroding mild steel tube. The properties and geometries of the tubes are summarized and listed in Table 3-1.

Both in case of the plain tubes and the finned tube, the length of the tube in contact with the working fluid was 1.33 meters. In either situation, the nominal area - outside diameter (0.0796 m^2) is used to express the heat transfer performance. Use of the nominal area allows a direct measure and comparison of the results if the finned tube replaces the plain one.



Figure 3-3. SKETCH OF A TWELVE LONGITUDINALLY FINNED TUBE

.

.





Figure 3-4. HORIZONTAL (a) AND VERTICAL (b) CROSS SECTIONS OF A FINNED TUBE

Table 3-1. SUMMARY OF THE PROPERTIES OF TUBES

Tube No.	Material	0.D.	I.D.		Propertio	es of Fin	De	Net Free Area			
		(mm)	(mm)	No.	Length (m)	Height (mm)	Width (mm)	(m)	x 10 ⁴ (m ²)		
1	S.S.	19.1	15.9	-	-	- ·	-	0.01791	7.9128		
2	Сор	19.1	15.3	_	-			0.01791	7.9128		
3	M.S.	19.1	13.3	-	-	-		0.01791	7.9128		
4	M.S.	19.1	14.7	12	1.2	6	0.5	0.00869	6.9528		

S.S. = stainless steel Cop = copper M.S. = mild steel

ì

3.5 Temperature Measurements

Hard water and steam temperatures are measured using iron-constantan thermocouples. The temperature of the hot stream is measured at the inlet and

outlet of the tube. Other thermocouples to measure the temperature of the test fluid are located at the feed tank exit, at the inlet and outlet of the shell and at the exit of the helical heat exchanger.

The thermocouples are connected to a switch, which is in turn connected to the Fluke Model 2240A datalogger (Figure 3-5). The datalogger measures the voltage with the thermocouples output referenced to 0° C by placing their cold junctions in an ice bath (Figure 3-6). Thermocouples were calibrated using a Hewlett Packard model 2801A guartz thermometer.



Figure 3-5. CIRCUIT DIAGRAM FOR DATALOGGER CONNECTION

- 88 -

<u>,</u> .



Figure 3-6. PHOTOGRAPH OF APPARATUS FROM THE FRONT

4. EXPERIMENTAL PROCEDURES

4.1 General Approach

Artificially hardened water was prepared by adding $Ca(NO_3)_2 \cdot 4H_2O$ or $CaCl_2 \cdot 2H_2O$ and $NaHCO_3$ to 210 liters of water. To avoid corrosion on mild steel by chloride ion, calcium nitrate was used as a source of calcium. Nitrogen gas was introduced into the water to remove the dissolved oxygen in the supply tank. When sufficient nitrogen had bubbled through the water, the nitrogen gas flow rate was measurably reduced and diverted only to maintain a nitrogen atmosphere above the liquid in the tank, to prevent further introduction of oxygen into the test fluid during the run.

It is important to have both steady temperatures and heat flow. These were to be achieved by keeping the temperature of water entering and leaving the exchanger constant. Where the inlet temperature could not be held constant due to limits on the cooling water temperature and flow rate, the temperature rise of the water in the exchanger was nevertheless kept as constant as possible.

After being assured, by taking several temperature readings, that the steady state condition was reached, $CaCl_2 \cdot 2H_20$ or $Ca(NO_3)_2 \cdot 4H_20$ was added. Two or three minutes later, $NaHCO_3$ was added. The first water sample was taken, allowing time for complete mixing to occur in the tank. The sample was filtered using a Gelman magnetic filter funnel equipped with a Whatman GF/B glass microfiber filter which removes particles down to 1 micron in size.

Several water samples were taken during each run, approximately 8-12 hours apart, and more chemicals were added to keep the water quality constant. It should be mentioned that the addition of chemicals was done during the run with higher concentration of chemicals since the water quality did not vary to a great extent during runs at lower concentration. The in-line filter in the recycle line was used to remove excessive amounts of suspended solids. However, for runs at lower concentration of chemicals, the direct recycle line was used because the suspended solids concentrations were negligible.

4.2 Solution Preparation

For the low concentration early runs, 97.35 gr of $CaCl_2 \cdot 2H_2O$ and 64.58 gr of NaHCO₃ were used to provide an artificially hardened water with the initial concentration of 0.126 g of Ca^{++} and 0.084 g/l of Na⁺. In later runs, $Ca(NO_3)_2 \cdot 4H_2O$ was used as a source of calcium to reduce corrosion arising from the chloride ion.

Since the deposit obtained using the above-mentioned concentrations of chemicals was muddy, the amount of substances was increased by a factor of 2.2 (0.00694 mole/liter of Ca⁺⁺ and 0.00805 mole/liter of Na⁺) in order to get a harder and better constructed scale. In either case, the chemicals were separately dissolved in the water before addition to the supply tank.

4.3 pH Measurements

pH is measured using an Orion Research Digital Ionalyzer/501. pH 4.01 \pm 0.01 (at 25°C) and pH 8.00 \pm 0.02 (at 25°C) buffer solutions are used to calibrate the pH meter.

4.4 Alkalinity Measurements

Total alkalinity (T.A.) is measured by titration using a bromocresol green-methyl red mixed indicator which is more suitable for alkalinities below 500 mg/l. $Na_2S_2O_3$ is used as an inhibitor for the removal of residual chlorine that would otherwise impair the indicator colour changes. 0.2 N HCl is used as a titrant and the end point is both determined by the indicator color change and by a pH meter (55).

1 drop of $Na_2S_2O_3$ and 3 drops of the mixed indicator are added to 50 ml of sample. The blue sample is titrated with 0.02 N HCl to the appearance of a light pink which occurs at pH 4.5. Total alkalinity is then calculated using the following formula:

Total Alkalinity =
$$\frac{B' \times N \times 50000}{V'}$$

- where B' = volume of the titrant (ml)
 N = normality of titrant
 - v' = volume of the sample (ml)

4.5 Hardness Measurements

Hardness is determined by a complexometery method using Erichrome Black T as an indicator. The presence of magnesium is required for satisfactory determination of the end point in the titration. Both the sharpness of the end point and the tendency of $CaCO_3$ and $Mg(OH)_2$ to precipitate increase with increasing pH; therefore, the pH value of $10.0 \pm$ 0.01 is recommended as a satisfactory compromise. A buffer solution is prepared by mixing 55 ml concentrated HCl with 400 ml distilled water, adding 310 ml 2-amino ethanol and 5 g magnesium salt of EDTA and diluting the whole mixture to 1 litre with distilled water (56).

To prevent the interfering effect of some metal ions which cause an indistinct end point, a mixture of 4.5 gr hydroxylamine hydrochloride in 100 ml of 95 percent ethyl-alcohol is used as an inhibitor. A solution of 0.01 M disodium ethylenediamine tetraacetate dihydrate (EDTA) is used as a titrant.

To carry out the titration, 1 ml of the buffer, 1 ml of the inhibitor and 2 drops of the indicator are added to 25 ml of sample which is already diluted to 50 ml. 0.01 M solution of EDTA is added to the reddish sample until the disappearance of the last reddish tinge. Duration of titration should not exceed 5 min in order to prevent precipitation of CaCO₃ at higher pH values. Using the following formula, the hardness (C_a^{++}) can be determined:

$$Ca^{++} = \frac{A' \times B'' \times 1000}{v'}$$

where

A' = volume of the titrant (ml)
B'' = mg of CaCO₃ equivalent to 1 ml of EDTA
v' = volume of the sample (ml)

4.6 Determination of Total Dissolved Solids

Since there is not a high quantity of dissolved solids in the tap water, it can be treated as distilled water regarding calculation of total dissolved solids (TDS). If solutions of x mole/liter of NaHCO₃ and y mole/liter of Ca(NO₃)₂ were brought together, the following basic reactions would take place (assuming $\frac{x}{2} < y < x$ which is applicable in this experiment:

$$x \text{ NaHCO}_{3} = x\text{HC}\overline{O}_{3} + x\text{Na}^{+}$$

$$y\text{Ca}(\text{NO}_{3})_{2} = y\text{Ca}^{+} + 2y\text{N}\overline{O}_{3}$$

$$x\text{HC}\overline{O}_{3} = \frac{x}{2} \text{CO}_{2} + \frac{x}{2} \text{C}\overline{O}_{3}^{-} + \frac{x}{2} \text{H}_{2}\text{O}$$

$$\frac{x}{2}\text{C}\overline{O}_{3}^{-} + y\text{Ca}^{+} - \frac{x}{2} \text{Ca}\text{CO}_{3} + (y-\frac{x}{2}) \text{Ca}^{+}$$

$$2y\text{N}\overline{O}_{3}^{-} + x\text{N}_{a}^{+} = x\text{NaNO}_{3} + (2y-x) \text{N}\overline{O}_{3}^{-}$$

$$(2y-x) \text{N}\overline{O}_{3}^{-} + (y-\frac{x}{2}) \text{Ca}^{+} = (y-\frac{x}{2}) \text{Ca}(\text{NO}_{3})_{2}$$

Therefore, summation of the above chemical reactions yields:

$$x NaHCO_3 + yCa(NO_3)_2 \longrightarrow \frac{x}{2} CaCO_3 + x NaNO_3 + (y-\frac{x}{2})Ca(NO_3)_2 + \frac{x}{2} CO_2 + \frac{x}{2} H_2O_3$$

Thus, following the assumption that the tap water is treated as distilled water:

Total Alkalinity (mole/liter) =
$$\frac{x}{2}$$
 CaCO₃

Hardness (mole/liter) =
$$(y-\frac{x}{2})$$
 Ca $(NO_3)_2 + \frac{x}{2}$ CaCO₃
TDS (mole/liter) = $\frac{x}{2}$ CaCO₃ + x NaNO₃ + $(y-\frac{x}{2})$ Ca $(NO_3)_2$

Therefore, all the above-mentioned quantities can be rewritten in terms of $mg/liter of CaCO_3$:

T.A. =
$$10^5 \frac{x}{2}$$

 $C_a^{++} = 10^5 y$
 $C_a^{++} - T.A. = 10^5 (y - \frac{x}{2})$
TDS = $10^5 (x+y)$

.

Also, TDS can be evaluated in terms of total weight of solids per volume (mg/l):

TDS =
$$(10^5 \frac{x}{2}) + (85000x) + (164) \left(\frac{Ca^{++} - T.A.}{100}\right)$$

TDS = T.A. + 1.7 T.A. + 1.64 $(Ca^{++} - T.A.) = 1.06$ T.A. + 1.64 Ca^{++}

4.7 Procedure For a Scaling Run

- (1) Install one of the clean test sections in the tube and shell heat exchanger. Connect it, from the top, to the steam line and, from the bottom, to the steam trap. Insulate the test heat exchanger.
- (2) Fill the supply tank to 210 liter line with the tap water. Connect the tap water hose to the cooling water line.
- (3) Put the cold junctions of the thermocouples in the thermos bottle already filled with the ice/water mixture.
- (4) Turn on the datalogger. Set the time and program the format of a scan sequence. Program the datalogger to give the output in millivolts.
- (5) Turn on the pump and set the recirculating water flow rate at a desired value.
- (6) Introduce the nitrogen gas by turning the control valve open.
- (7) Follow the solution preparation procedure.
- (8) Reduce the nitrogen gas flow and divert it from the water to the top of the tank.

- (9) Turn on the steam into the heating coil located inside the supply tank.
- (10) Cut off the steam to the tank after reaching the desired operating temperature (25°C for most of the runs).
- (11) Turn the steam on the exchanger at 115.15 kPa absolute pressure (used for almost all of the runs).
- (12) Set the cooling water flow rate.
- (13) Calculate the heat flux by measuring the inlet and outlet temperatures of hot and cold streams passing through the test heat exchanger.
- (14) After taking several readings and having the steady heat flow rates, add $CaCl_2 \cdot 2H_2O$ or $Ca(NO_3)_2 \cdot 4H_2O$ (as required) solution. Wait for 2-3 minutes until the chemical is completely mixed, then add NaHCO₃ solution and note the time. The chemicals are added using a plastic funnel passing through the hole in the center of the wooden lid of the tank.
- (15) When the mixing process is judged to be complete (approximately after 5 minutes), record both the temperature measurements and steam pressure and take the water sample using the side outlet of the supply tank. Register the time.

- (16) Take the temperature readings every 10 minutes during the first hour and every 30 minutes for the rest of the run. Record the steam pressure as often as possible.
- (17) Try to keep the exchanger at a steady heat flow by controlling both the steam pressure (which rises as the tube scales) and cooling water flow rate.
- (18) Since the recirculating water flow rate is not automatically controlled, control it manually to have a steady water flow rate.
- (19) Take a water sample every 8-12 hours and determine the water quality following the pH, total alkalinity and hardness measurements after filtration of the water sample through a Whatman GF/B glass microfibre filter.
- (20) When there is a significant change in total alkalinity and hardness during the two successive samplings, which usually occurred for the runs with higher concentrations of chemicals, addition of more chemicals is made to keep the water quality essentially constant. Assuming that the deposit is made purely out of CaCO₂, the change in both total alkalinity and

ţ

hardness should be exactly the same (see determination of TDS). Considering the above assumption, the addition process is carried out according to the following relationships:

weight of Ca(NO₃)₂ .
$$4H_20$$
 added gr = $\frac{210 \ Z \ M_{Ca(NO_3)_2} \cdot 4H_20}{10^5}$

weight of NaHCO₃ added gas (gr) =
$$\frac{420 \text{ Z } \text{M}_{\text{NaHCO}_3}}{10^5}$$

Ł

- (21) Take water samples after addition of make up chemicals and re-evaluate water quality.
- (22) Calculate the heat flux, the overall heat transfer coefficient and the fouling resistance for each reading.
- (23) After a period of 70 hrs, where there is not a big change in the fouling resistance for several hours, stop the run.
- (24) Cut off the steam to the tube. Turn off the valve to the main steam line and purge the line by turning on the steam valve to the steam heating coil.

- (25) Turn off the pump, tap water and datalogger. Drain the supply tank through both bottom and side exits. Let the system cool down. Remove the tube from the heat exchanger. Examine the tube's deposit for hardness. Measure the thickness of the deposit at several places throughout the length of the tube. Scrape or crack off the scale and collect it for comparison with the results of other runs.
- (26) Measure the outside diameter of the bare tube to see if there has been any change due to corrosion.

4.8 Cleaning

The tubes were used several times; therefore, their outside surface areas required cleaning after each run. Initially, after removing the tube from the exchanger and scraping off the heavier deposit, the tube was replaced in the exchanger and the water containing about 2400 ppm of Oakite 31 (containing phosphoric acid and biodegradable surfactants) was pumped around the system for a couple of hours to remove $CaCO_3$ deposits both on the tube and in the rest of the equipment.

Since the solubility of $CaCO_3$ in oakite solution increases with increasing temperature, the solution was heated up by turning on the steam to the heating coil located in the supply tank. Then the pump was stopped and the cleaning solution was drained out of the system. To remove any oakite from the equipment, pure water was pumped throughout the system for

30 minutes and then drained out via the supply tank. This action was repeated 3 to 4 times to clean the tube and the rest of the equipment from any residual oakite.

This cleaning procedure was not satisfactory for two reasons. Firstly, it was not possible to clean the tubes well. Secondly, the procedure enhanced the corrosion of mild steel tubes. One alternative was to clean the tubes by hand. First, the majority of scale was removed by a mild steel wire brush, then remaining deposits were removed by steel wool and emery cloth. However, this manual cleaning was neither time nor labor efficient and, most importantly, failed to remove all the scale.

A second alternative was to remove the deposits by a 20% solution of inhibited hydrochloric acid. Pickling inhibitor, Rodene, was added at concentrations of 0.1% by volume to prevent the corrosive effect of the acid. The tubes were chemically cleaned under the batch condition using a 120 cm high glass column having an inside diameter of 5 cm. The cleaning procedure was completed using an emery cloth. However, a mixture of oakite solution was pumped around the system to clean the rest of the equipment. In this case, two metal rods having the same outside diameter as the tubes were used to block the inlet and outlet of the hot stream in the exchanger. As mentioned above, the residual oakite was removed by recirculating pure water.

5. RESULTS AND DISCUSSION

5.1 General Outlook

5.1.1 Low Concentration Runs

Tests were first done on the stainless steel and copper tubes with initial concentration of $Ca^{++} = 0.126$ g/l and $Na^+ = 0.084$ g/l, using $CaCl_2$. $2H_2O$ and $NaHCO_3$ as sources of calcium and carbonate, respectively. These are termed "low concentration runs". As the initial operating condition, the steam was conducted to the tube at 115.15 kPa absolute pressure, after heating up the tank water temperature to $25^{\circ}C$ (see Section 4.7). The entrance and exit temperatures of both the hot and the cold streams were recorded and used for calculation of heat flow, q, and overall heat transfer coefficient, U (see Appendix II). Having obtained the steady state condition with water only, the chemicals were then added. This was followed by subsequent measurement of water quality and calculation of q and U.

As time passed and the tube scaled, the overall heat transfer coefficient dropped and the steam pressure was raised to supply a constant heat flow. The magnitude of the fouling resistance, R_f , was determined by intermittent measurement of U and hence the change in the total resistance of the exchanger with respect to time. The effect of scaling on the water quality was followed using alkalinity, hardness and pH measurements. Since the in-line filter was not used for these runs, the suspended solids content of the water sample was determined using the Whatman GF/B glass microfiber filter to investigate the extent of the bulk precipitation process.

Figure 5-1, corresponding to the copper tube at annular flow velocity of 0.503 m/s, shows a typical output for a run at low concentration of chemicals. At constant heat flux value of 14.1 kW, the overall heat transfer coefficient decreased with the time, reaching 2.09 kW/m²K after 70 hrs., which corresponded to an 18.4% drop in U from the clean condition $(2.56 \text{ kW/m}^2\text{K})$.

As is evident from this figure, the fouling process was accompanied by a drop both in total alkalinity and the hardness. The reductions in the magnitude of these parameters for the subsequent water samples were not high enough to require the addition of more chemicals. After 70 hrs., the changes in the magnitude of total alkalinity and hardness were respectively 61 and 63 mg $CaCO_3/1$. Considering the experimental errors, these values were close enough to deduce that the deposits were essentially pure $CaCO_3$ (see Section 4.6).

The suspended solids content of the water first increased with time and reached a maximum of 3.6 ppm after 24 hours. This suggests that some precipitation occurred both in the bulk and in the hot film adjacent to the interface. The slight reduction in the suspended solids content in the latter part of the run indicates the possible existence of a small extent of particulate fouling.



Figure 5-1. RUN 5 TYPICAL OUTPUT

84 -

Each tube was studied under three different annular flow velocities ranging from 0.29 m/s to 0.7 m/s. The results are listed in Table 5-1. Since heat flow increases with annular mass flow rate, both the heat flow rate and the overall heat transfer coefficient of a clean tube increased with increasing velocity, as is shown in the table. Comparing the non-corroding tubes under approximately identical operating conditions, the stainless steel tube has both lower heat flux and lower clean overall heat transfer coefficient than does the copper tube. This was understandable considering the lower value of unit thermal conductance for the stainless steel than for the copper. Regarding the dirty heat transfer coefficient. the advantage of the copper tube over the stainless steel one was also evident through the experimental results. The only irregularity occurred for the stainless steel tube at annular velocity of 0.503 m/s (Run 3) in which there was no change either in the magnitudes of water quality parameters or in the overall heat transfer coefficient with the passage of time due to an unknown reason.

For either tube, the scale thickness on the dirty surface decreased with distance from the steam entrance. The deposits could not be cracked off. Hence, they were taken out partly by scraping off with a spatula, resulting in a rusty coloured powdery residue. The remainder was removed by chemical cleaning.

5.1.2 High Concentration Runs

Tests were carried out on both the plain and the finned mild steel tubes under the above-mentioned operating conditions using low concentrations

v (m/s)	Stainless Steel											Copper												
	Run		T.A.		Ca++*		PHave	U (kW/m ² K)			Run	- <u></u> 0	T.A.*		Ca ⁺⁺ *		DHave	U (kW/m ² K)						
		(k₩)	Start	End	Start	End		Start	End	% Drop	(k	(kW)	Start	End	Start	End		Start	End	% Drop				
0.200	1A	7.0	188	168	205	285	7.84	1.37	1.16	15.3	6A	11.5	177	129	298	250	7.61 7.60	2.25	1.74	22.7				
0.299	1.	7.9		148	305	270	7.86		1.16	15.3	6		177	105		228			1.69	24.9				
	3A			175		302	7.89	1.68	1.68	0.0	5A	14.1	176	140		260	8.11	2,56	2.17	15.2				
0.503	3	10.4	182	175	310	302	7.93		1.68	0.0	5		176	115	298	235	8.11		2.09	18.4				
0.605	2A 2	11.6	182	158	202	280	7.92	1.83	9.9	4A			151	204	274	8.03		2.73	17.0					
0,095		11.5		125	302	247	7.95	2.03	1.73	14.8		18,0	170		294			3.29	2.55**	.55** 22.5**				

Table 5-1. SUMMARY OF RESULTS FOR RUNS AT LOW CONCENTRATIONS OF CHEMICALS

T.A. = total alkalinity Ca⁺⁺ = hardness * Units in mg CaCO₃/1 ** Values determined by extrapolation A = runs lasted 48 hours

I 98

Ŧ.

of chemicals. The plots of the overall heat transfer coefficient versus time showed erratic behaviour especially at higher flow rates (above 2.33 x 10^{-4} m³/s). The high velocity runs resulted in a drop of the overall heat transfer coefficient during approximately the first ten hours of the experiment followed by a subsequent rise in the magnitude of the coefficient even though there was a continuous reduction in the values of the total alkalinity and the hardness throughout the run. A typical example for such a run is shown in Figure 5-2.

The corroding effect of the mild steel tube was considered to be the major factor responsible for this behaviour especially in the case of the plain tube. Therefore, nitrogen gas was introduced to blanket the liquid in the tank and $Ca(NO_3)_2$. $4H_2O$ replaced $CaCl_2$. $2H_2O$ as the source of calcium to eliminate the corrosive effect of dissolved oxygen and chloride ion, respectively. The rise in the value of overall heat transfer coefficient could be due to the low concentration of chemicals which might prevent the complete adherence of the scale to the mild steel surface and enhance the probability of removal mechanism at high velocities or to roughness effects which would be of considerable importance with thin scales. Thus, the concentration of chemicals were raised by a factor of 2.2 to get a more pronounced result since the main objective was the determination of the fouling resistance and its variation with the time. These runs are designated "high concentration runs". With the increase in amount of scale deposited, the changes in the magnitude of the total alkalinity and hardness for two subsequent water samples taken during the run were high enough to require the intermittent addition of make up chemicals to keep the water quality as constant as possible (see Section 4.7). In



Figure 5-2. RUN 16 TYPICAL OUTPUT

- 88 -

addition, the in-line filter was used to eliminate the interfering effect of the suspended solids in the bulk. The initial absolute steam pressure was 115.15 kPa for all high concentration runs except for Runs 21 and 22 (P=129 kPa). The initial tank water temperature was also uniform for all runs except for Run 22 (T=30°C).

A typical output for a high concentration run is shown in Figure 5-3. At 0.299 m/s annular velocity, as the tube fouled, the overall heat transfer coefficient of the plain mild steel tube decreased approximately from 1.45 kW/m²K to 1.25 kW/m²K (13.8%) after 70 hours. Figure 5-3 also represents a rapid decrease in the magnitudes of total alkalinity and hardness during the first four hours (about 20 mg/l of CaCO₃ per hour) followed by a slower one (about 4.5 mg/l of CaCO₃ per hour).

Five different annular flow velocities, ranging from 0.29 m/s to 0.8 m/s $(2.33 \times 10^{-4} \text{ to } 5.50 \times 10^{-4} \text{ m}^3/\text{s})$, were used to investigate the effect of velocity on the fouling process of the mild steel tubes. The copper tube was also tested under high concentration of chemicals at three different annular water flow rates in the above-mentioned region. Table 5-2 summarizes the results of the high concentration runs. The data in the table indicate an increase in the overall heat transfer coefficient with the velocity for either tube. Under identical operating conditions, the rate of heat transfer and the clean and the dirty heat transfer coefficients were higher for the copper tube than the plain mild steel one. In the case of mild steel tubes, the finned tube had a higher heat flow rate than the plain one. The reasons for the advantage of the copper and the finned mild steel



Time (hrs)

Figure 5-3. RUN 24 TYPICAL OUTPUT

- 90 -

Vol. Flow Rate x 105 (m3/s)		Finned Hild Steel							Plain Nild Steel									Copper								
	Run	9 (KW)	T # # \	Ca ⁺⁺ *	0H	U _O (kW/m ² K)		Rup		T.A.+	Ca+++		U ₀ (kW/m ² K)		Run		T A +			U _G (kW/m ² K)						
			ave	8.46	448	Start	End	\$ Drop		(xW)	476	ave	ave	Start	End	1 Drop		(¥¥)	ave	446	Ave	Start	End	\$ Drop		
23.67	27	9.8	371	661	7.64	1.61	1.54	4.3	24	8.4	415	697	7.78	1.45	1.25	13.8	20									
	30	10.1	365	650	7.54	1.69	1.45	14.2									96	9.9	191	923	7.69	1.70	1.48	12.9		
32.00	28	11,3	347	633	7.60	2.00	1.78	11.0	29	9.1	351	639	778	1.58	1.32	16.5								·		
39.83		19.7	361	640	3 65			2 02		23	11.2	358	644	7.68	1.84	1,43	22.3		346				1 00	<u> </u>		
	20	12.7	301	040	7.96	2.24	2.02	9.8	33	9.4	353	640	7.93	1.63	1.48	9.2	3/	12.5	340	042	7.81	2.28	1.80	21.1		
47.50	31	13.85	340	626	7.64	2.32	2.10	9.5	25	10.9	356	651	7.45	1.90	1.63	14.2										
55.00	32	15.45	318	603	7.68	2.70	2.15	19.6	20 21 22	11.9 13.5 12.0	351 360 348	700 641 639	7.74 7.72 7.63	2.02 2.50 2.18	1.80 2,15 1,85	10.9 14.0 15.1	36	13.4	320	619	7.79	2.4	1,80	25.0		

Table 5-2. SUMMARY OF RESULTS FOR RUNS AT HIGH CONCENTRATION OF CHEMICALS

T.A. = total alkalinity Ca⁺⁺ = hardness * Units in mg CaCO₃/) U values based on nominal (bare-tube) outside area. Plain annular cross sectional area = 79.128 x 10⁻⁵ m² Finned annular cross sectional area = 69.528 x 10⁻⁵ m²

T

tubes over the plain mild steel tube are the higher unit thermal conductance of the metal in case of the former and the extended surface area in case of the latter. The overall heat transfer coefficient of the finned mild steel tube, based on the outside nominal (bare-tube) area (Appendix II.2), was also shown to be larger than the plain tube coefficient either before or after fouling. The experimentally determined ratio of the clean overall heat transfer of the finned tube to that of the plain tube (Table 5-2) was about 1.25. If the fins were 100% efficient, for tubes of the same wall thickness, one would expect the value of this ratio to be approximately 3. However, the wall thickness in the finned tube is less than that for the plain tube and the fin efficiency and the total surface efficiency of the surface are 33% and 53%, respectively. Therefore, the expected result would be in a close agreement with the experimental value of this ratio (1.25). The value of this ratio was slightly higher (1.3) based on the dirty overall heat transfer coefficients (Table 5-2).

The thickness of the deposits formed during the high concentration runs increased with distance from the water entrance. In the case of the finned tube, the deposits were mostly formed on the prime surface. The formation of the deposits on the fins was not uniform and the thickness of the scale decreased with distance from the base of the fins. Both the yellow shiny deposits formed on the copper tube and the gray ones on the plain mild steel tube were hard and brittle, and could be cracked off easily. However, the deposits on the finned mild steel tube had to be scraped off with a spatula, providing a light grayish powder. The photographs of the three disassembled dirty copper, plain and finned mild steel tubes are shown in Figures 5-4, 5-5, and 5-6.









Figure 5-6. PHOTOGRAPH OF DISASSEMBLED FOULED FINNED MILD STEEL TUBE
5.2 Effect of Reynolds Number on Overall Heat Transfer Coefficient

5.2.1 Copper and Stainless Steel Tubes

The test heat exchanger tubes were examined under three different annular flow velocities and the clean and the dirty heat transfer coefficients after 48 hours and 70 hours were determined in each case. Water quality parameters and operating conditions for each run are summarized in Tables 5-3 and 5-4. The hot and cold stream temperatures and the factors representing the chemistry of water were all averaged throughout a run. The wall temperatures were calculated using the heat flow rates and the average steam temperatures (see Appendix II.6).

Figure 5-7 represents the change in both the clean and the dirty overall heat transfer coefficients with respect to the Reynolds number based on the film temperature and hence the velocity. The positive effect of Reynolds number on the value of the overall heat transfer coefficient is evident from the graphs. In either clean or dirty conditions, the overall heat transfer coefficient was higher for the copper tube than the corresponding one for the stainless steel tube. However, the percentage of drop, over the same length of time, in the value of U was higher for the copper tube than for the stainless steel one. This might have been due to both the smoothness of the stainless steel tube which weakens the adherence of the scale to the surface and the higher bulk and wall temperatures in the case of the copper tube (Tables 5-3 and 5-4).



Figure 5-7. EFFECT OF REYNOLDS NUMBER ON CLEAN AND DIRTY OVERALL HEAT TRANSFER COEFFICIENTS OF COPPER AND STAINLESS STEEL TUBES

Vol. Flow Rate _x 10 ⁻⁵ (m ³ /s)	Run	v (m/s)	Reb	Refilm	Refilm T.A.* ave	CA ⁺⁺ * ave	TDS ave	pH T _b ave a	T _b ave	T _b T _s ave ave (°C) (°C)	T _w ave	q	U _o (kW/m ² K)			Rf		
							(mg/1)		(°C)		(°C)	(kW)	Start	End	% Drop	(m ² K/kW)		
23.67	1A	0.200	0.200	0.299	6971	11136	181	298	681	7.84	22.0	106.09	95.43		1 27	1.16	15.3	0.140
	١	0.299	0271	11194	173	290	659	7.86	23.0	106.80	96.14	7.3	1.37	1.16	15.3	0.140		
20.02	3A	0 503	0.503 10554	18576	177	304	686	7.89		103.95	89.91	10.4		1.68	0.0	0.000		
39.83	3	0.503		18599	177	304	686	7,93	27.3	104.12	90.08	10.4	1.08	1.68	0.0	0.000		
	2A	0 605	14572	24984	173	294	666	7.92		102.18	86.66		0 03	1.83	9.9	0.052		
55.00	2	0.033	0 . 695 14573	25261	161.4	283	635	7.95	27.0	103.66	88.14	11.5	2.03	1.73	14.8	0.085		

.

Table 5-3. SUMMARY OF RESULTS FOR THE STAINLESS STEEL TUBE

T.A. = total alkalinity Ca⁺⁺ = hardness TDS = total dissolved solids * Units in mg CaCO₃/l

Vol. Flow Rate_x 10 ⁻⁵ (m ³ /s)	Run	v (m/s)	Reb	Refilm	T.A.* ave	* CA ⁺⁺ * ave	TDS ave (mg/1)	рН ave	Tbave	T _s ave	T _w ave	q	Uo	(kW/m ² K)	R _f (m ² K∕k₩)	
									(°C)	(*C)	(°C)	(kW)	Start	End	% Drop		
	6A	0.200	6920	12921	154	275	641	7.61	21.0	109.81	109.01			1.74	22.7	0.130	1
23.6/	6	0.299	0829	12987	141	262	579	7.60	31.0	110.60	109.80	11.5	2.25	1.69	24.9	0.147	99
	5A	0 502	11404	21170	162	281	633	8.11	20. 6	106.42	105.43		0.55	2.17	15.2	0.070	I
39.83	5	0.503	11494	21 397	150	269	600	8,11	30.6	107.83	106,84	14.1	2.50	2.09	18.4	0.088	
	44	0.000	1070	29506	169	294	661	8.03		106.80	105.54			2.73	17.0	0.062	•
55.00	4	0.095) 158/U			31.8			18.0	3,29	2.55	22.5	0.088				

Table 5-4. SUMMARY OF THE RESULTS FOR THE COPPER TUBE AT LOW CONCENTRATIONS OF CHENICALS

T.A. = total alkalinity Ca⁺⁺ = hardness TDS = total dissolved solids * Units in mg CaCO₃/l

5.2.2 Copper and Plain Mild Steel Tubes

The copper tube was tested under high concentrations of chemicals and three different flow velocities to make the comparison between the non-corroding and the corroding (plain mild steel) tubes possible. The summary of results, representing the average water quality, the operating conditions, and the overall heat transfer coefficients, are given in Tables 5-5 and 5-6.

As is shown in Figure 5-8, the clean overall heat transfer coefficient has a greater value for the copper tube than for the mild steel tube, and both coefficients rose with increasing Reynolds number. However, in the case of the copper tube at high values of the Reynolds number, the positive effect of Reynolds number on the coefficient decreased for the clean condition and vanished for the dirty one. The result was a higher percentage drop in U at higher velocities. An increase in the dirty coefficient of the mild steel tube with the Reynolds number was observed, though no specific trend was found considering the percentage of the drop in the coefficient. The dirty coefficient of the copper tube was higher than that of the mild steel tube at lower Reynolds numbers.

Runs 23 and 33 (Table 5-6) both correspond to the mild steel tube at 0.503 m/s annular flow velocity. The clean coefficient for Run 23 was much higher than the one for Run 33 due to the higher steam temperature. However, they both had close values of dirty coefficients, presumably as a result of a greater wall temperature in the case of the former. Run 20 was



Figure 5-8. EFFECT OF REYNOLDS NUMBER ON CLEAN AND DIRTY OVERALL HEAT TRANSFER COEFFICIENTS OF COPPER AND PLAIN MILD STEEL TUBES

Vol. Flow Rate x 10 ⁻⁵ (m ³ /s)	Run	v (m/s)	Reb	Refilm	T.A.* ave	CA ⁺⁺ * ave	TDS ave (mg/l)	рН ave	Tbave	T _s ave	T _w ave (*C)	9	Uo	(kW/m ² K)		R _f
									(°C)	(°C)		(kW)	Start	End	% Drop	(m ² K/kW)
23.67	38	0.299	6801	12938	361	659	1463	7.69	30.84	110.07	109.38	9.9	1.70	1.48	12.9	0.087
39.83	37	0.503	12101	22488	346	642	1420	7,81	33.29	112.81	111.94	12.5	2,28	1.80	21.1	0.117
55.00	36	0.695	17087	32072	320	619	1354	7.79	34.27	117.06	116.12	13.4	2.40	1.80	25.0	0.139

.

.

Table 5-5. RESULTS FOR THE COPPER TUBE AT HIGH CONCENTRATION OF CHEMICALS

T.A. = total alkalinity Ca⁺⁺ = hardness TDS = total dissolved solids * Units in mg CaCO₃/l

Vol. Flow Rate_x 10 ⁻⁵	Run	v (m/s)	Reb	Refilm	T.A.* ave	CA++* ave	TDS ave (mg/1)	рН - аve	T _b ave	T _s ave	T _w ave	q	U _o (kW/m ² K)			R _f
(m ³ /s)									(°C)	(°C)	(*C)	(kW)	Start	End	% Drop	(m²K/kW)
23.67	24	0, 299	6500	11952	415	697	1583	7.78	28.7	107.75	99.69	8.4	1.45	1.25	13.8	0.110
32.00	29	0.404	8777	16253	351	639	1420	7.78	28.63	109.34	100.60	9.1	1.58	1.32	16.5	0.125
20.02	23	0 503	10767	21 31 2	358	644	1436	7.68	27.92	119.77	109.02	11.2	1.84	1.43	22.3	0.156
33.03	33	0.505	10831	19394	353	640	1424	7.93	28.21	103.99	94.97	9.4	1.63	1.48	9.2	0.062
47.50	25	0,600	12660	23078	356	651	1445	7.45	27.40	105.94	95.48	10.9	1.90	1.63	14.2	0.087
55.00	20 21 22	0.695	14592 14820 17180	26624 25770 29030	351 360 348	700 642 639	1 520 1 4 3 3 1 4 1 7	7.74 7.72 7.63	27.00 27.75 34.50	106.78 103.06 111.91	95,36 90,10 100,39	11.9 13.5 12.0	2.02 2.50 2.18	1.80 2.15 1.85	10.9 14.0 15.1	0.061 0.065 0.082

Table 5-6. SUMMARY OF RESULTS FOR THE PLAIN MILD STEEL TUBE

.

T.A. = total alkalinity Ca⁺⁺ = hardness TDS = total dissolved solids * Units in mg CaCO₃/l

- 103 I. also repeated, resulting in higher values of the clean coefficient (Runs 21 and 22). In the case of Run 22, an increase in both inside and outside convective heat transfer coefficients due to higher initial steam pressure and higher initial tank water temperature, respectively, was responsible for the higher magnitude of the overall heat transfer coefficient. However, in the case of Run 21, the rise in the overall heat transfer coefficient was inexplicable due to both the lower film temperature and higher temperature difference ($\overline{T}_{save} - T_{wave}$) of the hot stream (Table 5-6). Amongst these, Run 21 yielded the highest heat flow rate as the result of the greater magnitude of the temperature (Run 21) was accompanied by an increase in the percentage drop in U which was further enhanced due to the simultaneous rise in the initial tank water temperature (Run 22).

5.2.3 Plain and Finned Mild Steel Tubes

The operating conditions and the results of the tests done on the finned mild steel tube under five different velocities are given in Table 5-7. Figure 5-9 and both Tables 5-6 and 5-7 show the positive effect of velocity on both the clean and the dirty coefficients of the tubes. The higher value of the clean coefficient due to the greater heat transfer area in the case of the finned tube is evident from the figure. The beneficial effect of the enhanced tube over the plain tube remained valid even after the fouling process. However, at Reynolds numbers approximately above

Vol. Flow Rate x 10 ⁻⁵ (m ³ /s)	Run	v (m/s)	Reb	Refilm	T.A.* ave	CA ⁺⁺ * ave	TDS ave (ma(1))	pH ave	T _b ave (°C)	T _s ave	T _w ave (*C)	q	U _O (kW/m ² K)			Rf	
								(mg/1)		(°C)	(*C)	(°C)	(kW)	Start	End	\$ Drop	(m ² K/kW)
	27	0.240	3844	6917	371	661	1477	7.64	31.88	108.73	101.87	9.80	1.61	1.54	4.3	0.028	
23.07	30	0.340	3735	6849	365	650	1453	7.54	30. 57	111.73	104.66	10,10	1.69	1.45	14.2	0.0 9 8	ı
32.00	28	0.460	4946	8973	347	633	1406	7.60	29.61	107.43	99.52	11.30	2.00	1.78	11.0	0.062	105 -
39.83	26	0.573	61 70	10904	361	648	1445	7,58	29.70	104.84	95.95	12.70	2.24	2.02	9,8	0.049	
47.50	31	0.683	7716	13579	340	626	1387	7.64	31.89	109.74	100.05	13.85	2.32	2.10	9,5	0.045	
55.00	32	0. 791	9066	15882	318	603	1 326	7.68	32.54	111.71	100.90	15.45	2.70	2.15	19.6	0.095	

Table 5-7. SUMMARY OF RESULTS FOR THE FINNED MILD STEEL TUBE

.

T.A. = total alkalinity Ca⁺⁺ = hardness TDS = total dissolved solids * Units in mg CaCO₃/l

•

. .



Figure 5-9. EFFECT OF REYNOLDS NUMBER ON CLEAN AND DIRTY OVERALL HEAT TRANSFER COEFFICIENTS OF PLAIN AND FINNED MILD STEEL TUBES

11000, this advantage decreased since the curve (U_d vs. Re_{film}) tended to flatten out (Figure 5-9). No specific relationship was found between the percentage of the drop in the overall coefficient and Reynolds number in either case.

Due to the erratic behaviour of the fouling curve, Run 27 was repeated on the finned tube. Considering the experimental errors, both the first (Run 27) and the second (Run 30) trials resulted in reasonable agreement for the heat flow rate and the clean coefficient (Table 5-7). However, the percentage of the drop in U was much higher for Run 30, yielding a lower dirty coefficient presumably as a result of higher wall temperature.

5.3 Fouling Resistance With Respect to Time

As the fouling process took place, the overall heat transfer coefficient of the exchanger was reduced due to a rise in the overall resistance of the heat transfer unit as a result of the occurrence of the fouling resistance (R_f) . Since the scaling solution was not highly concentrated even during the "high concentration" runs, the variation of the fouling resistance with respect to the time was anticipated to be gradual, showing a linear behaviour. This was the case for most of the runs. Figures 5-10 to 5-14 show the typical fouling behaviour.

The fouling resistance versus time curves for Runs 1, 6, 23, 27, 31 and 38 were far from linear, as shown in Figures 5-15 to 5-20, respectively. Run 27 (Figure 5-18) for the finned tube at 0.340 m/s annular velocity showed an inexplicable erratic behaviour whereas Run 30 under approximately identical operating conditions, resulted in a relatively linear behaviour. The asymptotic behaviour of Run 23 (Figure 5-17) on the plain mild steel tube was also inconsistent since linearity was evident during a repeat trial (Run 33). This inconsistency might have been due to the use of a new tube in case of the former. The sawtooth behaviour resulting from Run 31 (Figure 5-19) on the finned tube at 0.683 m/s flow velocity must have been due to the change in the cooling water flow rate. As a consequence of the reduction in the usage of the main supply water at night, the cooling water flow rate increased resulting in a lower inlet recirculating water This situation increased the magnitude of the temperature temperature. driving force, the rate of heat flow and the overall heat transfer coefficient, resulting in a lower value of the fouling resistance.

Runs 1, 6 and 38 (Figures 5-15, 5-16 and 5-20) were all carried out on the non-corroding (stainless steel and copper) tubes at an annular flow velocity of 0.299 m/s. These runs, either under "low concentration" (Runs 1 and 6) or "high concentration" (Run 38) of chemicals, displayed asymptotic behaviour. The initial rise in resistance was extremely rapid, occurring usually over the first three to five hours. The rapid growth of the fouling deposit would usually result in a more porous and less tenacious scale, which has low values of density and thermal conductivity, resulting in a higher magnitude of fouling resistance than a non-porous tenacious scale for



Figure 5-10. RUN 2 FOULING RESISTANCE VERSUS TIME



Figure 5-11. RUN 4A FOULING RESISTANCE VERSUS TIME



Figure 5-12. RUN 20 FOULING RESISTANCE VERSUS TIME



Figure 5-13. RUN 26 FOULING RESISTANCE VERSUS TIME



Figure 5-14. RUN 36 FOULING RESISTANCE VERSUS TIME



Figure 5-15. RUN 1 FOULING RESISTANCE VERSUS TIME



Figure 5-16. RUN 6 FOULING RESISTANCE VERSUS TIME

- 115 -



Figure 5-17. RUN 23 FOULING RESISTANCE VERSUS TIME



Figure 5-18. RUN 27 FOULING RESISTANCE VERSUS TIME



Figure 5-19. RUN 31 FOULING RESISTANCE VERSUS TIME



Figure 5-20. RUN 38 FOULING RESISTANCE VERSUS TIME

- 119 -

a specific mass of foulant. However, the former is more prone to the removal than the latter. The asymptotic form of the fouling curve for these runs might have been due to the rapid formation of the scale and the absence or negligible effect of removal as the result of the low annular velocity value during the initial part of the run, followed by enhancement of the removal mechanism at later stages due to the deposit weakness.

The remainder of the fouling curves are shown in Appendix IV.

5.4 Effect of Reynolds Number on Fouling Resistance

5.4.1 Copper and Stainless Steel Tubes

Figure 5-21 represents the fouling resistance of the copper and stainless steel tubes under low concentration of chemicals both after 48 and 70 hours plotted versus the Reynolds number. The fouling resistance of both non-corroding tubes decreased with increasing Reynolds number (see Tables 5-3 and 5-4). At low velocity in either case, the majority of the fouling process occurred during the early part of the experiment while at higher annular flow velocities, the fouling resistance increased gradually throughout the run. This was deduced by comparison between the fouling resistance of each tube after 48 and 70 hours (i.e. Run 1 and 1A), which showed good agreement with the curves representing the fouling behaviour (see Section 5.3). The stainless steel tube provided a lower fouling resistance than the copper tube over the same length of time.



Figure 5-21. EFFECT OF REYNOLDS NUMBER ON FOULING RESISTANCE FOR COPPER AND STAINLESS STEEL TUBES

As is shown in Table 5-3, the only inconsistency occurred during the experiment on the stainless steel tube at flow velocity of 0.503 m/s which resulted in zero fouling resistance due to the inexplicable absence of fouling.

5.4.2 Copper and Plain Mild Steel Tubes

The fouling resistances pertaining to the tests done under high concentration of chemicals on both the copper and the plain mild steel tubes are given in Tables 5-5 and 5-6. The comparison between these two sets of data was carried out (Figure 5-22) to see the effect of the surface material on the fouling process with respect to the Reynolds number.

As the figure shows, the fouling resistance on the copper tube increases with increasing Reynolds number. However, in the case of the mild steel tube, the initial increase in the fouling factor was followed by a decrease at Reynolds numbers approximately above 18000. This might have been due to the reduction in the stickability of the scale to the mild steel tube and hence enhancement of the removal term at higher velocity rates.

The only inconsistency involving the mild steel tube occurred during Runs 23 and 33 which resulted in a very high fouling factor in the case of the former, presumably due to the high wall temperature, and a relatively low fouling resistance in the case of the latter for an unknown reason. The enhancement of the fouling resistance as a result of either high pressure or both high temperature and pressure (Runs 21 and 22) is also



Figure 5-22. EFFECT OF REYNOLDS NUMBER ON FOULING RESISTANCE FOR COPPER AND PLAIN MILD STEEL TUBES

1

evident from the figure. In general, even though the non-corroding copper tube showed lower fouling resistance than the mild steel one at low Reynolds numbers, its efficiency tended to slacken off at higher Reynolds numbers.

5.4.3 Plain and Finned Mild Steel Tubes

The fouling resistance of the finned tube based on the outside nominal area is calculated (Table 5-7) and plotted versus both the velocity and the Reynolds number based on the equivalent diameter in Figure 5-23. Except for Run 32, the fouling resistance decreased with increases in the Reynolds number in the velocity region tested. This could be due to the geometry of the tube and hence higher level of turbulence and enhancement of the removal process at high velocity values.

The effect of velocity on the fouling resistance at Reynolds numbers above 16000 could not be investigated due to the limited recirculating water flow rate. In general, it can be considered that the enhanced tube provided more favorable fouling resistances than the plain one at the corresponding velocity values. It should also be mentioned that since the fouling resistance of the finned tube is evaluated based on the prime area, the weight of the deposit per total area and hence the thickness of the deposit is smaller (about 14%) on the finned tube than that on the plain tube for the same magnitude of the fouling resistance based on the prime area (see Appendices II.5 and II.10).



Figure 5-23. EFFECT OF BOTH VELOCITY AND REYNOLDS NUMBER ON FOULING RESISTANCE FOR PLAIN AND FINNED MILD STEEL TUBES

5.5 Determination of the Fouling Rate

5.5.1 Measured Fouling Rate

The fouling rate was determined through numerical analysis using the U.B.C. DLQF curve fitting routine and the experimental fouling resistance data. Linear, asymptotic and 4th order polynomial functions were used for fitting (see Appendix III). Even though the linear function provided the most suitable fit for the majority of the runs, the asymptotic one could be considered the best fit for a few others as was evident from the fouling resistance curve versus the time (see Section 5.3).

5.5.2 Predicted Fouling Rate

The average water chemistry parameters and the operating condition data such as velocity and temperature were used to calculate the predicted fouling rate based on the ionic diffusion model of Hasson (Appendix III). The average concentrations of the carbon species and calcium ion for each run were determined. The fouling rates were calculated using both Hasson's low and high pH equations. Since the bicarbonate ion concentration was higher than the carbonate one, the fouling rate based on the low pH equation was considered to be the effective one for all runs. The results of model calculations are summarized in Table 5-8.

Run	^{pH} ave	HC03 x 10 ^{2*}	CO ₃ x 10 ^{4*}	CO ₂ x 10 ^{4*}	Ca ⁺⁺ x 10 ^{2*}	Calculated** Scaling Rate x 10 ⁶
1A	7.84	0.1748	0.2687	0.4237	0.2977	0, 0914
1	7.86	0.1668	0.2689	0.3860	0.2897	0.0844
3A	7.89	0.1703	0.2944	0.3672	0.3037	0.1353
3	7,93	0.1696	0.3218	0.3337	0.3037	0,1359
2A	7.92	0,1661	0.3032	0.3347	0.2937	0,1586
2	7.95	0.1540	0.3029	0.2895	0.2827	0.1403
6A	7.61	0.1504	0.1422	0.6256	0.2748	0.0853
6	7.60	0.1377	0.1273	0.5868	0.2618	0.0715
5A	8.11	0.1510	0.4488	0.1983	0.2807	0.1415
5	8.11	0.1396	0.4158	0.1835	0.2688	0.1242
4A	8.03	0.1593	0.3961	0.2513	0.2937	0.2075
24	7.78	0.4017	0.6056	1,1068	0.6964	0.4747
29	7.78	0.3397	0.5144	0.9352	0.6384	0.4815
23	7.68	0.3485	0.4240	1.2149	0.6434	0.6327
33	7.93	0.3375	0.7123	0.6565	0.6394	0,5480
25	7.45	0.3505	0.2457	2.0562	0.6504	0,6388
20	7.74	0.3406	0.4781	1.0212	0.6994	0, 7281
21	7.72	0.3502	0.4480	1.1040	0.6404	0.7048
22	7.63	0.3397	0.3683	1.3261	0,6384	0.7807
30	7.54	0.3579	0.3145	1.7205	0.6494	0,5233
27	7.64	0.3621	0.4007	1.3800	0.6604	0,5321
28	7.60	0.3394	0.3393	1.4154	0.6324	0, 5938
26	7.58	0.3536	0.3352	1.5412	0.6474	0,7316
31	7.64	0.3318	0.3654	1.2634	0.6254	0, 81 81
32	7.68	0.3096	0.3745	1.0763	0.6025	0.8464
38	7.69	0.3511	0.4413	1.1983	0.6584	0.4289
37	7.81	0.3334	0.5541	0.8679	0.6414	0.6416
36	7.79	0.3086	0.4915	0.8457	0.6184	0.7797

Table 5-8. SUMMARY OF MODEL CALCULATIONS

* Units in mole/1.
** Unit in m²K/kJ.

.

5.6 Effect of Reynolds Number on the Measured Fouling Rate

5.6.1 Copper and Stainless Steel Tubes

The experimental measurements of the fouling rate under low concentration of chemicals on the two non-corroding tubes accompanied by the type of fitted curves are given in Table 5-9. As is shown in the table, neither of the tubes showed any specific trend with respect to the Reynolds number. Asymptotic behaviour was the only common effect of low velocity value on the rate of fouling in either case, presumably due to the negligible effect or absence of the removal process in the early part of the run.

5.6.2 Copper and Plain Mild Steel Tubes

The copper tube under high concentration of chemicals also showed asymptotic behaviour at low velocity leading to a very high initial fouling rate in comparison with the corresponding ones at high velocity values. The high rate of fouling at low Reynolds number (Run 38) might have been due to the absence of a removal process and hence the rapid formation of the scale. However, the strength of the scale might be compromised by the high rate of fouling. Therefore, the growth in the deposit thickness and the reduction in the adherence factor would result in the asymptotic behaviour. At high velocities, both the gross deposition and the removal terms are effective from the start of the run resulting in both a more gradual increase in the

v (m/s)		Сорре	r		Stainless Steel							
	Run*	^{Re} film	$\frac{dR}{d\theta} \times 10^{6}$ (m ² K/kJ)	Fit	Run	Re _{film}	$\frac{dR}{d\theta} \times 10^{6}$ (m ² K/kJ)	Fit				
0.299	6A 6	12921 12987	1.9766 1.9334	Asymp Asymp	1A 1	11136 11194	12.3109 12.7204	Asymp Asymp				
0.503	5A 5	21170 21397	0.3545 0.4224	Asymp Asymp	3A 3	18576 18599	-0.0570 -0.0204	Lin Lin				
0.695	4A	29506	0.3678	Lin	2A 2	24984 25261	0.0808 0.3138	Lin Lin				

.

Table 5-9. COMPARISON OF MEASURED FOULING RATES ON COPPER AND STAINLESS STEEL

Note: A = runs lasted 48 hours Lin = linear Asymp = asymptotic * Low concentration runs scale thickness and a stronger deposit. Having a higher fouling rate for the run at 0.695 m/s than the one at 0.503 m/s (Figure 5-24) indicated that the effect of increasing velocity on the removal process was counterbalanced by the high gross deposition rate.

The net rate of fouling on the plain mild steel tube was also affected by the velocity (Figure 5-24). No asymptotic behaviour was shown. even at low values of Reynolds numbers. This was attributed to the surface material effect. The fouling rate was increased with increasing velocity up to a Reynolds number of about 17000 due to the enhancement of the gross However, the effect of the removal rate became more deposition rate. pronounced at higher Reynolds numbers, resulting in a reduction of the net fouling rate with further increases in the magnitude of the velocity. The simultaneous effects of both higher temperature and pressure leading to a higher fouling rate can be seen in comparison between Runs 20 and 22. However, an increase in the fouling rate with an increase only in the initial steam pressure (Run 21) was not evident due to either experimental errors or the curve fitting procedure which covered all data points. With regard to Figure 5-24, it can be considered that the plain mild steel tube provided lower fouling rates than the corresponding ones on the copper tube, but more data is needed to confirm this result.



FIGURE 5-24. EFFECT OF REYNOLDS NUMBER ON MEASURED FOULING RATE FOR COPPER AND PLAIN MILD STEEL TUBES - 131 -
5.6.3 Plain and Finned Mild Steel Tubes

Velocity was also one of the effective factors on the net rate of fouling of the finned tube. Figure 5-25 shows both the velocity and the Reynolds number effects on both the plain and the enhanced mild steel tubes. The fouling rate values were determined by fitting the entire data using a numerical analysis method (see Section 5.5.1). However, in the case of Run 31, the curve fitting method could not supply a reliable fit due to the periodic changes in the cooling water flow rate and hence the sawtooth behaviour of the fouling resistance curve (see Section 5.3). For this run, an attempt was made to visually fit the data points corresponding to the daytime (8:00 a.m. to 8:00 p.m.) cooling water flow rates.

Disregarding Run 32, the general outlook provided by the experimental data on the enhanced tube showed a drop in the fouling rate with increasing velocity. The reduction might have been due to higher magnitude of the removal term, the controlling effect of the surface reaction mechanism or both at the high velocity rates. The reason for rapid growth in the value of fouling rate at 0.791 m/s (Run 32) is not known.

The reduction in the rate of fouling took place at Reynolds numbers approximately above 16000 (V=0.404 m/s) for the plain tube, while at Reynolds numbers about 7000 (V=0.340 m/s) for the finned (Figure 5-25). This might be due to the fact that the same degree of turbulence occurs at the lower values of Reynolds number in case of the enhanced tube as a result

- 132 -



FIGURE 5-25. EFFECT OF BOTH VELOCITY AND REYNOLDS NUMBER ON MEASURED FOULING RATE FOR PLAIN AND FINNED MILD STEEL TUBES

of the existence of the fins. Generally, the finned tube provided lower and more favourable fouling rates than did the plain one, at equivalent annular flow velocity values.

5.7 Effect of Reynolds Number on the Predicted Fouling Rates

An increase in the predicted fouling rate with velocity was expected since the model was based on both the diffusion controlled mechanism and the absence of the removal term. Due to the fact that the model predicts a linear behaviour for the fouling curve and the rate should increase with an increase in the Reynolds number, an increase in the fouling resistance with the velocity is expected.

Figure 5-26 and Table 5-10 both represent the effect of Reynolds number on the non-corroding tubes under low concentrations of chemicals. Model calculations resulted in a lower magnitude of fouling rate for the copper tube than the stainless steel one under approximately identical operating conditions at Reynolds numbers below 22000. The fact that the plot of the predicted fouling rate versus Reynolds number for the stainless steel tube had a reduction in its slope at higher annular flow velocity value might have been due to the smoothness of the surface which became more effective in diminishing the occurrence of the fouling process in the presence of higher degree of turbulence. The basis of the model, the constancy of the fouling rate with respect to the time, was evident considering the close values of predicted rate corresponding to a run over two different periods of time (i.e. 5 and 5A). A decrease in experimental fouling resistance with velocity (Figure 5-21) was in contradiction with the model's prediction (Figure 5-26).



FIGURE 5-26. EFFECT OF REYNOLDS NUMBER ON PREDICTED FOULING RATE FOR COPPER AND STAINLESS STEEL TUBES

v		Сорре	r		Stainless :	Steel
(m/s)	Run*	Re _{film}	$\frac{dR}{d\Theta} \times 10^{6}$ (m ² K/kJ)	Run	Re _{film}	$\frac{dR}{d\theta} \times 10^{6}$ (m ² K/kJ)
0.299	6A 6	12921 12987	0.0893 0.0757	1A 1	11136 11194	0.0956 0.0886
0.503	5A 5	21170 21397	0.1497 0.1323	3A 3	18576 18599	0.1416 0.1424
0.695	4 A	29506	0.2180	2A 2	24984 25261	0.1666 0.1484

Table 5-10. COMPARISON OF PREDICTED FOULING RATES ON COPPER AND STAINLESS STEEL

Note: A = runs lasted 48 hours. * Low concentration runs.

The effect of Reynolds number on predictions for both the copper and the plain mild steel tubes under high concentration of chemicals is shown in Figure 5-27. An increase in the fouling rate with the velocity was also held to be generally in effect for these tubes. The slight inconsistency occurred during Run 24, on the plain mild steel tube, which was presumably due to the relatively high values of total alkalinity and hardness in comparison to the other runs. The single effect of the initial steam pressure and the combined effect of both the inlet water temperature and the initial steam pressure are shown by Runs 21 and 22, respectively. Comparing Runs 20 and 21, the lower value of the fouling rate for Run 21 is understandable and can be attributed to its lower average wall temperature (Table 5-6). The enhancement of the experimental fouling resistance with the velocity (Figure 5-22) was in good agreement with Hasson's model in the case of the copper tube. However, the reverse relationship in the case of the plain mild steel tube (Figure 5-22) was contradictory with the predicted fouling rate results (Figure 5-27), suggesting the ionic diffusion plus reaction did not control the fouling rate on mild steel. In general, the predicted rates on both tubes were similar at corresponding annular flow velocities.

As is shown in Figure 5-28, the predicted fouling rate also increased with increasing velocity in the case of the finned mild steel tube. This trend was opposite to the variation of the fouling resistance with the Reynolds number provided by the experimental data (Figure 5-23). Predicted fouling rate values were indicative of the fact that the plain tube behaved more favourably than the enhanced one in the velocity region



FIGURE 5-27. EFFECT OF REYNOLDS NUMBER ON PREDICTED FOULING RATE FOR COPPER AND PLAIN MILD STEEL TUBES

138



FIGURE 5-28. EFFECT OF BOTH VELOCITY AND REYNOLDS NUMBER ON PREDICTED FOULING RATE FOR PLAIN AND FINNED MILD STEEL TUBES

investigated. This is due to the effects of the fins on the hydraulic diameter, and hence the Reynolds number. This trend was also in contradiction with the results based on the fitted value of fouling rates (Figure 5-25).

5.8 Comparison Between the Predicted and the Measured Fouling Rates

Tables 5-11, 5-12 and 5-13 represent the ratio of the predicted to the measured fouling rates for all runs. The graphical representation of the change in this ratio for low concentration runs with respect to Reynolds number (Tables 5-11 and 5-12) was not feasible due to the wide range of variation in its magnitude in the velocity region tested. Data corresponding to the copper tube under low concentration of chemicals suggest that the model underpredicts the fouling rate, particularly at low values of the annular velocity. This is also the case for stainless steel tube at low Reynolds numbers. However, no other conclusion could be drawn regarding the other runs on the stainless steel tube.

For the copper and both mild steel tubes under high concentration of chemicals, the model overpredicted the experimental fouling rates (Table 5-13). The only inconsistency occurred with respect to Run 38 due to the asymptotic behaviour of the fouling curve (see Section 5.3). Figure 5-29 shows the effect of Reynolds number on the ratio of the measured to the predicted fouling rates for these high concentration runs. In the case of the two mild steel tubes, the consistency between the model and the experimental data decreases with increasing Reynolds number, presumably due

Run*	v	Resim	dR/d⊖ × (m ² K	10 ⁶ /kJ)	(dR/də) _{meas}
	(m/s)	F I 101	Predicted	Measured	(dR/de)pred
6A 6	0.299	12921 12987	0.0893 0.0757	1.9766 1.9333	22.134 25.539
5A 5	0.503	21170 21397	0.1497 0.1323	0.3545 0.4224	2.368 3.193
4A	0.695	29506	0.2180	0.3678	1.687

Table 5-11. COMPARISON OF PREDICTED AND MEASURED FOULING RATES ON COPPER

Note: A = runs lasted 48 hours. * Low concentration runs.

Table 5-12. COMPARISON OF PREDICTED AND MEASURED FOULING RATES ON STAINLESS STEEL

Run*	v	Recta	dR/d⊖ x (m ² K,	10 ⁶ /kJ)	(dR/de) meas
	(m/s)	TIM	Predicted	Measured	(dR/de)pred
1A	0.299	11136	0.0956	12.3109	128.775
1		11194	0.0886	12.7204	143.57
3A	0.503	18576	0.1416	-0.0570	-0.403
3		18599	0.1424	-0.0204	-0.143
2A	0.695	24984	0.1666	0.0808	0.485
2		25261	0.1484	0.3138	0.473

Note: A = runs lasted 48 hours. * Low concentration runs.

	Tube Material		$\frac{dR}{d\Theta} \times 10^6$				
Run	and Geometry	v	Refilm	(m ² k	(/kJ)	(dR/də) meas	
	-	(m/s)	1 1 1 11	Predicted	Measured	(dR/de)pred	
38	Сор	0.299	12938	0.4288	3.3839*	7.8897	
37 36	Сор Сор	0.503 0.695	22488 32072	0.6423 0.7826	0.3369 0.6442	0.6055 0.8262	
24	P.M.S.	0.299	11952	0.4723	0.3552	0.7483	
29 33 25	P.M.S. P.M.S.	0.404	19394	0.4818	0.3453	0.9784 0.6301	
25	P.M.S. P.M.S.	0.600	23078	0.6389 0.7286	0.3196 0.2541	0.5003 0.3490	
22	P.M.S. P.M.S.	0.695	25770 29030	0.7047 0.7814	0.2156 0.3369	0.3050 0.4315	
30	F.M.S.	0.340	6849	0.5230	0.4173	0.7974	
28	F.M.S. F.M.S.	0.460	8973 10904	0.5943 0.7314	0.2415 0.1455	0.4067 0.1989	
31 32	F.M.S. F.M.S.	0.683	13579 15882	0.8194 0.8497	0.1000** 0.3920	0.1222 0.4631	

Table 5-13. COMPARISON OF PREDICTED AND MEASURED FOULING RATES FOR HIGH CONCENTRATION RUNS

Cop = copper
P.M.S. = plain mild steel
F.M.S. = finned mild steel
* Asymptotic fit
** Visually fitted

to the enhancement of removal process at high velocity values. The copper tube provided good agreement between the predicted and experimental value of the fouling rate at annular flow velocity of 0.695 m/s. No other specific relationship was found between velocity and the magnitude of this ratio using the copper tube. In general, for either the copper tube or the mild steel ones, the effect of Reynolds number on the magnitude of this ratio followed the same trend as did the value of the corresponding measured fouling rate (see Figures 5-24 and 5-25).



PREDICTED BY THE IONIC DIFFUSION MODEL

- 144 -

6. CONCLUSION

The effect of the fouling process on the two non-corroding (copper and stainless steel) tubes and the performance of the exchanger either before or after the deposit accumulation was examined. The copper tube showed higher values of heat flow rate and both clean and dirty overall heat transfer coefficients than did the stainless steel tube. The same trend was applicable regarding the percentage of drop in the clean overall heat transfer coefficient, over the same period of time, due to the smoothness of the stainless steel tube and higher bulk and wall temperatures for the runs With respect to the Reynolds number, the fouling on the copper one. resistance of the copper tube was also higher than the corresponding one on the stainless steel tube. In either case, the fouling resistance was generally linear in time except for the low velocity (0.299 m/s) runs which showed an asymptotic behaviour. The magnitude of the fouling resistance, for either tube, decreased with increasing velocity, unlike the values of heat flow rate, clean and dirty overall heat transfer coefficients which showed increases with the Reynolds number. Therefore, even though the gain in performance of the copper tube over the stainless steel one diminished due to the fouling process, the copper tube was more favourable. The predicted rates of fouling were slightly lower for the copper tube than the corresponding ones on the stainless steel tube at Reynolds numbers below 22000. In either case, the predicted fouling rates increased with increasing Reynolds number. In general, even though the copper tube was more prone to fouling, it provided a higher degree of performance than the stainless steel one both before and after the scaling process.

Under high concentration of chemicals, i.e. alkalinity \sim 300-400 mg $CaCO_2/1$, the copper tube also had higher values of heat flow rate and both initial and final overall heat transfer coefficients than did a plain mild steel tube. In both cases, the magnitude of these parameters increased with the velocity. The fouling curves generally followed linear behaviour. For the copper tube, the magnitudes of fouling resistance and percentage of drop in the clean overall heat transfer coefficient increased with increasing However, for the plain mild steel tube, these values were velocity. increased with the velocity only for Reynolds numbers approximately below 18000 (v=0.503 m/s). The result was a lower magnitude of fouling resistance for the copper tube than the plain mild steel one at low velocity values (v<0.503) followed by a higher value for the former than the latter at Reynolds numbers above 18000 (v>0.503 m/s). For either tube. the experimental values of the fouling rate with respect to the Reynolds number followed the same trend as did the fouling resistance except for Run 38. In general, the measured fouling rates indicated lower values for the plain mild steel tube than the corresponding ones of the copper tube. Though calculations based on Hasson's ionic diffusion model provided close values for the corresponding runs on either tube, the model over-predicted the experimental rates of fouling except for Run 38. The predicted rates increased with the velocity resulting in a better agreement with the measured values at high Reynolds numbers in the case of the copper tube, while giving rise to more suitable predictions at low velocities using the plain mild steel tube. Totally, it can be said that the copper tube had always a better performance, regarding the dirty heat transfer coefficient, than the plain mild steel one, especially at low annular flow velocities.

In comparison between the two corroding mild steel tubes, the plain tube provided lower values of heat flow rate and both clean and dirty heat transfer coefficients. In addition, a near linear increase of these parameters with the velocity was evident. The linearity of the fouling curves with respect to time was also generally evident, except for a run on a new tube. The fouling resistance for the finned tube generally decreased with increasing velocity. However, in the case of the plain tube, the fouling resistance reached its maximum at Reynolds number of approximately 18000 as was mentioned earlier. The enhanced tube always gave rise to a lower value of fouling rate than did the plain tube in the velocity region investigated. The measured rates of fouling followed the same trend with respect to velocity as did the fouling resistances. The predicted rates, in either case, rose with the velocity. However, unlike the measured values, they were lower for the plain tube than the finned one. For both tubes, the model over-predicted the experimental values of fouling rates providing a better agreement with the measured rates at lower velocities. In general, the finned tube was shown to be more favourable than the plain one regarding both the overall heat transfer coefficient and the fouling resistance build up over the same period of time.

Testing the tubes under low concentration of chemicals (alkalinity \sim 180 mg CaCO₃) resulted in rather soft and powdery deposits, on the copper and stainless steel tubes, and muddy deposits on the mild steel tubes which precluded firm conclusions of on the effect of the fouling process on the mild steel tube under these conditions. When a more concentrated scaling solution was used, the deposits were much stronger and

more resistant to removal. The corroding effect of the mild steel tube which enhanced the fouling process providing a higher magnitude of the fouling resistance for the mild steel tube than the copper at low velocity was counterbalanced for copper by an increase in the strength of the scale at high velocity values. Between the two mild steel tubes, the enhanced one had lower values of fouling resistance than the plain one at identical Reynolds numbers. At the same Reynolds number, the velocity in the finned tube is higher than that of the plain tube. Thus, the more favourable behaviour of the finned tube can be attributed either to the higher degree of turbulence or the soft and powdery structure of the deposits which were more prone to the removal mechanism. The reducing effect of velocity on the fouling resistance was also reasonable considering the structure of the deposits formed on the enhanced tube. In conclusion, the greater values of both clean and dirty heat transfer coefficients along with the lower magnitudes of fouling resistance made the enhanced tube a better choice, where any hard-water scaling is expected. In addition, even though the ionic diffusion model does

not yield good agreement with the experimental results, it can be safely used to predict the magnitude of the fouling rates for the copper and both mild steel tubes under high concentration of chemicals in the velocity region tested.

7. NOMENCLATURE

<u>Symbol</u>

Description

^a 0, ^a 1, ^a 2,, ^a 9	deposition constants
A	surface area
Ā	logarithmic mean area
A _{eff}	effective area
A _f	surface area of the fins
A _p , A _{Prime}	prime or unfinned area
А _о	nominal (bare-tube) outside area
At	total surface area
Α'	volume of the titrant (EDTA)
Asymp	asymptotic
^b ₁ , ^b ₂ , ^b ₃	removal constants
b	length of the fin
b'	degree of supersaturation
В'	volume of the titrant (HCl)
B''	mg of CaCO ₃ equivalent to 1 ml of EDTA
с _ь	bulk concentration
С _р	specific heat
° _s	concentration of saturated liquid at
	the surface
с _т	total carbon species concentration

Symbol	Description
D	diameter
D _{AB}	diffusivity
D _e	equivalent diameter
D _s	diameter of the shell
Dt	diameter of the tube
E	activation energy
h, h _c	convective heat transfer coefficient
h _d	dirt coefficient, reciprocal of fouling
	resistance for unit surface area
HTRI	Heat Transfer Research Incorporated
J	constant in the Arrhenius equation
k	molar solubility constant of CaCO ₃
k'	rate coefficient for bicarbonate
	decomposition reaction
k _{BF}	mass transfer coefficient for diffusion
	of Ca(HCO ₃) ₂
k _f	thermal conductivity of deposit
^k fin	thermal conductivity of fin
k ₁	thermal conductivity of liquid
k _w	average unit thermal conductance of wall

.

- 150 -

-

<u>Symbol</u>

De	sc	ri	р	t	i	on
 _	_		_	_	-	_

К	constant for the Reitzer gross
	deposition rate
К1	first molar dissociation constant of
	carbonic acid
K ₂	second molar dissociation constant of
	carbonic acid
К _D	mass transfer coefficient for diffusion
К _О	overall mass transfer coefficient
κ _R	rate constant for surface reaction
К _{sp}	molar solubility product of CaCO ₃
1	height of the fin
L	length of the tube
Lin	linear
LMTD	Logarithmic Mean Temperature Difference
LSI	Langelier Saturation Index
m	clean fin efficiency parameter
^m d	dirty fin efficiency parameter
^m f	mass of deposit per unit area
^m t	total mass of the deposit
m	net rate of fouling
^m d	gross rate of deposition
^m r	rate of removal
М	constant for the mechanical strength of
	the deposit

Symbol	Description
N	normality of the titrant (HCl)
NFA	Net Free Area
Ρ	fin perimeter
Poly	polynomial
Pd	probability function of velocity
q	heat flux
9 _C	heat flux by convection
۹ _f	rate of heat flow by the finned area
۹ _p	rate of heat flow by the prime
	(unfinned) area
9 _t	total rate of heat flow
R	unit thermal resistance
R _b	deposit bond resistance
R _f	fouling resistance
Rg	gas constant
R _o	unit thermal resistance at time zero
RSI	Ryzner Stability Index
R*	asymptotic fouling resistance
S	annular cross sectional area
s _f	fin spacing
t	thickness of the fin

..

Symbol	Description
т	temperature
т _b	bulk temperature
T _c	temperature of cold stream
T _f	film temperature
T _h	temperature of hot stream
T _S	surface temperature
T _s	steam temperature
T _w	wall temperature
TEMA	Tubular Exchanger Manufacturers
	Association
TWP	Total Wetted Perimeter
U .	overall heat transfer coefficient
U _c	clean overall heat transfer coefficient
Ud	dirty overall heat transfer coefficient
V	velocity
۷	volume of the water sample
v*	critical friction velocity
٧	volumetric flow rate
W	rate of deposition in Hasson ionic
	diffusion model
W	mass flow rate
W _c	mass flow rate of cold stream
W _h	mass flow rate of hot stream
× _f	deposit thickness

1

Dimensionless Groups Description

De	22	ri	nt	in	n

-

Pr	Prandtl number, Cpµ/k
Nu	Nusselt number, hD/k
Re	Reynolds number, PVD/ $_{\mu}$
Sc	Schmidt number, μ/PD

Greek Letters

Description

α	distribution fraction of carbon species					
^β 2	a function dependent on wall, scale					
	surface and bulk temperatures					
Δ	difference between values					
۵T _m	logarithmic mean temperature difference					
	(LMTD)					
^η f	fin efficiency					
ⁿ fd	dirty fin efficiency					
ⁿ t	total surface efficiency					
ⁿ td	total dirty surface efficiency					
π	constant = 3.14					
θ	time					
°с	time constant					
ө _D	delay time					
μ	viscosity					
ρ	density of the liquid					
٩	density of the deposit					

τ	liquid shear stress
Ψ	deposit structure factor
Ω	water characterization factor

<u>Subscripts</u>

Description

ave	average
b	bul k
i, in	inside
k	conduction
max	maximum
meas	measured
o, out	outside
pred	predicted

Superscripts

Description

В	exponent
f	exponent
g	exponent
n	exponent
r	exponent

.

8. REFERENCES

- Standards of Tubular Exchanger Manufacturers Association, 5th Edition, New York, 1968.
- 2) Thackery, P.A., "The Cost of Fouling in Heat Exchanger Plant". In: Proceedings of the Conference "Fouling - Science or Art_{μ} ", Surrey, Guildford, England, March 27-28, 1979.
- 3) Hasson, D., Sherman, H. and Biton, M., "Prediction of Calcium Carbonate Scaling Rates". Proceedings 6th International Symposium on Fresh Water from the Sea, Vol. 2, p. 193, 1978.
- Epstein, N., "Fouling in Heat Exchangers". Proceedings of 6th International Heat Transfer Conference, Vol. 6, p. 235-253, August 1978.
- 5) Kern, D.G. and Seaton, R.E., "A Theoretical Analysis of Thermal Surface Fouling". British Chemical Engineering, Vol. 4, p. 258, May 1959.
- Kern, D.G. and Seaton, R.E., "Surface Fouling . . . How to Calculate Limits". Chemical Engineering Progress, Vol. 55, p. 71, June 1959.
- 7) Banchero, J.T. and Gordon, K.F., "Scale Deposition on a Heated Surface". Advances in Chemistry Series, Vol. 27, p. 105, April 1960.

- 8) Hasson, D., "Precipitation Fouling". In "Fouling of Heat Transfer Equipment", Somerscales, E. and Knudsen, J.G. (Editors), Hemisphere Publishing Corporation, p. 527-568, 1981.
- 9) Larson, T.E. and Buswell, A.M., "Calcium Carbonate Saturation Index and Alkalinity Interpretation". Journal of American Water Work Association, Vol. 34, p. 1667, 1942.
- Loewenthal, R.E. and Marais, G.V.R., "Carbonate Chemistry of Aquatic Systems: Theory and Application", p. 125, Ann Arbor Science Publishers Incorporation, Ann Arbor, Michigan, 1976.
- 11) McCabe, W.L. and Robinson, C.S., "Evaporator Scale Formation". Industrial and Engineering Chemistry, Vol. 16, No. 5, p. 478, January 18, 1924.
- 12) Hasson, D., "Rate of Decrease of Heat Transfer due to Scale Deposition". Dechema-Monographien, Vol. 47, p. 233, 1962.
- 13) Hixon, A.W. and Knox, K.L., "Effect of Agitation on Rate of Growth of Single Crystals". Industrial and Engineering Chemistry Engineering and Process Development, Vol. 43, No. 9, p. 2144, September 1951.
- 14) Treybal, R.E., "Mass Transfer Operations", Third Edition, Chemical Engineering Series, McGraw-Hill Book Company, New York, 1980.

- 15) Reitzer, B.J., "Rate of Scale Formation in Tubular Heat Exchangers". Industrial and Engineering Chemistry Process Design and Development, Vol. 3, No. 4, p. 345, October 1964.
- 16) Bransom, S.H., "Factor in Design of Continuous Crystallisers". British Chemical Engineering, Vol. 5, No. 12, p. 838, 1960.
- 17) Hasson, D., Avriel, M., Resnick, W., Razenman, T. and Shlomo, W., "Mechanism of Calcium Carbonate Scale Deposition on Heat-Transfer Surfaces". Industrial and Engineering Chemistry Fundamentals, Vol. 7, No. 1, February 1968.
- 18) Taborek, J., Aoki, T., Ritter, R.B. and Palen, J.W., "Fouling: the Major Unresolved Problem in Heat Transfer Transfer". Chemical Engineering Progress, Vol. 68, No. 2, February 1972.
- 19) Taborek, J., Aoki, T., Ritter, R.B. and Palen, J.W., "Predictive Methods for Fouling Behavior". Chemical Engineering Progress, Vol. 68, No. 7, July 1972.
- 20) Watkinson, A.P. and Martinez, O., "Scaling of Heat Exchanger Tubes by Calcium Carbonate". Transactions of the American Symposium of Mechanical Engineering, Journal of Heat Transfer, Vol. 97, p. 504, November 1975.

- 21) Morse, R.W. and Knudsen, J.G., "Effect of Alkalinity on the Scaling of Simulated Cooling Tower Water". The Canadian Journal of Chemical Engineering, Vol. 55, p. 272, June 1977.
- 22) Nancollas, G.H. and Reddy, M.M., "The Crystallization of Calcium Carbonate: II Calcite Growth Mechanism". Journal of Colloid and Interface Science, Vol. 37, No. 4, p. 824, December, 1971.
- 23) Wiechers, H.N.S., Sturrock, P. and Marais, G.V.R., "Calcium Carbonate Crystallization Kinetics". Water Research, Vol. 9, p. 835, 1975.
- 24) Hasson, D. and Gazit, E., "Scale Deposition From an Evaporating Falling Film". Desalination, Vol. 17, p. 339, 1975.
- 25) Lord, R.C., Minton, P.E. and Slusser, R.P., "Design of Heat Exchangers". Chemical Engineering, p. 96, January 26, 1970.
- 26) Fanaritis, J.P. and Bevevino, J.W., "Designing Shell and Tube Heat Exchangers". Chemical Engineering, p. 62, July 5, 1976.
- 27) Cleaver, J.W., Yates, B., "Mechanism of Detachment of Colloidal Particles from a Flat Substrate in a Turbulent Flow". Journal of Colloid and Interface Science, Vol. 44, p. 464, September 1973.

- 28) Cleaver, J.W. and Yates, B., "The Effect of Re-Entrainment on Particle Deposition". Chemical Engineering Science, Vol. 31, p. 147, 1976.
- 29) Suitor, J.W., Marner, W.J. and Ritter, R.B., "The History and Status of Research in Fouling of Heat Exchangers in Cooling Water Service". The Canadian Journal of Chemical Engineering, Vol. 55, p. 374, August 1977.
- 30) Knudsen, J. and Roy, B.V., "Studies on the Scaling of Cooling Tower Water". International Conference of Heat Exchanger Surfaces, Engineering Foundation, White Haven, November 1982.
- 31) "Principles of Industrial Water Treatment", Third Edition, Drew Chemical Corporation, New Jersey, 1979.
- 32) Langelier, W.F., "The Analytical Control of Anti-Corrosion Water Treatment". Journal of the American Water Work Association, Vol. 28, p. 1500, 1936.
- 33) Langelier, W.F., "Chemical Equilibria in Water Treatment". Journal of the American Water Work Association, Vol. 38, No. 2, p. 169, February 1946.
- 34) Watkinson, A.P., "Effects of Water Quality on Hard Water Scaling". Proceedings of the 30th Canadian Chemical Engineering Conference, National Heat Transfer Symposium, Vol. 2, p. 616, October 1980.

- 35) Watkinson, A.P., "Process Heat Transfer: Some Practical Problems". The Canadian Journal of Chemical Engineering, Vol. 58, p. 553, October 1980.
- 36) Ryznar, J.W., "A New Index for Determining Amount of Calcium Carbonate Scale Formed by a Water". Journal of the American Water Work Association, Vol. 36, p. 472, April 1944.
- 37) HTRI set of Electron Scanning Microscope (ESM) Pictures showing various stages of crystalline growth and behaviour. Available upon request.
- 38) Thurston, E.F., "Experimental Plant for Studying Methods of Controlling Scale Formation in Boilers". Chemistry and Industry, p. 1238, July 10, 1965.
- 39) Watkinson, A.P., Water Quality Effects on Fouling from Hard Waters. In: "Heat Exchangers - Theory and Practice", Taborek, J., Hewitt, G.F. and Afgan, N. (Editors), Hemisphere Publishing Corporation, 1983.
- 40) Peters, R.W. and Stevens, J.D., "Effect of Iron as a Trace Impurity on the Water Softening Process". The American Institute of Chemical Engineers Sympsoium Series, p. 46, 1982.

- 41) Peters, R.W. and Stevens, J.D., "Additivity of Crystal Size Distributions in the Simultaneous Precipitations of $CaCO_3$ and $Mg(OH)_2$ ". Proceedings of the Second World Congress of Chemical Engineering, Vol. 4, 76-81, October, 1981.
- 42) Kemmer, F.N., "Water: The Universal Solvent". Nelco Water Handbook, Second Edition, p. 58, Nelco Chemical Company, February 1979.
- 43) Somerscales, E.F.C., Fundamental Ideas in Corrosion Testing in the Presence of Heat Transfer. In: Symposium "Corrosion in Heat Transfer Conditions", Teddington, England, November 3, 1982.
- 44) "BETZ Handbook of Industrial Water Conditioning", Eighth Edition, BETZ Laboratories Incorporated, Trevose, Pennsylvania, 1980.
- 45) Uhlig, H.H., "Corrosion and Corrosion Control", 4th Printing, John Wiley a Sons Incorporated, New York, 1967.
- 46) Kreith, F., "Principles of Heat Transfer", Third Edition, Harper Row Publishers, Incorporated, New York, 1973.
- 47) Kern, D.G. and Kraus, A.D., "Extended Surface Heat Transfer". McGraw-Hill Book Company, New York, 1972.
- 48) Kern, D.G., "Process Heat Transfer". McGraw-Hill Book Company, New York, 1950.

- 49) Knudsen, J.G. and Katz, D.L., "Heat Transfer and Pressure Drop in Annuli". Chemical Engineering Progress, Vol. 46, No. 10, p. 490, October 1950.
- 50) Knudsen, J.G. and McCluer, H.K., "Hard Water Scaling of Finned Tubes at Moderate Temperatures". Chemical Engineering Symposium Series, Vol. 55, No. 29, 1959.
- 51) Katz, D.L. et al., University of Michigan. Engineering Research Institute, Project M592, July 1953.
- 52) Webber, W.O., "Under Fouling Conditions Finned Tubes Can Save Money". Chemical Engineering, Vol. 67, No. 6, p. 149, March 21, 1960.
- 53) Watkinson, A.P., Louis, L. and Brent, R., "Scaling of Enhanced Heat Exchanger Tubes". The Canadian Journal of Chemical Engineering, Vol. 52, p. 558, October 1974.
- 54) Epstein, N. and Sandhu, K., "Effect of Uniform Fouling Deposit on Total Efficiency of Extended Heat Transfer Surfaces". In: Sixth International Heat Transfer Conference, Vol. 4, p. 397, Toronto, Ontario, August 1978.

- 55) "Standard Methods of Chemical Analysis", Sixth Edition, Vol. 2, p. 2399. Welcher, F.J. (Editor). D. Van Nostrand Company Incorporated, Princeton, N.J., 1963.
- 56) "Standard Methods for the Examination of Water and Wastewater", Twelfth Edition, APHA, AWWA, WPCF, 1965.
- 57) "CRC Handbook of Chemistry and Physics", 57th Edition, Weast, R.C. (Editor), CRC Press Incorporated, Cleveland, Ohio, 1976-1977.

Appendix I. CALIBRATION OF THERMOCOUPLES AND ROTAMETERS

Table I.1 Calibration Table* for Thermocouples (57)

(Electromotive Force in Absolute Millivolts. Temperatures in Degrees C (Int. 1945). Reference Junctions at 0° C.)										
Ψ.	0		1	1 3	1	5	8	17	1 3	9
-190 -180 -170 -160 -180	-7.66 -7.40 -7.12 -6.62 -6.50	-7.69 -7.43 -7.16 -6.85 -6.53	-7.71 -7.46 -7.18 -6.88 -6.56	-7.73 -7.49 -7.21 -6.91 -6.60	-7.76 -7.51 -7.24 -6.94 -6.63	-7.78 -7.54 -7.27 -6.97 -6.66	-7.56 -7.20 -7.00 -6.69	-7.59 -7.32 -7.03 -8.72	-7.61 -7.35 -7.08 -6.76	-7 84 -7.38 -7.09 -8.79
140	- 8.18	-6.19	-6.22	6.28	-6.29	-6.33	-6.36	-6.40	-6.43	- 6.46
130	- 5.80	-5.84	-5.87	5.91	-5.94	-5.98	-6 01	-6.05	-6.08	- 6.12
120	- 5.42	-5.46	-5.50	-5.54	-5.58	-5.61	-5.65	-5.69	-5.72	- 5.76
110	- 5.03	-5.07	-5.11	-5.15	-5.19	-5.23	-5.27	-5.31	-5.35	- 5.38
100	- 4.63	-4.67	-4.71	-4.75	-4.79	-4.83	-4.87	-4.91	-4.95	- 4.99
- 90 - 80 - 10 - 150	-4.21 -3.78 -3.34 -2.89 -2.43	-4.25 -3.82 -3.38 -2.94 -2.48	-4.30 -3.87 -3.43 -2.98 -2.52	-4.34 -3.91 -3.47 -3 03 -2.67	-4.38 -3.96 -3.52 -3.07 -2.62	-4 42 -4,00 -3,56 -3,12 -2,66	-4.46 -4.04 -3.60 -3.16 -2.71	-4.50 -4 08 -3.65 -3 21 -2.75	-4.55 -4 13 -3.69 -3.25 -2.80	-4.59 -4.17 -3.74 -3.30 -2.84
- 40	-1.98	-2.01	-2.06	-2.10	-2.15	-2.20	-2.24	-2.29	-2 34	-2.38
- 30	-1.49	-1.53	-1.58	-1.63	-1.67	-1.72	-1.77	-1.82	-1.87	-1.91
- 20	-1.00	-1 04	-1.09	-1.14	-1.19	-1.24	-1.29	-1.34	-1.39	-1.43
- 10	-0.50	-0.55	-0.60	-0.65	-0.70	-0.75	-0.80	-0.85	-0 90	-0.95
(-)0	0.00	-0.05	-0.10	-0.15	-0.20	-0.25	-0.30	-0.35	-0.40	-0.45
(+)0 10 20 80 60	0.00 0.50 1.02 1.54 2.06 2.58	0.05 0.56 1.07 1.59 2.11 2.64	0.10 0.61 1.12 1.64 2.16 2.69	0.15 0.66 1.17 1.69 2.22 2.74	0.20 0.71 1.22 1.74 2.27 2.80	0.25 0.76 1.28 1.80 2.32 2.65	0.30 0.81 1.33 1.85 2.37 2.90	0.35 0.86 1.38 1.90 2.42 2.96	0.40 0.91 1.43 1.95 2.48 3.01	0.45 0.97 1.48 2.00 2.53 3.06
60 70 80 90	3.11 3 65 4.19 4.73 5.27	3.17 3.70 4.24 4.78 5.32	3.22 3.76 4.29 4.83 5.38	3.27 3.81 4.35 4.89 5.43	3.33 3.80 4.40 4.94 5.48	3.38 3.92 4.40 5.00 5.54	3.43 3.97 4.51 5.06 5.69	3.49 4.02 4.56 5.10 5.65	3.54 4.08 4.62 5.16 5.70	3.60 4.13 4.67 5.21 5.76
110 120 130 140	5.81 6.36 6.90 7.45 8.00	5.66 6.41 6.96 7.51 8.08	5.92 6.47 7.01 7.56 8.12	5.97 6.52 7.07 7.62 8.17	6.03 6.58 7.12 7.67 8.23	6.08 6.63 7.18 7.73 8.28	6.14 6.68 7.23 7.78 8.34	6.19 6.74 7.29 7,84 8,39	6.25 6.79 7.34 7.89 8.45	6.30 6.85 7.40 7.95 8.50
280	8.56	8.61	8,67	8.72	8.78	8.84	8.89	8.95	9.00	9.06
170	9.11	9.17	9,22	9.28	9.33	9.39	9.44	9.50	9.56	9.61
180	9.67	9.72	9,78	9.83	9.89	9.95	10.00	10.06	10.11	10.17
190	10.21	10.28	10,34	10.39	10.45	10.50	10.55	10.61	10.67	10.72
200	10.78	10.84	10,89	10.95	31.00	11.06	11.12	11.17	11.23	11.28
210	11.34	11.39	11.45	11.50	11.56	11.62	11.67	11.73	11.78	11.84
220	11.89	11.95	12.00	12.06	12.12	12.17	12.23	12.28	12.34	12.39
230	12.45	12.50	12.56	12.62	12.67	12.73	12.78	12.84	12.89	12.95
240	13.01	13.06	13.12	13,17	13.23	13.28	13.34	13.40	13.45	13.81
250	13.56	13.62	13.67	13.73	13.78	13.84	13.89	13.95	14.00	14.06
280	14.13	14.17	14, 13	14.28	14.34	14.39	14.45	14.50	14.56	14.61
270	14.67	14.72	14, 78	14.83	14. 19	14.94	15.00	15.06	15.11	15.17
280	15.22	15.28	18, 33	15.39	13.44	15.50	15.55	15.61	15.66	15.72
290	15.77	15.83	15, 89	15.94	16.00	16.05	16.11	16.16	16.22	16.27
300	16.33	16.38	16, 44	16.49	16.55	16.60	16.66	16.71	16.77	16 82
810	10.63	16. <i>4</i> 3	16.99	17.04	17.10	17.15	17.21	17.26	17.31	17.37
820	17.43	17.48	17.54	17.60	17.65	17.71	17.76	17.82	17.87	17.93
830	17.98	18.04	18.09	18.15	18.20	18.26	18.32	18.37	16.43	18.48
840	18.54	18.59	18.65	18.70	18.76	18.81	18.87	18.92	18.98	19.03
860	19.09	19.14	19.20	19.25	19.31	19.37	19.42	19.48	19.53	19.59
360	19.64	19,70	19.75	19.81	19.88	19.92	19.97	20.03	20.08	20.14
270	20.20	20.25	20.31	20.36	20.42	20.47	20.53	20.58	20.64	20.69
380	20.75	20.80	20.86	20.91	20.97	21.02	21.08	21.13	21.19	21.24
390	21.30	21.35	21.41	21.45	21.52	21.57	21.63	21.68	21.74	21.79
400	21.85	21.90	21.96	22.02	22.07	22.13	27.18	22.24	22.29	22.35
410	22.40	22.46	22.51	22.57	22.62	22.68	22,73	22.79	22.84	22.90
420	22.95	23.01	23.06	\$3.12	23.17	13.23	23,28	23.34	23.39	23 45
430	23.50	23.56	23.61	\$3.67	23.72	23.78	23,83	23.89	23.94	24.00
440	24.06	24.11	24.17	\$4.22	24.28	24.33	24,39	24.44	24.50	24.55
450	24.61	24.56	34.72	\$4.77	24.83	24.88	24,94	25.00	25.05	25.11
450	25.18	25.22	25.27	25.33	25,38	23.44	25.49	25.55	25 60	25 66
470	25.72	25.77	25.83	25.88	25,94	25.99	26.05	26.10	26.16	26.22
480	26.27	26.33	26.38	26.44	26,49	26.55	26.61	26.66	26 72	26 77
490	26.83	26.89	26.94	27.00	27,05	27.11	27.17	27.22	27.28	27.33
500	27.39	27.45	27.50	27.56	27,61	27.67	27.73	27.78	27.84	27.90
\$10 \$20 \$30 \$40 \$50	27.95 28.52 29.05 29.65 30.22	28.01 28.57 29.14 29.71 30.28	28.07 28.63 29.20 29.26 30.34	28.12 28.69 29.25 29.82 30.39	28.18 28.74 29.31 29.88 30.45	28.23 28.80 29.37 29.94 80.51	28,29 28,86 29,42 29,99 30,57	28, 35 28, 91 29, 48 30, 05 30, 62 of 1948	25,40 28,97 29,54 30 11 20,88	26.46 29 02 29 59 30 16 30 74

* Used in Runs 1 to 8.

Location of The Thermocouple	ر °C)	b (°C/mv)	(°C ² /mv)	Range (°C)	Run
Tank out	-1.67865	21.65324	-0.71004	25-50	9-38
Exch in	0.25190	19.79763	-0.29185	25-50	9-21
	0.37845	19.10536	0.0	20-40	22-38
Exch out	2.40695 0.20770 0.15810 0.34057	17.71571 19.34461 19.27142 19.12506	0.24891 0.0 0.0 0.0 0.0	25-50 20-40 20-40 20-40	9-21 22 23-26 27-38
Steam in	9.60230	15.75524	0.18182	100-130	9-21
	0.65260	18.80879	0.0	90-130	22-38
Steam out	9.97206	15.60082	0.18964	100-130	9-21
	0.39281	18.85247	0.0	90-130	22-38

Table I.2. CONSTANTS CORRESPONDING TO THE CALIBRATION EQUATION* FOR THERMOCOUPLES

* $T = av^2 + bv + c$ (T in °C, v in milivolts). Exch = exchanger.



Figure I-1. CALIBRATION CURVE FOR LARGE ROTAMETER


Figure I-2. CALIBRATION CURVE FOR SMALL ROTAMETER

Appendix II. SAMPLE CALCULATIONS

II.1 Determination of Heat Flow Rate

The rate of heat flow (q) to the recirculating water was calculated by:

$$q = W_c c_p \Delta T_c$$
(II-1)

where m, c_p and ΔT are the mass flow rate, the specific heat and the temperature change of the water, respectively. Subscript c is denoted to the cold stream (water). ΔT_c was calculated using the calibration table or calibration equations shown in Appendix I. W_c was determined using the volumetric flow rate (V) as follows:

$$W_{c} = \rho V$$
 (II-2)

Physical properties (c_p and ρ) of water were evaluated at the bulk temperature (T_b) using:

$$\rho(kg/m^3) = -0.3269 \ \bar{T}_{\rm b} + 1005.4$$
 (II-3)

$$c_p (k /kg^{\circ}C) = -0.00108 T_b + 4.1818$$
 (II-4)

where \overline{T}_{b} is in degrees centigrade.

The outside nominal area of the tubes were used as a reference area in calculation of the overall heat transfer coefficient

$$A_{O} = \pi D_{O}L \qquad (II-5)$$

where D_0 is the outside diameter (19.1 mm) and L is the length (1.33 m) of the tubes in contact with the water.

II.3 Determination of Logarithmic Mean Temperature Difference

Following the evaluation of the inlet and outlet temperatures of both streams via either the calibration table or equations, the logarithmic mean temperature difference (LMTD) was calculated using:

LMTD =
$$\frac{T_{c_{out}} - T_{c_{in}}}{\ln (T_{h} - T_{c_{in}})/(T_{h} - T_{c_{out}})}$$
(II-6)

where T is the temperature and subscripts c, h, in and out are denoted to the cold, the hot, the incoming and the outgoing streams. Since the steam (hot stream) did not undergo any sub-cooling, the average of the inlet and outlet temperatures, (\bar{T}_h) of that stream, which were slightly different due to the experimental errors, was used in the determination of LMTD.

II.4 Determination of Overall Heat Transfer Coefficient

The following equation was applied to compute the overall heat transfer coefficient based on the outside nominal area:

$$U_{0} = \frac{q}{A_{0}(LMTD)}$$
(II-7)

II.5 Determination of the Fouling Resistance

The change in the magnitude of the overall heat transfer coefficient over the length of the run was due to the fouling resistance (R_{f}) which was calculated through:

$$R_{f} = \frac{1}{U_{d}} - \frac{1}{U_{c}}$$
(II-8)

where U_c and U_d are the clean and the dirty overall heat transfer coefficients evaluated, respectively, at the start and end of a run. It should be noted that the fouling resistance of the finned tube was calculated based on the nominal area (bare-tube) outside diameter (see Appendix II.2).

To relate the mass and thickness of the deposit on the finned tube to that of the plain tube, the effective area (A_{eff}) should be taken into consideration. The relationship between the fouling resistance based on the prime area (R $_{\rm f_1}$) and the one based on the effective area (R $_{\rm f_2}$) can be expressed by:

$$R_{f_2} = \frac{A_{eff}}{A_o} R_{f_1}$$
(II-9)

The unit thermal resistance (R_f) , the mass per unit surface area (m_f) and the thickness of the deposit (x_f) , are inter-related as:

$$\mathbf{m}_{\mathbf{f}} = \mathbf{\rho}_{\mathbf{f}} \mathbf{x}_{\mathbf{f}} = \mathbf{\rho}_{\mathbf{f}} \mathbf{k}_{\mathbf{f}} \mathbf{R}_{\mathbf{f}}$$
(II-10)

Therefore, the total mass of the deposit on the finned tube would be

$${}^{m}t_{1} = {}^{\rho}f^{k}f^{R}f_{1} {}^{A}prime \qquad (II-11)$$

or

$${}^{m}t_{2} = {}^{\rho}f^{k}f^{R}f_{2} {}^{A}eff \qquad (II-12)$$

Thus, the mass of the deposit per unit prime area (m_{f_1}) and per unit total area (m_{f_2}) would be

$$m_{f_1} = \rho_f k_f R_{f_1}$$
(II-13)

$$m_{f_2} = \frac{\rho_f k_f R_{f_2} A_{eff}}{A_t}$$
(II-14)

$$\frac{{}^{m}f_{2}}{{}^{m}f_{1}} = \frac{{}^{R}f_{2}^{A}eff}{{}^{R}f_{1}^{A}t}$$
(II-15)

Considering equation II-10 and substituting equation II-9 into equation II-15 would yield:

$$m_{f_2}/m_{f_1} = x_{f_2}/x_{f_1} = A_{eff}^2/A_oA_t$$
 (II-16)

 A_{eff} can be evaluated using the fin efficiency (n_f) through:

$$m = \left(\frac{h_c^P}{k_{fin}^A f}\right)$$
(II-17)

$$n_{f} = \frac{\tanh(ml)}{ml}$$
(II-18)

$$A_{eff} = A_t^{\eta} t = (A_t - A_f) + A_f^{\eta} f$$
 (II-19)

where, 1, P, A_f and k_{fin} are the height, the perimeter, the surface area and the conductive heat transfer coefficient of the fin. The convective heat transfer coefficient h_c was approximately 2.8 kW and evaluated by:

$$Nu = \frac{h_c D}{k_1} = 0.023 \text{ Re}_{film}^{0.8} \text{ Pr}_{film}^{0.333}$$
(II-20)

A sample calculation is given in Section II-11.

II.6 Determination of the Wall Temperature

The average wall temperature throughout a run was evaluated using the radial rate of heat flow equations for concentric cylinders in the steady state condition. Wall temperatures were required for model calculations only.

~.

$$q = \pi D_{i}Lh_{i}(T_{save} - T_{w_{i}}) = \frac{2\pi Lk_{w}}{Tn(D_{o}/D_{i})}(T_{w_{i}} - T_{w}) = \pi D_{o}Lh_{o}(T_{v} - T_{o})$$

$$i_{ave} ave ave ave ave ave (II-21)$$

where T is the temperature and subscripts s, w_i and w_o are denoted to the steam, inside and outside tube wall, respectively. Assuming that the inside wall and the steam temperatures are equal, the following equations can be written:

$$q = \frac{2\pi Lk_{W}}{\ln (D_{0}/D_{i})} (\bar{T}_{save} - T_{w_{0}})$$
(II-22)

$$T_{w_{o_{ave}}} = T_{s_{ave}} - \frac{q \ln(D_{o}/D_{i})}{2\pi L k_{w}}$$
 (II-23)

II.7 Determination of the Equivalent Diameter

The equivalent diameter (D_e) was required for the subsequent determination of the Reynolds number. The magnitude of this parameter (D_e) should be evaluated by different means considering the geometry of the tubes.

In the case of plain tubes, the following equation was applied:

$$D_e = D_{s_i} - D_{t_o}$$
(II-24)

where D_{s_i} and D_{t_o} are, respectively, the inside diameter of the shell and the outside diameter of the tube.

The equivalent diameter of the enhanced tube was calculated using:

$$D_{e} = 4 \frac{\text{Net Free Area (NFA)}}{\text{Total Wetted Perimeter (TWP)}}$$
(II-25)

where

NFA = Inside cross sectional area of the shell -Outside cross sectional area of the tube -Cross sectional area of the fins

$$= \frac{\pi}{4} \left(D_{s_{1}}^{2} - D_{t_{0}}^{2} \right) - 12 t 1$$
 (II-26)

and

$$TWP = \pi (D_{s_{1}} + D_{t_{0}}) + 12(2)$$
 (II-27)

where t and 1 are, respectively, the thickness and the height of the fins.

II.8 Determination of the Reynolds Number

Reynolds numbers were calculated based on both the bulk (\overline{T}_{b}) ave and the film (T_{f}) temperatures using the following equation:

$$Re = \rho v D_e / \mu = \rho V D_e / S_\mu \qquad (II-28)$$

where ρ , v, V and μ are, respectively, the density, the velocity, the volumetric flow rate and the viscosity of the water passing through the annular cross sectional area, S.

To calculate the physical properties (ρ , μ) of the water at the average bulk temperature, linear regression was applied in the 20°-30°C temperature region yielding to equation II-3 in case of density, and:

$$\mu$$
 (kg/m.s) = -0.1731 x 10⁻⁴ T + 1.3177 x 10⁻³ (II-29)

where $\overline{\mathbf{T}}_{\mathbf{b}_{ave}}$ is in degrees centigrade.

For the evaluation of the density and viscosity at the average film temperature, the following relationships were applied:

$$\rho$$
 (kg/m³) = 1181.32 - 0.593 T_f (II-30)

$$\mu(kg/m.s.) = 0.1/[2.148 ((T_f - 281.435) + (8078.4 + (T_f - 281.435)^2)^{0.5}) - 120.]$$
(II-31)

where ${\sf T}_{\sf f}$ is in degrees kelvin and evaluated by:

$$T_{f} = (\overline{T}_{b_{ave}} + \overline{T}_{w_{o_{ave}}})/2$$
(II-32)

II.9 Determination of Water Quality Parameters

Total alkalinity (T.A.) was evaluated using the following expression:

$$T.A. = \frac{B' \times N \times 5000}{V}$$
(II-33)

where B', N and v' are the volume of the titrant (HCl), normality of the titrant and volume of the sample, respectively.

Hardness (C_a^{++}) of the water sample was evaluated using:

$$C_{a}^{++} = \frac{A' \times B'' \times 1000}{V'}$$
(II-34)

where A' and v' are volumes of the titrant (EDTA) and of the sample, respectively; B'' is the weight (mg) of $CaCO_3$ equivalent to 1 ml of EDTA.

The magnitude of total dissolved solids (TDS) was required for model calculations and as explained in Section 4.6, can be approximated by:

TDS =
$$1.06$$
 T.A. + 1.64 Ca⁺⁺ (II-35)

The values of total alkalinity and hardness, obtained from equations II-21 and II-22, were averaged throughout a run and used in equation II-23 for evaluation of average total dissolved solids.

II.10 Numerical Example Using Run 26

Run 26 corresponds to the test on the finned tube having rotameter setting of 50 (V = $39.83 \times 10^{-5} \text{ m}^3/\text{s}$). Using the calibration equations (Table I.2), the temperature readings for the start of the run are:

$$T_{c_{in}} = 25.18^{\circ}C$$

 $T_{c_{out}} = 32.80^{\circ}C$
 $\overline{T}_{s} = 100.07^{\circ}C$

$$\bar{T}_{b} = 28.99^{\circ}C$$

. .

Applying equations II-3 and II-4:

$$p_{\rm b} = 995.969 \, {\rm kg/m}^3$$

c_{pc} = 4.1787 kJ/kg°C

Using equations II-1 and II-2:

$$q = (39.83 \times 10^{-5}) (995.969) (4.1787) (32.80-25.18) = 12.68 kW$$

Substituting the temperature values in equation II-6 yields:

LMTD =
$$\frac{32.80 - 25.18}{1n \frac{100.07 - 25.18}{100.07 - 32.80}}$$
 = 70.97°C

Applying equation II-5:

$$A_0 = \pi(0.0191) (1.33) = 0.0796 \text{ m}^2$$

Therefore, by plugging the above numerical values in equation II.7, the clean overall heat transfer coefficient would be obtained:

$$U_{o_c} = \frac{12.68}{(0.07976) (70.97)} = 2.24 \text{ kW/m}^2\text{K}$$

The dirty overall heat transfer coefficient would be evaluated using the temperature readings for the end of the run followed by the above procedure:

$$U_{o_{d}} = \frac{(39.83 \times 10^{-5}) (995.44) (4.1785) (34.35-26.84)}{(0.07976) (34.35-26.84)} = 2.02 \text{ kW/m}^{2}\text{K}$$

$$\frac{(0.07976) (34.35-26.84)}{1n (\frac{107.78-26.86}{107.78-34.35})}$$

Then, the fouling resistance can be calculated according to equation II-8

$$R_f = \frac{1}{2.02} - \frac{1}{2.24} = 0.049 \text{ m}^2 \text{K/kW}$$

The average steam temperature and the heat flow rate throughout a run were, respectively, 104.84°C and 16.7 kW. Thus, according to equation II-23

$$T_{w_{o_{ave}}} = 104.84 - \frac{12.7 \ln(19.1/15.9)}{2\pi(44.99)(1.33)} = 95.95^{\circ}C$$

Having $\overline{T}_{b_{ave}} = 29.70^{\circ}C$ yields:

$$T_f = \frac{95.95 + 29.70}{2} = 62.32^{\circ}C$$

$$p_b = 995.74 \text{ kg/m}^3$$

 $\mu_b = 0.0008051 \text{ kg/m.s}$
 $p_f = 982.081 \text{ kg/m}^3$
 $\mu_f = 0.0004484 \text{ kg/m.s}$

Using equations II-25, II-26 and II-27 yields:

$$D_{e} = 4 \frac{\pi (0.037^{2} - 0.0191^{2})/4 - 12(0.006)(0.0005)}{\pi (0.037 + 0.0191) + 12(2)(0.006)} = 0.00869 \text{ m}$$

Therefore, the Reynolds numbers can be calculated knowing $S = NFA = 69.528 \times 10^{-5} m^2$ and using equation II-28:

$$\operatorname{Re}_{b} = \frac{(995.74) (0.0239) (0.0869)}{(69.528 \times 10^{-5}) (0.0008035)} = 6170$$

$$\operatorname{Re}_{f} = \frac{(982.081) \ (0.0239) \ (0.0869)}{(69.528 \times 10^{-5}) \ (0.0004484)} = 10652$$

50 ml of the water sample taken at the start of the run was titrated with 19.1 ml of 0.02 N HCl. Thus, according to equation II.33:

T.A. =
$$\frac{(19.1) (0.02) (50000)}{50}$$
 = 382 ppm CaCO₃

For the hardness determination, 25 ml of the sample was titrated with 0.01 M EDTA (1 mg $CaCO_3$ per 1 ml of EDTA). 16.8 ml of the titrant was used until the occurrence of the indicator colour change. Equation II-34 yields:

$$C_a^{++} = \frac{(16.8) (1) (1000)}{25} = 672 \text{ ppm } CaCO_3$$

TDS = 1.06 (361) + 1.64 (648) = 1445 mg/liter

II.11 Numerical Determination of Fin Efficiency and Deposit Thickness

For a 1.2 m long and 6 mm high and 0.5 mm wide fin, the fin efficiency was calculated through equations II-17 and II-18:

$$m = \left[\frac{(2.8 \times 10^3) (2) (1.2)}{(44.99) (1.2) (0.0005)}\right]^{0.5} = 498.94$$

$$n_f = \frac{\tanh [(498.94) (0.006)]}{(498.94) (0.006)} = 0.33$$

Therefore, the effective area and the total efficiency of the finned tube would be evaluated through equation II-19 having

$$A_f = 2 (12) (1.2) (0.006) = 0.1728 m^2$$

and

 $A_{+} = 0.0796 - 12 (1.2) (0.0005) + 0.1728 = 0.2452 m^{2}$

$$A_{eff} = (0.2452 - 0.1728) + 0.1728 (0.33) = 0.1294 \text{ m}^2$$

and

$$^{\eta}t = \frac{0.1294}{0.2452} = 0.53$$

Therefore, using equation II-16:

$$\frac{{}^{m}f_{2}}{{}^{m}f_{1}} = \frac{{}^{x}f_{2}}{{}^{x}f_{1}} = \frac{(0.1294)^{2}}{(0.0796)(0.2452)} = 0.86$$

Appendix III. COMPUTER PROGRAMS

Four computer programs were used to investigate the results of each run. The first program (Figure III.1) converts the raw datalogger data to temperatures and computes the rates of heat flow, the overall heat transfer coefficients and the fouling resistances. The second one contains three separate subprograms, each of which fits the fouling resistance data to a linear (Figure III.2), an asymptotic (Figure III.3) or a fourth degree polynomial (Figure III.4) function. The third one plots both the fouling resistance data and the best fit with respect to time (Figure III.5). The last program (Figure III.6) calculates the predicted fouling rates based on the Hasson's ionic diffusion model.

Figure III-1. PROGRAM TO EVALUATE OVERALL HEAT TRANSFER COEFFICIENTS AND FOULING RESISTANCES

```
REAL LMTD
    INTEGER RUN
    DIMENSION TIME(200),R1(200),R2(200),R3(200),R4(200),EI(200),E0(200)
    DIMENSION DT(200), SD(200), SI(200), TS(200), TW(200), LMTD(200), DE(200)
    DIMENSION HC(200), QA(200), UA(200), UI(200), RE(200)
    TT=O.
    SS=O.
    AA=O.
    READ (5.1) RUN, GG, VE, N
    FORMAT (15.2F10.5,15)
    FORMAT (F6.0,4F6.3)
2
    DO 10 I=1.N
    READ (5.2) TIME(I),R1(I),R2(I),R3(I),R4(I)
    EI(I)=19.10536289*R1(I)+0.3784450408
    ED(I)=19.125505776*R2(I)+0.3405731389
    DT(I) = EO(I) - EI(I)
    SO(I)=18.8524736*R3(I)+0.39280770
    SI(I)=18.80878684*R4(I)+0.6525958763
    TS(I) = (SD(I) + SI(I))/2.
    TW(I) = (EO(I) + EI(I))/2.
    TT=TT+TW(I)
    SS=SS+TS(I)
    LMTD(I)=DT(I)/ALOG((TS(I)-EI(I))/(TS(I)-EO(I)))
    DE(1)=(-0.326893)*TW(1)+1005.445464
    HC(I) = (-0.000108) * TW(I) + 4.181817
    QA(I) = (DT(I) * DE(I) * HC(I) * VE) / 60.
    UA(I)=QA(I)/(0.07976*LMTD(I))
    UI(I)=1/UA(I)
    RE(I)=UI(I)-GG
    AA=AA+QA(I)
10
    CONTINUE
    WRITE (6,3)
    FORMAT (
                                                           LMTD
з
                  ΕI
                        FO
                               DT
                                     SO
                                             ST.
                                                    TS
                                                                  TIME
                                                                          ο
                                                                               ٢.
             R ()
   1' U
4
    FORMAT (1X, 3F6.2, 4F7.2, F6.0, 2F6.2, F7.3)
    Z=0.
    DO 7 I=1,N
    WRITE (6.4) EI(I).EO(I).DT(I).SO(I).SI(I).TS(I).LMTD(I).TIME(I).QA(I).
   1UA(I), RE(I)
    Z=Z+1.
7
    CONTINUE
    WA=TT/Z
    SA=SS/Z
    FA=AA/Z
    WRITE (6,5)
    FORMAT ( '
5
                    TW(a∨e)
                              TS(ave) Q(ave) ')
    FORMAT (1X,2F10.2,F8.2)
8
    WRITE (6,8) WA, SA, FA
    STOP
    END
```

Figure III-2. PROGRAM TO LINEARLY FIT FOULING RESISTANCES

```
DIMENSION X(200), Y(200), YF(200), WT(200), E1(6), E2(6), P(6)
      COMMON M
З
      READ(5,1) N.M.NI
      FORMAT(315)
1
      READ(5,2) EPS
2
      FORMAT(F10.5)
12
      FORMAT(6F10.5)
      FORMAT(F10.1, F10.4)
11
      READ (5,12) (P(I), I=1,M)
      DD 10 I=1.N
10
      READ(5, 11) X(I), Y(I)
      EXTERNAL AUX
      CALL LQF(X,Y,YF,WT,E1,E2,P,O.O,N,M,NI,ND,EPS,AUX)
      IF (ND.NE.1) STOP
      WRITE (6,4)
4
      FORMAT( ' ESTIMATES OF ROOT MEAN SQUARE TOTAL ERROR IN THE PARAMETERS ')
      WRITE (6,5) (E2(I) ,I=1,M)
5
      FDRMAT (1X,8G15.5)
      WRITE (6,6)
6
      FORMAT(' VALUES OF X VALUES OF Y FITTED VALUES OF Y')
      DO 7 I=1,N
7
      WRITE (6.5) X(I),Y(I),YF(I)
WRITE (6.8) P(1),P(2)
8
      FORMAT(/, ' a= ',G12.5, ' b= ',G12.5)
30
      STOP
      END
      FUNCTION AUX(P,D,X,L)
      DIMENSION P(6),D(6)
      COMMON M
      D(1)=1.
      D(2)=X
      AUX = P(1) + X * (P(2))
      RETURN
      END
```

Figure III-3. PROGRAM TO ASYMPTOTICALLY FIT FOULING RESISTANCES

.

	DIMENSION X(200), Y(200), YF(200), WT(200), E1(5), E2(5), P(5)
3	READ(5.1) N.M.NT
1	FORMAT(315)
•	READ(5, 2) EPS
2	FORMAT(F10.5)
12	EDRMAT (3E10-5)
11	FDRMAT(F10, 1, F10, 4)
•••	READ (5, 12) (P(T), T=1 M)
10	READ(5, 11) X(1), Y(1)
	EXTERNAL AUX
	CALL LOF(X, Y, YE, WT, E1, E2, P.O.O. N. M. NI, ND, EPS, AUX)
	IF (ND NF 1) GD TD 3
	WRITE (6.4)
4	FORMAT(' ESTIMATES OF ROOT MEAN SQUARE TOTAL ERROR IN THE
	1 PARAMETERS ()
	WRITE (6.5) (E2(I) .I=1.M)
5	FORMAT (1X,8G15.5)
	WRITE (6,6)
6	FORMAT(' VALUES OF X VALUES OF Y FITTED VALUES OF Y')
	DO 7 I=1,N
7	WRITE (6,5) X(I),Y(I),YF(I)
	WRITE (6.8) P(1),P(2),P(3)
8	FDRMAT(/,'a= ',G12.5,'b= ',G12.5,'c= ',G12.5)
30	STOP
	END
	FUNCTION AUX(P,D,X,L)
	DIMENSION P(3),D(3)
	COMMON M
	D(1)=1.
	D(2)=1EXP(-P(3)*X)
	D(3)=P(2)*X*EXP(-P(3)*X)
	AUX=P(1)+P(2)*(1-EXP(-P(3)*X))
	RETURN
	END

Figure III-4. PROGRAM TO FIT FOULING RESISTANCES TO A POLYNOMIAL FUNCTION

.

/

```
DIMENSION X(200), Y(200), YF(200), WT(200), E1(6), E2(6), P(6)
      COMMON M
З
      READ(5,1) N.M.NI
      FORMAT(315)
1
      READ(5,2) EPS
      FORMAT(F10.5)
2
      FORMAT(6F10.5)
12
      FORMAT(F10.1, F10.4)
11
      READ (5,12) (P(I), I=1,M)
      DD 10 I=1,N
      READ(5, 11) \times (I), Y(I)
10
      EXTERNAL AUX
      CALL LQF(X,Y,YF,WT,E1,E2,P,O.O,N,M,NI,ND,EPS,AUX)
      IF (ND.NE.1) STOP
      WRITE (6,4)
      FORMAT(' ESTIMATES OF ROOT MEAN SQUARE TOTAL ERROR IN THE PARAMETERS ')
4
      WRITE (6,5) (E2(I) ,I=1,M)
5
      FORMAT (1X,8G15.5)
      WRITE (6,6)
6
      FORMAT(' VALUES OF X VALUES OF Y FITTED VALUES OF Y')
      DO 7 I=1,N
7
      WRITE (6,5) X(I), Y(I), YF(I)
      WRITE (6,8) P(1),P(2),P(3),P(4),P(5)
FORMAT(/,' a = ',G12.5,' b = ',G12.5,' c = ',G12.5,
8
   #/,' d= ',G12.5,' e= ',G12.5 )
30
      STOP
      END
      FUNCTION AUX(P,D,X,L)
      DIMENSION P(6),D(6)
      COMMON M
      D(1)=1.
      D(2)=X
      D(3)=X*X
      D(4)=X*X*X
      D(5)=X*X*X*X
      AUX=P(1)+X*(P(2)+X*(P(3)+X*(P(4)+X*(P(5)))))
      RETURN
      END
```

L

Figure III-5. PROGRAM TO PLOT THE FOULING RESISTANCE DATA AND THE BEST FIT

```
INTEGER N
   DIMENSION X(200), Y(200), Z(200), XX(200)
   READ (5,3) A.B.C.D.E
 3 FORMAT (5E12.9)
   READ (5,2) N, RUN, Q, V, RE, TA, CA, PH
 2 FORMAT (15, F5.0, F5.1, F6.3, F7.0, 2F5.0, F5.2)
   READ (5.4) M.K.W
 4 FORMAT (215, F5.0)
   READ (5,1) (X(I),Y(I),I=1,N)
 1 FORMAT (F10.1, F10.4)
   CALL AXIS (2.,2., 'Rf (M**2.K/kW)', 14,6.,90.0,+0.04,0.04)
   CALL PLOT (2.,2.,3)
CALL PLOT (2.,1.9,3)
   CALL PLOT (2.,2.,2)
   T=3.
   DO 5 I=1,7
CALL PLOT (T,2.,2)
   CALL PLOT (T, 1.9,2)
   CALL PLOT (T,2.,3)
   T=T+1
 5 CONTINUE
   CALL PLOT (2.,2.,3)
   XS=O.
   CALL NUMBER (1.975,1.75,0.1,XS,0,-1)
   XS=XS+600.
   CALL NUMBER (2.875, 1.75, .1, XS, 0, -1)
   R=3.825
   DD 30 I=1,6
   XS=XS+600
   CALL NUMBER (R. 1.75,0.1,XS,0,-1)
   R=R+1
30 CONTINUE
   CALL PSYM (4.95,1.55,0.15,'TIME (min)',+0.,10)
   CALL PLOT (2.,8.,3)
   F=3.
   DD 40 I=1,7
   CALL PLOT (F.8.,2)
CALL PLOT (F,7.9,2)
   CALL PLOT (F.8.,3)
   F = F + 1
40 CONTINUE
   ₽=7.
   DO 50 I=1.6
   CALL PLOT (9.,P,2)
CALL PLOT (8.9,P,2)
   CALL PLOT (9.,P,3)
   P=P-1.
50 CONTINUE
   DD 20 I=1.N
   X(I)=2.+X(I)/600.
   Y(1)=2.+(Y(1)+.04)/0.04
   CALL SYMBOL (X(I),Y(I),0.08,0,0.,-1)
20 CONTINUE
   XX(1)=0.0
   IF (C.EQ.O.) GD TD 60
   IF (D.EQ.O.) GO TO 70
   DD 90 J=1,K
   Z(J) = A + XX(J) * (B + XX(J) * (C + XX(J) * (D + XX(J) * (E))))
   XX(J+1)=XX(J)+30.
90 CONTINUE
```

Figure III-5. PROGRAM TO PLOT THE FOULING RESISTANCE DATA AND THE BEST FIT (Continued)

```
DD 91 J=1,K
     XX(J)=2.+XX(J)/600.
     Z(J)=2.+(Z(J)+0.04)/0.04
 91 CONTINUE
     GD TD 100
 70 CONTINUE
     DO 80 J=1,K
     Z(J)=A+B*(1-EXP(-C*XX(J)))
     XX(J+1)=XX(J)+30.
 80 CONTINUE
     DO 81 J=1,K
     XX(J)=2.+XX(J)/600.
     Z(J)=2.+(Z(J)+0.04)/0.04
 81 CONTINUE
     GO TO 100
 60 CONTINUE
     DO 10 J=1,K
     Z(J) = A + B + X X(J)
     XX(J+1)=XX(J)+30.
 10 CONTINUE
     DD 11 J=1,K
     XX(J)=2.+XX(J)/600
     Z(J)=2.+(Z(J)+0.04)/0.04
 11 CONTINUE
100 CONTINUE
     CALL PLOT (XX(M),Z(M),3)
     DD 15 J=M.K
     CALL PLOT (XX(J),Z(J),2)
 15 CONTINUE
     WRITE (6,8) X(N),Z(M),Y(N),A,B
  8 FORMAT (/, 'X=',G12.5,'Z=',G12.5,'Y=',F9.5,'A=',G12.5,'B=',G12.5)
CALL SYMBOL (3.0.7.4,0.1,'RUN'.0.,3)
     CALL SYMBDL (3.0,7.25,0.1,'Q',0.,1)
     CALL SYMBOL (3.0,7.1,0.1, 'V',0.,1)
    CALL SYMBOL (3.0,6.95,0.1,'Re',0.,2)
CALL SYMBOL (3.0,6.8,0.1,'T.A.',0.,4)
CALL SYMBOL (3.,6.65,0.1,'Ca',0.,2)
     CALL SYMBOL (3.17,6.69,0.10, '++',0.,2)
    CALL SYMBOL (3.,6.5,0.1,'PH',0.,2)
CALL NUMBER (3.6,7.4,0.1,RUN,0.,-1)
    CALL NUMBER (3.6.7.25.0.1.0.0.,1)
    CALL NUMBER (3.6.7.1,0.1,V,0.,3)
    CALL NUMBER (3.6,6.95,0.1,RE,0.,-1)
    CALL NUMBER (3.6,6.8,0.1,TA,0.,-1)
    CALL NUMBER (3.6,6.65,0.1,CA,0.,-1)
    CALL NUMBER (3.6.6.5.0.1.PH.O..2)
CALL PSYM (4.15.7.25.0.1.'kW',+0..2)
    CALL PSYM (4.15,7.1,0.1, 'm/s',+0.,3)
CALL PSYM (4.15,6.8,0.1, 'mg/1',+0.,4)
CALL PSYM (4.15,6.65,0.1, 'mg/1',+0.,4)
    IF (W.EQ.O.) GD TD 120
     CALL PSYM (3.7,7.4,0.1, 'A',+0.,1)
120 CALL PLOTND
     STOP
```

END

Figure III-6. PROGRAM TO DETERMINE THE RATES PREDICTED BY THE HASSON'S IONIC DIFFUSION MODEL

```
/COMPILE
     IMPLICIT REAL (A-Z)
     INTEGER RUN
     INPUT UNITS OF TA, TDS ARE PPM, TEMP=DEG K, V=CM/SEC, D=CM
READ (5,99) RUN, V, D, TW, TB, TDS, CA, TA, PH
С
9
     FORMAT (12, F6.2, F6.3, F7.2, F6.2, F6.0, 2F5.0, F5.2)
     TW=TW+273.16
     TB=TB+273.16
     IF (RUN .EQ. O) GO TO 3
     TEMP=,5*(TW+TB)
     PRINT, 'RUN=', RUN
      FIND LIQUID DENSITY AND VISCOSITY AS A FUNCTION OF TEMP
С
      LIQUID DENSITY=P LIQUID VISCOSITY=U
C
     P=(1181.32-0.593*TEMP)/1000.
     U=1./(2.148*((TEMP-281.435)+(8078.4+(TEMP-281.435)**2)**.5)-120.0)
С
      FIND REYNOLDS NUMBER (RE) AND SCHMIDT NUMBER (SC)
     RE=V*D*P/U
     SC=U/(P*1.E-05)
     PRINT, 'P=',P, 'U=',U, 'RE=',RE, 'SC=',SC
      CALCULATE MASS TRANSFER COEFFICIENT KD
С
     KD=0.023*RE**(-0.17)*SC**(-0.666)*V
С
      CALCULATE KR- REACTION RATE CONSTANT
     KR=EXP(38.74-20700./(1.987*TEMP))
С
      CHANGE UNITS OF KR TO KR1 (CM/S)/(G CACO3/CM**3) TO (CM/S)/(GMOLECA/L))
     KR1=KR*2,497239*40,08/1000.
     KR1=100.*KR1
     DETERMINE IONIC STRENGTH OF THE SOLUTION I = (1.E-O3)*TDS/40.
С
С
      DETERMINE FI
     FI=(I**.5)/(1+I**.5)-.3*I
      CHANGE UNITS OF TA, TDS, AND CA (AS PPM OF CACO3) TO MOLES/L
С
     TA=TA/(1000.*100.08935)
     CA=CA/(1000.*100.08935)
     PRINT, 'I=', I, 'TA=', TA, 'TDS=', TDS
      FIND FM AND FD, ALSO K1 AND K2 AND HENCE K11 AND K21
С
     FM = 10. **(-0.51 * FI)
     FD=10.**(-2.04*FI)
с
      USE PH TO DETERMINE (H+) AND PKW TO FIND (OH-)
     H=(10.**(-1.*PH))/FM
     OH=((10.**(-4787.3/TEMP-7.1321*ALOG10(TEMP)-0.01037*TEMP+22.801))
    1/H)/FM
     PRINT, 'FM=', FM, 'FD=', FD
     K1=10.**(-17052/TEMP-215.21*ALDG10(TEMP)+0.12675*TEMP+545.56)
     K2=10.**(-2902.39/TEMP-0.02379*TEMP+6.498)
     PRINT, 'K1=', K1, 'K2=', K2, 'TEMP=', TEMP
     K11=K1/FM**2
     K21=K2/FD**2
     PRINT, 'K11=',K11, 'K21=',K21
NOW USE THE VALUES OF TA,H,OH,K11,K21 TO DETERMINE (HCO3),(CO3),(CO2)
С
     HCO3=(TA+H-OH)/((2.*K21/H)+1.)
     CO3=(TA+H-DH)/(2.+H/K21)
     CO2=(H*(TA+H-OH))/(K11+2.*K11*K21/H)
     PRINT, 'PH=', PH, 'H=', H
     PRINT, '(HCO3)=',HCO3,'(CO3)=',CO3,'(CO2)=',CO2
     PRINT, '(OH)=', OH, 'K11=', K11, 'K21=', K21
      NOW WE CALCULATE W USING CA++ CONC.
С
     KSPU=10.**(-0.01183*(TEMP-273)-8.03)
     KSP=KSPU/(FD*FD)
     PRINT, 'CA++=', CA, 'KD=', KD, 'KR=', KR, 'KSP1=', KSP
     W=(KD*CA/2.)*(1.+CO3/CA+KD/(KR1*CA))*(1.-(1.-((4.*CO3/CA)*(1.-KSP/
```

Figure III-6. PROGRAM TO DETERMINE THE RATES PREDICTED BY THE HASSON'S IONIC DIFFUSION MODEL (Continued)

```
1(CA*CO3)))/(1.+CO3/CA+KD/(KR1*CA))**2)**.5)
     A=1.-4.*K21*KR1*CA/(K11*KD)
     B=C02/CA+4.*K21*KR1*HC03/(K11*KD)+KSP*KR1/(KD*CA)
     C=K21*KR1*HCD3**2/(K11*KD*CA)-KSP*CD2*KR1/(CA**2*KD)
     CC=(1.+4.*A*C/B**2)**.5-1.
     VV=CC+1.
     WLPH= . 5* . 1*KD*CA*B*CC/A
     WP=KD*CO3*(1.-KSP/(CA*CO3))/(KD/(KR1*CA)+4.*CO3/HCO3+KSP/(CA*CO2))
С
     CORRECT UNITS OF W FROM CM/S * MOLE/L TO GM/CM**2-S
     W=(W/1000.)*100.08935
     WP=.1*WP
     PRINT, 'VV=', VV, 'WLPH=', WLPH, 'WHPH=', W, 'IN G/(CM**2-S)'
PRINT, 'W LPH APP=', WP
     WRITE(6,12)
  12 FORMAT(//)
З
      CONTINUE
     STOP
     END
/EXECUTE
```

Appendix IV. FOULING CURVES

The remainder of fouling curves with respect to time are shown in Figures IV.1 to IV.17.



Figure IV-1. RUN 1A FOULING RESISTANCE VERSUS TIME



Figure IV-2. RUN 2A FOULING RESISTANCE VERSUS TIME



Figure IV-3. RUN 3A FOULING RESISTANCE VERSUS TIME



Figure IV-4. RUN 3 FOULING RESISTANCE VERSUS TIME



Figure IV-5. RUN 5A FOULING RESISTANCE VERSUS TIME



Figure IV-6. RUN 5 FOULING RESISTANCE VERSUS TIME



Figure IV-7. RUN 6A FOULING RESISTANCE VERSUS TIME



Figure IV-8. RUN 21 FOULING RESISTANCE VERSUS TIME

- 200 -



Figure IV-9. RUN 22 FOULING RESISTANCE VERSUS TIME



Figure IV-10. RUN 24 FOULING RESISTANCE VERSUS TIME



Figure IV-11. RUN 25 FOULING RESISTANCE VERSUS TIME


Figure IV-12. RUN 28 FOULING RESISTANCE VERSUS TIME



Figure IV-13. RUN 29 FOULING RESISTANCE VERSUS TIME



Figure IV-14. RUN 30 FOULING RESISTANCE VERSUS TIME



Figure IV-15. RUN 32 FOULING RESISTANCE VERSUS TIME

•



Figure IV-16. RUN 33 FOULING RESISTANCE VERSUS TIME

.;`•



Figure IV-17. RUN 37 FOULING RESISTANCE VERSUS TIME