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DYNAMIC MASS TRANSFER OF STAINLESS STEEL IN
SODIUM UNDER HIGH-HEAT-FLUX CONDITIONS

AEC Research and Development Report

ATOMICS INTERNATIONAL

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Printed in the United States of America
Available from
Clearinghouse for Federal Scientific and Technical Information
National Bureau of Standards, U.S. Department of Commerce
Springfield, Virginia 22151
Price: Printed Copy \$3.00; Microfiche \$0.65

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H.C. 3.00; REV. 65

**DYNAMIC MASS TRANSFER OF STAINLESS STEEL IN
SODIUM UNDER HIGH-HEAT-FLUX CONDITIONS**

By

**J. HOPENFELD
D. DARLEY**

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**CONTRACT: AT(04-3)-701
ISSUED: JULY 15, 1967**

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ABSTRACT

Equations governing the corrosion rate of fuel cladding material in sodium-cooled reactors are presented. Using corrosion data obtained from various loop tests, attempts were made to evaluate the mass transfer coefficients which appear in the basic equations. It was found that available loop data are inadequate to provide the needed information, mainly because the local bulk liquid concentration of the corrosion products cannot be determined to any degree of certainty in these loops. A test loop facility is described where this difficulty is eliminated by setting up concentration gradients between the surface and the bulk liquid by means of high-heat-flux heaters.

Preliminary data is presented on the corrosion behavior of Type 304 stainless steel exposed to 1200°F sodium, with a 1250°F surface temperature. The results obtained demonstrate the feasibility of using the present method to study mass transfer mechanisms in sodium.

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I. INTRODUCTION

As part of the AEC fast reactor cladding program, work has been initiated at Atomics International to:

- 1) Design, fabricate, and operate dynamic sodium test facilities which simulate a fast reactor operating environment
- 2) Test, evaluate, and demonstrate performance capability of samples of candidate fast reactor cladding and structural materials in these dynamic sodium facilities.

During the course of this work it became increasingly apparent that complete simulation of the reactor environment in experimental loops is not possible; therefore, some extrapolation of the experimental data from the loop to the reactor environment will always be required. To minimize the extrapolation that such engineering procedures usually entail, a good understanding of the parameters which may affect the structural integrity of materials in sodium is necessary. A parameter which is presently recognized as potentially capable of affecting material properties is sodium mass transfer.

The purpose of the present report is to

- 1) Present a theoretical formulation of the sodium mass transfer problem
- 2) Review briefly the recent literature on sodium mass transfer
- 3) Present preliminary data on the corrosion behavior of stainless steel in sodium under high-heat-flux conditions.

II. THEORETICAL BACKGROUND

A. BASIC EQUATIONS

From the point of view of mass transport phenomena, the environment of a system, such as the liquid-metal-cooled fast reactor, may be characterized by the existence of chemical potential gradients and a moving liquid metal. The mass flux⁽¹⁾ in such a system may be written as:

$$\bar{J}_i = C_i \bar{V} - \alpha \text{grad } \mu_i (T, p, C_i) \quad \dots (1)$$

Here $\bar{J}_i$ is the flux of species i , C_i is the concentration, $\bar{V}$ is the velocity of the fluid, α is a proportionality constant, and μ_i is the chemical potential, which in general is a function of temperature, pressure, and concentration. The first term on the right of Equation 1 represents the transport of solute due to convection, while the second term accounts for transport by diffusion.

Since Equation 1 describes the transport of matter in the liquid phase only, an additional equation which describes the dissolution of solid into the liquid phase is required. Such dissolution of solid surfaces takes place when there exists a chemical potential difference between the solid and the liquid phase. The rate of metal dissolution is obtained from the chemical equation which describes the metal dissolution process. For example, a metal M can enter directly into the solution:

or it may react at the surface first with some species X in the solution and then dissolve as a compound.

By performing a mass balance on species i in an elemental volume of the liquid and using Equation 1, the basic convective diffusion equation is obtained:

$$\frac{\partial C_i}{\partial t} + \bar{V} \text{grad } C_i = D_i \nabla^2 C_i + S_i, \quad \dots (5)$$

where D_i is the diffusion coefficient defined as

$$D_i = \alpha \left(\frac{\partial \mu}{\partial C} \right)_{T,p}$$

and S_i is the rate of production or disappearance of the species per unit volume. The appearance of the velocity in the above equation requires an additional equation which describes the velocity field; this is provided by the Navier-Stokes equations. The task of calculating mass transfer rates is to solve the diffusion equation together with the Navier-Stokes equations and the proper boundary conditions.

Unfortunately there are very few geometries where this can be achieved and only when the flow is laminar. In most practical engineering applications, however, the flow is turbulent throughout most of the region of interest and one must resort to empirical correlation.

The convective diffusion equation, the Navier-Stokes equations, and the heat transfer equation are all of the same form and therefore there exists an analogy between momentum, heat, and mass transfer. As it is the practice in heat transfer problems to introduce a film heat transfer coefficient, it is convenient to introduce analogously a mass transfer coefficient h_m in mass transfer calculations.

$$J_i = h_{m,i} (C_{s,i} - C_{b,i}) \quad \dots (6)$$

According to this equation the transport of species from the surface of a dissolving solid to the bulk liquid stream is governed by the respective concentration difference and the mass transfer coefficient $h_{m,i}$.

Since the surface concentration C_s is not in general a measurable quantity it will be useful to eliminate C_s from Equation 6. From the overall reaction, Equation 2, the net metal dissolution rate is

$$\frac{d(M)_{\text{solid}}}{dt} = K_1 - K_2 C_s \quad \dots (7)$$

at equilibrium; however,

$$\frac{d(M)_{\text{solid}}}{dt} = 0$$

and C_s^0 equals the concentration of the saturated solution. Thus

$$K_1 = K_2 C_s^0 \quad \dots (8)$$

and

$$\frac{d(M)_{\text{solid}}}{dt} = K_2 (C_s^0 - C_s). \quad \dots (9)$$

Under steady-state conditions, the rate of metal dissolution equals the rate of the diffusion of metal away from the surface. Consequently, C_s can be eliminated from Equation 6 and 7 and the equation for the mass flux becomes:

$$J = \frac{C_s^0 - C_b}{\frac{1}{h_m} + \frac{1}{K_2}} \quad \dots (10)$$

To account for the experimental observation, that mass transfer in liquid metal is oxygen concentration dependent, Epstein⁽²⁾ suggested expressing K_2 as follows:

$$K_2 = K_o + KC_{O_2} , \quad \dots (11)$$

where K_o accounts for any metal dissolution at zero oxygen content, combining Equations 10 and 11:

$$J = \frac{C_s^o - C_b}{\frac{1}{h_m} + \frac{1}{K_o + KC_{O_2}} + \frac{1}{h_r}} \equiv h'(C_s^o - C_b) . \quad \dots (12)$$

The term $1/h_r$ was inserted in Equation 12 to account for any possible barrier to mass transfer which may exist between the liquid and the solid metal surface.

B. REACTOR SYSTEM

In a complex geometry such as the reactor, the parameters h_m , $C_{s,i}$, and $C_{b,i}$ vary throughout the system. To illustrate the source of these variations, it is convenient to represent the reactor system by two heat exchangers and the piping connecting them (Figure 1).

In Figure 1, T_{Hb} and T_{Hc} designate the bulk temperature of the sodium and the cladding temperature, respectively, in the proper generated channels. T_{Cb} represents the bulk sodium temperature in the heat exchanger.

Typical sodium bulk and cladding temperatures in fast reactors are shown in Figure 2.⁽³⁾

1. Variations in h_m

a. Entrance Region

As the sodium enters the heated channel, it forms a boundary layer on the channel walls. Initially, the thickness of the boundary layer is small in comparison with the hydraulic diameter of the channel and the flow there is laminar.

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Figure 1. Schematic of a Reactor System Showing Driving Force for Mass Transfer

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Figure 2. Typical Sodium Bulk and Cladding Temperatures in Fast Reactors

Further downstream the flow in the boundary layer undergoes transition to turbulent flow. With further increase in distance from the entrance, the thickness of the boundary layer becomes comparable with the radius of the channel, and the entire channel is filled with turbulent flow which is subjected to viscosity effects. Because of the similarity between momentum, mass, and heat transfer, it can be expected that similar phenomena will take place in the case of diffusion or heat transfer from the surface towards the bulk sodium. The local thickness of these boundary layers, however, is not the same. The ratio of the thermal to hydrodynamic layer thickness depends on the Prandtl number, while the ratio of the diffusion to the hydrodynamic layer thickness depends upon the Schmidt number. Since the Schmidt number for iron diffusion in sodium is of the same order of magnitude as the Prandtl number for water, the entrance length can be estimated from heat transfer data.⁽⁴⁾ Depending on the Reynolds number, the entrance length ranges from $x/d = 20$ at a Reynolds number of 101,000 to $x/d = 70$ for a Reynolds number of 12,700. For typical velocities and channel dimensions it can be expected that the entrance length in a reactor channel will extend to about one-sixth of the fuel pin length.

In the entrance region the mass transfer is significantly larger than in the fully developed regime.

b. Fully Developed Regime

The distribution of the diffusing species in the fully developed flow regime can be characterized by four zones:

- 1) Main turbulent stream
- 2) Turbulent boundary layer
- 3) Viscous sublayer
- 4) Diffusion sublayer

In the main turbulent core the concentration is constant. In the turbulent boundary layer, mass transfer takes place by eddies; in Zones 3 and 4, mass transfer takes place predominantly by molecular diffusion.

Drawing an analogy between heat and mass transfer, the coefficient h_m in the fully established turbulent regime is given by

$$\frac{h_m d_h}{D} = 0.023 (Re_{d_h})^{0.8} (Sc)^{0.4}. \quad \dots (13)$$

2. Variations in C_s^O

The equilibrium solubility C_s^O of the dissolving species in the liquid next to the surface is a function of the local temperature.

By chemical analysis, it was found that C_s^O for iron in sodium, and the reaction rate constant K_2 over the range of 225 to 500°C in the presence of 100 to 300 ppm oxygen are:⁽²⁾

$$C_{s(ppm)}^O = -8.7 + 0.0277 T^O \quad \dots (13a)$$

$$\log_{10} K = -0.007 - 2,388/T^O. \quad \dots (13b)$$

Radiochemical measurement techniques yielded solubility values which are by more than two orders of magnitude lower than those given above.⁽⁵⁾

3. Variations in C_b

The variations in the bulk concentration of the dissolved products along a channel can be derived by performing a mass balance on a volume element of the sodium, dv :

$$h'(C_s^O - C_b)Pdx = QdC_b, \quad \dots (14)$$

where P is the wetted perimeter and Q is the volumetric flow rate of the sodium. Since h' and C_s^0 are functions of x alone, Equation 14 can be written as

$$\frac{dC_b}{dx} + f_1(x)C_b = f_2(x), \quad \dots (15)$$

where

$$f_1(x) = \frac{P}{Q} h'$$

$$f_2(x) = \frac{h' P C_s^0}{Q},$$

subject to boundary condition

$$x = 0 \quad C = C_{b,in}.$$

Since $C_{b,in}$ in general is not known, equations similar to Equation 15 can be written for the cold leg of the reactor and the piping which connect the cold and hot sections. These equations are sufficient to determine the local corrosion rate at any point in the system, provided, of course, that the rate constant h' and the equilibrium solubility as a function of temperature are known. Depending on system geometry and operating temperatures, special cases which may arise can be visualized:

a) $C_{b,in} = C_{b,in}^0$ (equilibrium solubility at entrance temperature)

b) $C_{b,in} \ll C_{b,in}^0$

c) $C_{b,in} = C_{b,in}^0 + \text{"floating corrosion products in the liquid stream"}$

If $h'(x)$ and $C_s^0(x)$ in Equation 15 are replaced by some constant average value, then it can be shown that the local bulk concentration at any given location x along the channel is

$$C_b = \bar{C}_s^0 - \left(\bar{C}_s^0 - C_{b,in} \right) \exp \left(- \frac{[(P)x]h'}{Q} \right) \quad \dots (16)$$

Equation 16 indicates that when

$$\frac{[(P)x]h'}{Q} \rightarrow \infty \quad C_b \rightarrow \bar{C}_s^0 \quad \dots (16a)$$

$$\frac{P(x)h'}{Q} \rightarrow 0 \quad C_b \rightarrow C_{b,in} \quad \dots (16b)$$

In the first case, the bulk concentration tends towards the maximum concentration, i.e., the equilibrium solubility at the surface; however, since the bulk temperature is lower than the surface temperature, C_b cannot attain the value of C_s^0 and will assume the equilibrium solubility value at the bulk temperature $C_b \rightarrow C_b^0$. In the second case, the bulk concentration will remain essentially constant throughout the channel, having the value of the bulk concentration at the inlet. If the inlet flow, however, is rich in suspended particulate corrosion products, this case will be reduced to Case 1 which was discussed above.

In summary, the maximum and minimum corrosion rates (R) in the channel will be attained when

$$R = h' \left(\bar{C}_s^0 - C_{b,in} \right), \quad C_{b,in} \ll \bar{C}_s^0 \quad \dots (17)$$

$$R = h' \left(\bar{C}_s^0 - \bar{C}_b^0 \right), \quad \text{either when } \frac{[(P)x]h'}{Q} \rightarrow \infty, \quad \dots (18)$$

or the inlet flow is rich with insoluble corrosion products.

III. REVIEW OF SODIUM MASS TRANSFER STUDIES

Table 1 summarizes the corrosion rates of stainless steels in sodium and potassium as observed in loop experiments during the past few years. As it can be seen from the table, the reported corrosion rates vary over a wide range. These variations in corrosion rates would tend to indicate that the rate of material loss in flowing sodium is very dependent on the experimental conditions. Unless this dependence is well understood, any information about corrosion which was obtained in loop studies will remain of limited use for reactor application.

A. MASS TRANSFER STUDIES AT GENERAL ELECTRIC

The most comprehensive study to date of corrosion of stainless steel in sodium has been conducted by General Electric. This work was begun in 1959 and is a part of the AEC Sodium Components Program.

Generally speaking, the observed corrosion behavior in stainless steel can be characterized by overall corrosion of the test specimen and by selective leaching.

1. Overall Corrosion

For the overall corrosion rate, GE⁽⁶⁾ obtained the following empirical equation for the hot leg region:

$$R = V^{0.884} C^{1.156} \exp \left(12.845 - \frac{23,827}{T + 460} - 0.00676 \frac{L}{D_i} + \frac{2.26}{t + 1} \right), \quad \dots (19)$$

where

R = metal loss rate (mg/dm²-Mo)

V = sodium velocity (ft/sec)

C = oxide impurity concentration (ppm)

T = local sodium temperature (°F)

TABLE 1
COMPARISON OF VARIOUS CORROSION TEST DATA

Convection Type Loop	Fluid	Maximum Temperature (°F)	ΔT (°F)	Velocity (ft/sec)	Oxygen Content (ppm)	Material	Corrosion Rate (mil/yr)	Reference
Forced	Sodium	1200	250-500	21	11	316 SS	0.3 - 3.0	General Electric ⁽⁸⁾
Forced	Sodium	1100-1330	-	10-27	5-25	316 SS	0.16 - 4.0	British ⁽⁹⁾
Forced	Sodium	1600	550	25	-	16 SS	30 - 140	Pratt & Whitney ⁽¹⁰⁾
Natural	Sodium	1575	475	0.167	-	316 SS	0.36	United Nuclear ⁽¹¹⁾
Natural	Potassium	1575	475	0.34	-	316 SS	0.12	United Nuclear ⁽¹¹⁾
Forced	NaK	1300	200	2.1	30	347 SS	0.21	Oak Ridge ⁽¹²⁾
Forced	Potassium	1500	300	5	32-38	316 SS	0.20	Allison ⁽¹³⁾
Forced	Sodium	1472	-	10	300-500	20% Cr 15% Ni Steel	12	Russian ⁽¹⁴⁾
Forced	Sodium	1472	-	10	30-5	20% Cr 15% Ni Steel	2.58	Russian ⁽¹⁴⁾

L/D_i = number of diameters downstream in an isothermal region (where
L = length and D_i = hydraulic diameter*)

t = time (months)

A similar expression was derived for the cold leg region.

In order to account for the observed variation of mass transfer on velocity, oxygen concentration, temperature, and system geometry, Mottley⁽⁷⁾ proposed the following reaction mechanism:

where the subscript, s, refers to conditions at the metal surface.

In order to test the validity of this model the following assumptions were required:

- 1) The rate limiting step is the diffusion of FeO away from the surface.
- 2) The concentration of FeO in solution next to the surface is the equilibrium FeO concentration and is dependent on the sodium oxide concentration.
- 3) The isothermal region which comprises the piping, sample holder, and the sample proper can be broken into two regions: (a) the noncorrosion region, which includes the piping and the sample holder, and (b) the corrosion sample.
- 4) Sodium velocity was assumed to be independent of system coordinates.
- 5) The equilibrium solubility of iron in sodium is that which corresponds to data obtained from activity rather than from chemical analysis.

Comparison of the theoretical predictions with the experimental data give an order of magnitude agreement.

*Not consistent with nomenclature used elsewhere in this document.

One of the major objections to Mottley's model is that it employs the solubility values from activity measurements. Recent measurements^(15,16,17) of equilibrium solubility of iron in sodium are in general agreement with the higher value of the solubility (in the ppm range). Also, the presence of oxide in the system⁽¹⁵⁾ does not affect the solubility of iron such as to lower it by two orders of magnitude.

The dependence of the corrosion rate on the linear velocity would seem to indicate that the corrosion rate is not a strictly solution-rate-controlled process. If mass transfer through the liquid layer next to the surface plays a role in the corrosion, then the so-called downstream effect as predicted by Equation 1 could be attributed both to the entrance effect and the downstream enrichment in corrosion products. Work is now in progress at General Electric⁽¹⁸⁾ to provide quantitative information about each of these effects.

2. Selective Leaching

With Type 316 stainless steel, it has been observed that initially chromium, nickel, and manganese dissolve in sodium at a faster rate than their stoichiometric ratio. As a result, within a thin layer of approximately 0.4 mil, the alloy is rich in iron and depleted of chromium, nickel, and manganese.

Brush⁽¹⁹⁾ postulated the leaching process to be composed of selective and non-selective corrosion rate components. The total rate of dissolution, J_i , of a particular specimen is the sum of these components:

$$J_i = J_{i(\text{selective})} + J_{i(\text{non-selective})} \quad \dots (22)$$

The non-selective component of an active species is simply proportional to the overall alloy corrosion rate and its concentration in the metal at the surface:

$$J_{i(\text{non-selective})} = C_{i0} R. \quad \dots (23)$$

To obtain the selective component, $J_{i(\text{selective})}$, the transport of a particular species from the bulk metal to the surface was described by the diffusion equation:

$$\frac{\partial C_i}{\partial t} = \frac{\partial}{\partial x} \left(D_i \frac{\partial C_i}{\partial x} \right) + \frac{R}{\rho_a} \frac{\partial C_i}{\partial x}, \quad \dots (24)$$

where D_i is the diffusion coefficient of species i in the alloy, x is a distance coordinate with the origin at the liquid-metal interface, and ρ_a is the density of the alloy. In the first term on the right-hand side of Equation 24, the diffusion coefficient was inserted inside the parenthesis to account for its variation when the layer is depleted to such extent that transition from γ -iron to α -iron takes place.

Assuming a constant alloy corrosion rate and constant flux $J_{i(\text{selective})}$ a solution for the concentration profile in the depleted zone for each species in terms of the diffusivity coefficients, the flux $J_{i(\text{selective})}$ and the overall corrosion rate R was obtained.

From the results of this model, it was concluded that the depleted zone will attain its maximum thickness when

$$\frac{R \sqrt{t}}{2\rho_a \sqrt{D_i}}$$

and then each element will dissolve in accordance with its stoichiometric proportions. Experimental observations were in agreement with the theoretical predictions.

B. MASS TRANSFER STUDIES IN GREAT BRITAIN

The effect of temperature (450 to 725°C), velocity (0.5 to 40 ft/sec), and oxygen content (5 to 25 ppm) on the corrosion behavior of austenitic steels and nickel alloys has been studied by Thorley and Tyzack in England.⁽¹⁵⁾ Ferritic layer formation of a thickness of about 0.5 mil and corrosion rates up to 2.1 mil/yr have been observed by these investigators.

At relatively low velocity, the corrosion rate was found to increase approximately linearly with the velocity; as the velocity was further increased the corrosion rate reached a plateau and became independent of velocity. The height of this plateau was directly proportional to the oxygen concentration.

Examination of Equation 12 would indicate that such a behavior is possible where there is a transition from mass transfer in the liquid phase to a surface-reaction-controlled process. Thorley and Tyzack attributed the observed deviation of velocity dependent corrosion rate to slow deposition of iron in the cooler parts of the circuit and/or some chemical process involving iron removal from the sodium via a double oxide.

C. RUSSIAN WORK

Agapora et al,⁽¹⁴⁾ studied the behavior of iron, nickel, chromium, and austenitic steels in flowing sodium at temperatures up to 880°C, in the velocity range of 2 to 29 ft/sec, and with an oxygen content of 30 to 500 ppm. It was observed that at 30 to 50 ppm oxygen concentration and at a velocity of 2 ft/sec the corrosion rates of nickel and chromium increase with the temperature while iron gains weight with increasing temperature. Keeping the temperature constant at 880°C and increasing the flow rate and oxygen content to 500 ppm show an increase in the corrosion rate of iron with the increase in velocity and a weight gain by the nickel specimen. The period of testing at which these results were obtained was 500 hr. At a velocity of 17 ft/sec the corrosion rate of iron was 4.2 mil/yr.

Depending on the oxygen content, an increase in the nickel content in iron may increase or decrease the rate of corrosion of the alloy. In austenitic steel with 15% Ni and an oxygen level of 300 to 500 ppm, an increase in chromium content increases the rate of corrosion. At a chromium content of 20%, a velocity of 10 ft/sec, and a temperature of 800°C, the corrosion rate was found to be 12 mil/yr. At an oxygen level of 30 to 50 ppm, this value dropped to 2.5 mil/yr. Selective leaching with α -iron lattice in austenitic stainless steel was also observed. These changes indicated an abrupt reduction in the tensile strength of the steel.

D. MASS TRANSFER STUDIES WITH ROTATING DEVICES

Kassner⁽²⁰⁾ studied the dissolution of Type 304 stainless steel in Bi⁴² wt % Sn eutectic alloy by means of a rotating disc. The temperature and velocity dependence of the corrosion rate indicated that at relatively low velocity corrosion was controlled by a liquid-diffusion process. At higher velocities, however, a transition to a surface-reaction-controlled process was observed.

Johnson⁽¹⁶⁾ studied the corrosion behavior of Type 304 stainless steel by means of a rotating cylinder which was immersed in sodium at 1200°F containing oxygen less than 20 ppm. The solubility of iron in sodium was found to be in the range of 17 to 26 ppm. The mass transfer coefficient h' was found to be between 6.5×10^{-6} to 7.5×10^{-6} cm/sec. The rotational speed and oxygen content were not varied in these experiments. Because of the low value of the mass transfer coefficient h' obtained in these experiments, it was concluded that corrosion, at least in these experiments, was surface-reaction-controlled.

E. ANALYSIS

It would be desirable to apply Equations 12, 13, 13a, 13b, and 15 to a typical fast reactor and calculate the corrosion rate of fuel cladding in such a system. The difficulty in pursuing such a procedure is that neither the mass transfer coefficient h' or the equilibrium solubility C_s^0 can be calculated to any degree of certainty. Therefore, before performing the rigorous calculations for a reactor system it would be necessary to test the validity of various mass transfer models against observed mass transfer rates in various loops. However, since loop data introduces additional unknowns (namely, the bulk concentration), such a test cannot be expected to be conclusive.

Employing the two values of solubility as reported in the literature, and assuming either the liquid-controlled or surface-reaction-controlled corrosion process, theoretical calculations were made for predicting the corrosion rate in two loops.^(16, 17) In each case, the bulk concentration was assumed to be constant through the test section and equal to the equilibrium solubility

TABLE 2
COMPARISON BETWEEN CALCULATED AND EXPERIMENTALLY
DETERMINED CORROSION RATES

Source	Material Type	Solution	Maximum Temp (°F)	ΔT (°F)	Reynolds Number	O ₂ Content (ppm)	Corrosion Rate (mil/yr)		
							Obs.	Calc. (2)	Calc. (3)
Ref. 10 ⁽¹⁾	316 SS	Na	1600	550	9.4×10^4	Not reported	7-35	12	24 ⁽⁴⁾
Ref. 12	347 SS	NaK	1300	200	4×10^4	30	0.21	0.67	1.9

- (1) Loop Test SSSA-2C2. Two to three times higher corrosion rates were obtained for Loop Test SSSA-3C3.
- (2) Liquid-mass-transfer-controlled process, using lower values of solubility data.
- (3) Surface reaction controlled process, using higher values of solubility data.
- (4) This value corresponds to an oxygen level of 100 to 300 ppm.

corresponding to the temperature at the entrance to the heater. Two cases were considered:

1) The corrosion process is controlled by diffusion of FeO away from the surface and the equilibrium solubilities are predicted by the equation

$$S = 2.49 \times 10^{-8} - 1.55 \times 10^{-10}T + 2.95 \times 10^{-13}T^2 \quad \dots (25)$$

This case corresponds to using Mottley's model.⁽⁷⁾

2) The corrosion process is controlled by surface reaction rates. The rate constant and the solubilities are given by Equations 13a and 13b.

The results of the calculations are compared with the experimental data in Table 2. From the comparison, and because of the many approximations which are introduced, neither method of calculation can be ruled out. It should be noted, however, that using a liquid-controlled mass-transfer model together with the higher solubility data would predict corrosion rates which are larger by two orders of magnitude than those observed. Similarly, using the surface-reaction-controlled model together with the lower solubility data will yield results which are smaller by two orders of magnitude than those observed. The first case, however, cannot be ruled out because the method used for calculating the surface-to-bulk concentration gradient could have overestimated this value; on the other hand, using the lower solubility value could underestimate the absolute magnitude of the concentration. The combination of these two possible errors could have resulted in the fair agreement of the calculations with the experimental observations. However, when a liquid mass-transfer model is applied to systems where the minimum concentration gradient is presumably known (the SRE or the EBR, where the minimum concentration gradient corresponds to the surface-to-bulk temperature difference) and the higher values of the solubilities are employed, the corrosion rates thus calculated are considerably larger than those observed in these reactors. Experimental evidence, which indicates a velocity and entrance effect on the corrosion rate, favors a liquid mass-transfer-controlled process. The use of this model together with the solubilities in the 10^{-2} ppm range, although showing a fair

agreement with corrosion data, is not consistent with the many independent measurements, indicating that the solubility is in the ppm range.

Other mechanisms of mass transfer could be explored. For example, the transfer of corroding species, instead of being controlled in a liquid boundary layer or at the surface, could feasibly be controlled in a thin, partially compact, oxide layer which adheres to the surface. The effect of flow velocity on mass transfer could then be attributed to the "blowing off" of the film as a result of the shear stress at the oxide liquid interface. Exploration of the above model would require a study into the differences in properties of iron oxide and nickel oxide films, since it was observed that the corrosion behavior of these two metals differs considerably.⁽¹⁴⁾ Since the solubility of oxygen in sodium is in the 10^{+2} ppm range at temperatures of interest, a model which assumes the diffusion of oxygen to be the rate limiting process would predict corrosion rates which are much higher than those observed.

In summary, available loop data on corrosion of steels in sodium do not permit conclusive determination of the mass transfer coefficient h' which is needed for predicting corrosion rates in a reactor system. The main reason that prevents the determination of h' is that the bulk concentration is not known in loops where there is simultaneous dissolution and deposition of material in the various parts of the loop. What is needed, therefore, is a system where the number of variables is kept at a minimum. A system where the surface and bulk concentration and the liquid-mass-transfer coefficient can be determined within a large degree of certainty will fulfill these requirements. A system which approaches these requirements is shown in Figure 3. It consists of an annulus through which liquid sodium is flowing with known velocity V and uniform temperature T_L . At point A there is a step change in surface temperature due to heat generation in Section A-C. The heat flux from the surface to the bulk sodium is sufficiently large to establish about a 50°F gradient between the surface and the bulk stream. At point B the diffusional, thermal, and hydrodynamic flows are fully established. The surface-area-to-volume ratio upstream of the test section is sufficiently large so that the bulk concentration at point B corresponds to the bulk liquid temperature. Thus, knowing the bulk and surface temperatures fixes the bulk-to-surface concentration gradient. In such a system, varying the velocity while keeping the surface and bulk temperature constant should not have an effect on

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Figure 3. Schematic of a System with Surface-to-Bulk Temperature Gradient for Mass Transfer Studies

the corrosion rate if the process is surface-reaction controlled. Conversely, if the process is liquid-phase mass-transfer controlled, from the measured corrosion rate and the calculated liquid-mass-transfer coefficient h_m , the concentrations at the surface and in the bulk can be calculated and compared with both the high and low values of solubilities published in the literature.

In order to estimate the magnitude of the corrosion rate of such a system, Mottley's model was applied to calculate the material loss from an inner tube of 0.2 in. in diameter. The average surface temperature of the tube was assumed to be 1250°F and that of the bulk sodium, 1200°F. For a hydraulic diameter of 0.2 in. and flow rate of 6 gpm, a corrosion rate of 2.7 mil/yr was calculated for this system (Appendix II).

IV. EXPERIMENTAL STUDIES AT HIGH-HEAT-FLUX

In connection with the cladding test program at AI, a loop similar to the system described above was designed and built, thus also permitting the study of corrosion in sodium.

A. DESCRIPTION OF TEST LOOP

Figures 4 and 5 show the general layout of the basic test installation. The loop is constructed of Type 304 stainless steel, Schedule-40 pipe with valves of Type 316 stainless steel. The loop is designed for 1400°F operation at a maximum operating pressure of 50 psig. Normal operation is between 1200 and 1300°F with a helium cover gas pressure of 1 to 2 psig. The design sodium flow rate is 40 gpm maximum with 15 psig ΔP across the EM pump. Sodium purity is maintained at less than 10 ppm dissolved oxygen using a forced circulation cold trap. The oxygen concentration is monitored with a plugging meter. Sodium samples are taken weekly during operation to check on the impurity levels in the sodium system. The loop contains 51 gal of sodium when filled to normal operating level. The surface area exposed to 1200°F sodium is 105 ft².

The test section and high-heat-flux heater are shown in Figure 6. There are five parallel sets of two test sections in series. The test section is designed to achieve a maximum velocity of 20 ft/sec recorded at a nominal flow rate of 4 to 6 gpm. The test section is 8 in. long with an ID of 0.461 in. There is a helical wire wrap in the test section which centers the 0.295-in.-OD (including cladding) high flux heater to assure uniform velocity around the heater.

B. LOOP OPERATION

The high-heat-flux heaters to be tested are installed in the test sections. The loop is then preheated and filled with sodium. After sodium circulation is established, the immersion heater is used to bring the loop up to an operating temperature of 500°F. The loop plugging temperature is measured with the plugging meter and the cold trap is operated as required to lower the oxygen concentration in the loop to less than 10 ppm. The loop temperature is then increased to the required test section inlet temperature of 1200 to 1300 $\pm$ 3°F

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Figure 4. Cladding Test Loop

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Figure 5. General Layout of Test Loop

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Figure 6. High-Heat-Flux Heater and Test Section Schematic

Figure 7. Heater Sheath After Exposure to Flowing Sodium at Average Temperature of 1218°F for 464 hr (showing shiny surface appearance on the heated section and dull surface appearance on the unheated section)
(heat flux: 10^6 Btu/ft²-hr)

of set point. The high-heat-flux heaters are operated at heat fluxes up to 1.0×10^6 Btu/ft²-hr for periods up to 600 hr.

C. RESULTS

Only preliminary data have been obtained to date on the behavior of stainless steel in sodium under high-heat-flux conditions. The exposure time of the heater sheath in sodium was too short for weight change measurements. The results thus far obtained are of sufficient interest to warrant presentation.

Figure 7 shows the surface appearance of a heater exposed for 463 hr at a sodium temperature of 1218°F. The heated section had a shiny surface, while the surface of the unheated section was dull.

Figure 8 shows the surface of a heater sheath after exposure to 1200°F sodium for 433 hr. It can be seen that a reaction layer was formed on the heated section only.

Figure 9 shows a photomicrograph (500 to 1500 magnification) of the reaction layer which was estimated to be 0.22 mil thick. The preparation procedure for these photomicrographs was as follows:

- 1) The specimens were ground on 320 to 600 grid silicon carbide papers and then diamond polished by using a 15 μ , 6 μ , and 1 μ diamond on netcloth.
- 2) Final polishing was done with a slurry of 0.3 μ alpha alumina and then to a slurry of 0.05 μ gamma alumina.
- 3) The specimen was then lightly etched in a 10% chromic acid solution to delineate its entire structure.

The diffraction patterns obtained by x-ray fluorescence indicated that the layer was predominately that of α -iron with small amounts of γ -iron and a phase which could be either Cr_2C or Cr_2N . The depletion in Ni and Cr was estimated to be approximately 30%.

Figure 10 shows the distribution of Fe, Ni, and Cr in the reaction layer as obtained by a microprobe trace. (These data were not obtained for the same sample as shown in Figure 9.) The bulk concentration of Ni and Cr in the steel was obtained by chemical analysis and found to be 8.9 (wt %) Ni, 19 Cr, and 68.4 Fe.

b. HEAT FLUX ZONE
FERRITE LAYER
(AREA 1)

500X

c. NOSE END—
NO LAYER
(AREA 2)

500X

Figure 8. Ferrite Layer Location

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a. MAGNIFIED 500X

b. MAGNIFIED 1500X

FERRITE VERIFIED BY: X-RAY DIFFRACTION
 X-RAY FLUORESCENCE
 PREFERENTIAL ETCHING
 MAGNET (TENTATIVE)
 ELECTRON PROBE
 LOCATION: UNFAILED HEATER HIGH HEAT FLUX ZONE.

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Figure 9. Heater Sheath Ferrite Layer

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Figure 10. Ferrite Layer Cr-Ni Depletion
(Electron-Microprobe)

A 0.43 mil surface layer which is enriched in Fe and depleted of Ni and Cr is distinguished on the trace. The discontinuity in the concentration gradients of Cr and Ni is attributed to the transition from γ -iron to α -iron.

D. DISCUSSION

As already mentioned above, the exposure time of the heater sheath was too short to obtain measurable corrosion rates. However, indirect estimation of the material dissolved can be made from the observed thickness of the ferritic layer. Before such an estimate is made it would be of interest to determine whether the thickness of the ferritic layer is consistent with simplified calculations of the rate at which a given species (nickel, for example) can be depleted from the bulk metal. For these calculations the motion of the interface due to the overall corrosion rate will be neglected, and it will be assumed that the concentration of nickel is lowered to zero immediately after exposing the heater sheath surface to the flowing sodium and is held at this value during the testing lifetime of the heater.

The appropriate differential equation and the boundary conditions which describe the transport of nickel in the alloy can be written as

$$\frac{\partial C}{\partial t} = \frac{\partial}{\partial x} \left(D_{Ni}^{alloy} \frac{\partial C}{\partial x} \right) \quad \dots (26)$$

$$\begin{array}{lll} \text{At } t = 0 & C = C_{\infty} & \text{for } x > 0 \\ \text{At } x = 0 & C = 0 & \text{for } t > 0 \\ \text{At } x = \infty & C = C_{\infty} & \text{for } t > 0 \end{array}$$

Here C denotes the concentration of nickel, t is the time, and x is the distance from the metal-sodium interface into the material. Thermal diffusion and effects of curvature and axial diffusion were neglected in Equation 1.

Because of the phase transformation which takes place in the depleted layer, it is necessary to divide the layer into two regions: one which consists of γ -iron and the other consisting of α -iron. The diffusion coefficients of nickel vary considerably in the respective regions. Thus, in order to obtain a solution for the present problem, a phase diagram for the alloy and data about the diffusion coefficients in the two phases are required.

Information about diffusion coefficients of nickel in stainless steel is not readily available. A first approximation would be to consider the system as consisting of nickel and iron alone. For such a system, the diffusion coefficient of nickel in γ -iron is given by: ⁽²¹⁾

$$D_{Ni}^{\gamma-Fe} = (0.344 + 0.012 \times \text{wt.}\% \text{ Ni}) \exp \left(- \frac{67,500}{RT} \right) \quad \dots (27)$$

in the temperature range from 1050 to 1450°C. The diffusion coefficient of nickel in α -iron can be assumed to be approximately equal to the self-diffusion coefficient of iron in α -iron. ⁽¹²⁾

Experimental values of diffusion coefficients of iron in α -iron at 1250° F are:^(22,23)

$$D_{\text{Fe}}^{\alpha\text{-Fe}} = 4.6 \times 10^{-13} \text{ cm}^2/\text{sec}$$

$$D_{\text{Fe}}^{\alpha\text{-Fe}} = 3.16 \times 10^{-14} \text{ cm}^2/\text{sec}$$

The latter value of the diffusion coefficient was obtained from more recent experiments and is also believed to be more reliable.⁽²⁴⁾

For the case when the diffusion coefficient is assumed to be constant, the following simple solution is obtained from Equation 26:

$$\frac{C(x, t)}{C_{\infty}} = \text{erf}\left(\frac{x}{2\sqrt{Dt}}\right)$$

$$\frac{C(x, t)}{C_{\infty}} \rightarrow 1, \quad \dots(28)$$

particularly when

$x \rightarrow \delta$ = thickness of depleted layer.

Using the diffusion coefficients of nickel in γ -iron and α -iron, the corresponding values of δ are as shown in Table 3.

TABLE 3
COMPARISON OF MAXIMUM OBSERVED WITH THEORETICAL
PREDICTION LAYER

Calculated (in. $\times 10^4$)			Observed (in. $\times 10^4$)	
$D_{\text{Ni}}^{\gamma\text{-Fe}}$ (Ref. 21)	$D_{\text{Fe}}^{\alpha\text{-Fe}}$ (Ref. 23)	$D_{\text{Fe}}^{\alpha\text{-Fe}}$ (Ref. 22)	Electron Microprobe (Fig. 9)	X-ray Diff. (Fig. 8)
0.22	0.875	3.34	4.3	2.2

Using Equation 28, the effective diffusion coefficient required to yield the observed thickness of 0.22 mils would be $1.25 \times 10^{-14} \text{ cm}^2/\text{sec}$.

Table 3 shows that the observed layer thickness is consistent with theoretical calculations. Because of the simplified approach employed in these calculations, an order of magnitude agreement is considered satisfactory.

In Figure 11 the ferritic layer observed in the present experiment is compared with the ferritic layers which were observed by the British and at General Electric. The figure shows a qualitative agreement between the concentration profiles in the layers which were formed under no heat flux conditions and the layer formed due to the heat flux at the surface of the specimen.

In order to estimate the overall corrosion rate from the ferritic layer thickness, the results of Brush's model will be employed in the present calculations.

For a constant corrosion rate and neglecting diffusivity changes through the layer, the rate of nickel depletion is given by⁽¹⁹⁾

$$J_{\text{Ni}} = R(C_{\text{io}} - C_{\text{i}\infty}), \quad \dots(29)$$

where

$C_{\text{io}}, C_{\text{i}\infty}$ = concentration of the element at the sodium-alloy interface and at the bulk of the alloy, respectively

R = overall alloy corrosion rate.

Using the following numerical values

$$t = 430 \text{ hr}$$

$$\rho_a = 8 \text{ gm/cm}^3$$

$$C_{\text{io}} = 3\% \text{ by weight}$$

$$C_{\text{i}\infty} = 9\% \text{ by weight}$$

$$\delta = 4.3 \times 10^{-4} \text{ in.},$$

$$J_{\text{Ni}} = (C_{\text{i}\infty} - C_{\text{io}}) \frac{\delta}{2} \frac{\rho_a}{t} = 1.7 \times 10^{-10} \frac{\text{gm}}{\text{cm}^2 \cdot \text{sec}}$$

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Figure 11. Comparison of the Formation of Ferritic Layers in Various Loops

in Equation 29 a corrosion rate of 4.3 mil/yr is obtained. Using Mottley's model for predicting corrosion rate and using Equation 29 without correcting for the fact that the flow was not fully developed, a corrosion rate of approximately 1.67 mil/yr was calculated.

Since data at only one velocity and one temperature have been obtained thus far, no conclusions can be drawn about mass transfer mechanisms. Preliminary observations, however, indicate that surface changes and corrosion rates under high heat flux conditions appear to conform to the data obtained in loops where the specimens were not subjected to a heat flux. The experimental method presented here should prove useful in further investigation of mass transfer in sodium, since it eliminates the dependence of the bulk liquid concentration of corrosion products on loop geometry and operation.

V. SUMMARY AND CONCLUSIONS

At present the mechanism for mass transfer of metals in flowing sodium is poorly understood. This current lack of understanding is due to the paucity of reliable solubility data of iron in sodium and the difficulty of obtaining the required data for theoretical computations from loop experiments. In a reactor system, there is a permanent driving force for corrosion due to the difference between cladding surface and bulk sodium temperatures. In current studies of mass transfer in loops, the driving force for corrosion is established by the deposition of metal in the cold leg of the loop. This practice makes it difficult to determine the bulk liquid concentration in the hot zone which is, in turn, needed for calculating the corrosion rate coefficient. Reliable predictions of mass transfer in a reactor system cannot be made until this coefficient is well known.

It was concluded that a loop test facility in which the driving force for corrosion, similar to that in a reactor, is set up by means of high heat flux, and where the geometry of the test section is amenable to analytical treatment, offers the potential for providing information about mass transfer mechanisms.

A loop of this type was designed and constructed. Preliminary results from initial tests show the formation of a ferritic layer, approximately 0.4 mils thick, on the heated section only, and indicate that the bulk liquid was saturated with respect to the corrosion products throughout the loop. Compared with the practice in conventional loops, the difficulty of determining the bulk concentration is thus eliminated.

APPENDIX I

COMPARISON OF CALCULATED CORROSION RATES WITH THOSE OBSERVED EXPERIMENTALLY

A. PRATT & WHITNEY DATA

The loop used by Pratt and Whitney for corrosion rate calculations is shown in Figure 12.

1. Dimensions of Loop and Test Conditions

Heater leg ID - 0.126 in.

Effective heater length - 31.6 in.

Maximum sodium temperature - 1600°F

Temperature gradient through heater - 550°F

Heater fluid velocity - 25 fps

Test time - 250 hr

Corrosion Rate:	Wall Thickness (mil)	
Heater Leg	<u>Before</u>	<u>After</u>
Entrance	61.0-62.0	62.0
Middle	61.0-62.0	61.6-61.8
Exit	35.5-36.0	34.5-35.

2. Calculations

a. Liquid Mass Transfer Controlled Process

The mass transfer coefficient is expressed as follows:

$$\frac{h_m}{D_{\text{FeO}}} = 0.023 (\text{Re}_d)^{0.8} (\text{Sc})^{0.4}$$

$$\text{Re}_d = \frac{ud}{\nu} = \frac{25 \times 0.126}{12 (2.8 \times 10^{-6})} = 9.4 \times 10^4$$

$$\text{Sc} = 15.3$$

$$(\text{Re}_d)^{0.8} = (9.4 \times 10^4)^{0.8} = 9.45 \times 10^3$$

a. SODIUM MASS TRANSFER LOOP

$$R = h_m (\bar{C}_w - C_\infty)$$

$\bar{C}_w$ = SOLUBILITY OF FeO IN Na AT AVERAGE HEATER TEMPERATURE (gm/cm³)

C_∞ = SOLUBILITY OF FeO IN Na AT ENTRANCE TEMPERATURE (gm/cm³)

$\bar{h}_m$ = MASS TRANSFER COEFFICIENT (cm/sec)

R = CORROSION RATE (gm/cm²-sec)

b. SIMPLIFIED CORROSION MODEL

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Figure 12. Pratt and Whitney Loop and Simplified Model for Corrosion Rate Calculations

$$(Sc)^{0.4} = (15.3)^{0.4} = 3$$

$$h_m = \frac{2.3 \times 3 \times 9.45 \times 10 \times 1.65 \times 10^{-4}}{0.126 \times 2.54} = 0.325 \text{ cm/sec}$$

The corrosion rate is given by

$$R = h_m (\bar{C}_w - C_b)$$

Here it is assumed that the sodium is saturated with corrosion products at the entrance to the heater (Figure 12). The concentration at the surface is taken at the solubility corresponding to the average temperature, 1325°F.

$$\bar{C}_w - C_b = 2.4 \times 10^{-8} \text{ gm Fe as FeO/cm}^3$$

$$R = (0.325)(2.4) \times 10^{-8} = 7.8 \times 10^{-9} \text{ gm/cm}^2\text{-sec} = 12 \text{ mil/yr}$$

b. Surface Reaction Controlled Process

$$\log K = -0.007 - 2,388/T$$

$$T = 1325^\circ\text{F} = 990^\circ\text{K}$$

$$\log K = -0.007 - \frac{2,388}{990} = -2.42$$

$$K = 0.384 \times 10^{-2} \text{ cm/sec}$$

At

$$T = 990^\circ\text{K}$$

$$C_s^O = -8.7 + (0.0277) 990 = 18.7 \text{ ppm .}$$

a. NaK MASS TRANSFER LOOP

b. SIMPLIFIED CORROSION RATE MODEL

$$R = h_m(\bar{C}_w - C_\infty) (K)$$

$\bar{C}_w$ = SOLUBILITY OF FeO IN Na AT 1300°F

C_∞ = SOLUBILITY OF FeO IN Na AT 1100°F

K = FACTOR CONVERTING CORROSION OF STAINLESS STEEL
IN NaK TO CORROSION IN SODIUM

h_m = MASS TRANSFER COEFFICIENT (cm/sec)

R = CORROSION RATE (gm/cm²-sec)

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Figure 13. ORNL Loop and Simplified Model
for Corrosion Rate Calculations

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At

$$T = 1050^{\circ}\text{F} = 840^{\circ}\text{K}$$

$$C_s^0 = -8.7 + (0.0277) 840 = 14.6 \text{ ppm}$$

$$\begin{aligned} R &\approx (3.84) \times 10^{-3} (18.7 - 14.6) \times 10^{-6} = (3.84) (4.1) \times 10^{-9} \\ &= 15.8 \times 10^{-9} \text{ gm/cm}^2\text{-sec} = 24 \text{ mil/yr} \end{aligned}$$

B. ORNL DATA

The loop used by ORNL for corrosion rate calculations is shown in Figure 13.

1. Test Conditions

Material - Type 347 SS (0.622 ID)

NaK flow velocity - 2.1 fps

Maximum temperature - 1300°F

Minimum temperature - 1100°F

Test time - 2659 hr

Corrosion rate - 1.46 mg/cm²

2. Calculations

a. Liquid Mass Transfer Controlling Process

$$\text{Re}_d = \frac{2.1 \times 0.622}{12 (2.8 \times 10^{-6})} = 4 \times 10^4$$

$$(\text{Re}_d)^{0.8} = 4.8 \times 10^3$$

$$(\text{Sc})^{0.4} = 3$$

$$\frac{h_m d}{D_{\text{FeO}}} = 0.023 (4.8 \times 10^3) \times (3) = 328$$

$$h_m = \frac{328 \times 1.65 \times 10^{-4}}{0.622 \times 2.54} = 3.43 \times 10^{-2} \text{ cm/sec}$$

$$R = h_m (\bar{C}_w - C_\infty)$$

C_∞ = solubility at the lowest loop temperature

$$3.4 \times 10^{-8} \text{ gm FeO/cm}^3 \text{ Na}$$

$\bar{C}_w$ = solubility at 1300°F ($5.25 \times 10^{-8} \text{ gm FeO/cm}^3 \text{ Na}$)
(no temperature drop through specimen)

$$R_{\text{NaK}} = (5.25 - 3.4)10^{-8} \times 3.43 \times 10^{-2} = 6.4 \times 10^{-10} \text{ gm/cm}^2\text{-sec}$$

$$= 166 \text{ mg/gm}^2\text{-Mo} = \approx 1 \text{ mil/yr}$$

$$R_{\text{Na}} \approx \frac{1}{1.5} = 1.67 \text{ mil/yr}$$

b. Surface Reaction Controlled Process

$$\log K = -0.007 - 2,388/T$$

$$T = 1300^\circ\text{F} = 980^\circ\text{K}$$

$$\log K = -0.007 - \frac{2,388}{980} = -2.44$$

$$K = (0.372) 10^{-2} \text{ cm/sec}$$

Since the oxygen concentration was only 30%, this value is corrected by dividing it by 10:

$$K = 3.72 \times 10^{-4} \text{ cm/sec} .$$

At

$$T = 1300^{\circ}\text{F} ,$$

$$C_s^o = -8.7 + (0.0277)(980) = 18.5 \text{ ppm} .$$

At

$$T = 1100^{\circ}\text{F} ,$$

$$C_s^o = -8.7 + (0.0277) 870 = 15.3 \text{ ppm} .$$

$$R \approx (3.72 \times 10^{-4}) \times (18.5 - 15.3) \times 10^{-6} = 3.72 \times 0.32 \times 10^{-9}$$

$$1.2 \times 10^{-9} \text{ gm/cm}^2\text{-sec} = 1.9 \text{ mil/yr}$$

To correct for the fact that NaK was used instead of sodium, this number is divided by a factor of 1.5⁽²¹⁾:

$$R \approx \frac{1.9}{1.5} = 1.27 \text{ mil/yr}$$

APPENDIX II

ESTIMATION OF CORROSION RATE UNDER HEAT FLUX CONDITIONS

The geometry of the system is illustrated in Figure 14, where the inner cylinder represents the specimen surface. The length to hydraulic diameter ratio exceeds 40 to ensure fully developed flow.

Figure 14. Geometry for Mass Transfer
Studies Under Heat Flux Conditions

Application of the mass-heat transfer analogy for a fully developed turbulent flow in an annuli gives the following expression for the mass transfer coefficient h_m :

$$\frac{h_m d_h}{D_{FeO}} = 0.023 \left(Re_{d_h} \right)^{0.8} (Sc)^{0.4} \left(\frac{d_2}{d_1} \right)^{0.45}, \quad \dots (2)$$

where

$$Sc = \text{Schmidt number} = \nu / D_{FeO} = 15.3$$

$$D_{FeO} = \text{diffusion coefficient of FeO in sodium at } 1200^\circ\text{F} \\ (\text{as estimated in Reference 7}) = 1.65 \times 10^{-4} \text{ cm}^2/\text{sec}$$

$$\nu = \text{kinematic viscosity at } 1200^\circ\text{F} = 2.52 \times 10^{-3} \text{ cm}^2/\text{sec}$$

$$Re_{d_h} = \text{Reynolds number} = 4Q / \nu \pi_a (d_2 + d_1) = 1.2 \times 10^5$$

$$Q = \text{flow rate} = 6 \text{ gpm}$$

$$d_1, d_2 = \text{specimen and outer tube diameters} = 0.2 \text{ and } 0.4 \text{ in, respectively}$$

$$d_h = \text{hydraulic diameter} = d_2 - d_1 = 0.2 \text{ in.}$$

Solving for h_m :

$$h_m = 0.023 \left(1.2 \times 10^5\right)^{0.8} (15.3)^{0.4} (2)^{0.45} \frac{1.65 \times 10^{-4}}{0.51} = 0.354 \frac{\text{cm}}{\text{sec}}$$

The corrosion rate R is given by

$$R = h_m (\bar{C}_w - C_\infty) \quad \dots (3)$$

where

$$\bar{C}_w^o = \text{solubility at } 1250^\circ\text{F} = 4.75 \times 10^{-8} \text{ gm Fe as FeO/cm}^3$$

$$C_\infty^o = \text{solubility at } 1200^\circ\text{F} = 4.25 \times 10^{-8} \text{ gm Fe as FeO/cm}^3$$

$$R = 0.354 (4.75 - 4.25) \times 10^{-8} = 1.77 \times 10^{-9} \text{ gm Fe/cm}^2\text{-sec}$$

$$= 1.77 \times 10^{-9} \times 2.6 \times 10^{11} = 460 \text{ mg/dm}^2\text{-Mo}$$

$$\approx 2.7 \text{ mil/yr}$$

NOMENCLATURE

α	= proportionality constant	
C	= concentration	(gm/cm ³)
d_h	= hydraulic diameter	(cm)
D_i	= diffusion coefficient	(cm ² /sec)
h	= mass transfer coefficient	(cm/sec)
K	= chemical rate constant	(cm/sec)
P	= perimeter	(cm)
p	= pressure	(gm/cm ²)
Q	= flow rate	(cm ³ /sec)
R	= corrosion rate	(gm/cm ² -sec)
S	= source term	(gm/cm ³ -sec)
T	= temperature	(°K)
t	= time	(sec)
V	= velocity	(cm/sec)
x	= distance coordinate	(cm)
μ	= chemical potential	(cal/mole)
ν	= kinematic viscosity	(cm ² /sec)
ρ_a	= density	(gm/cm ³)
Re	= Reynolds number	(gm/cm ² -sec)
Sc	= Schmidt number	(gm/cm ² -sec)

Subscripts

b	= bulk conditions
i	= species i
m	= transfer through liquid phase
o	= equilibrium conditions
r	= mass transfer coefficient (as defined in Equation 12)
s	= surface conditions
'	= overall mass transfer coefficient (as defined in Equation 12)

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