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for the HRT Chemical Processing Plant

TO: H. E. Goeller and W. E. Unger

FROM: W. L. Carter

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0.0 SUMMARY

Radiolytic decomposition of heavy water in the decay-storage vessels of the HRT chemical processing plant requires that a recombiner be installed for D_2 and O_2 recovery. Decomposition gases will be diluted with steam, passed through an entrainment separator, superheated and passed over a platinized alumina catalyst. D_2 and O_2 will be present in stoichiometric amounts at concentrations well into the non-explosive region. After recombination the D_2O vapor will be totally condensed and refluxed to the decay-storage tank.

The rate of water decomposition in the chemical plant is quite small when compared to the same reaction in the HRT core system. In fact sufficient steam will be produced by absorption of radiation in the fuel solution to dilute the combustibles below their explosive concentration. However, auxiliary heat has been provided for additional boil-up. To avoid poisoning the catalyst surface with condensed moisture, the recombiner is cradled in a mantle heated by 250 psig steam.

The following tabulation gives pertinent data on the recombiner design:

Catalyst	= platinized alumina containing 0.3% wt pure platinum
Catalyst size	= $1/8"$ x $1/8"$ right circular cylinders
Catalyst volume	= 682 ml
Recombiner size	= 3" pipe x 5" long
Normal operating gas rate (total)	= 10.3 lb/hr
Mole fraction $D_2 + O_2$ in gas	= 0.0075
Temperature of catalyst bed	= 145°C
Pressure drop across catalyst	= 9.6 in D_2O
Operating pressure	= 14 psia

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1.0 INTRODUCTION

It is planned to operate the HRT chemical processing plant two days each week and to store the concentrated fuel solution-solids mixture for decay of radioactivity. Being directly from the reactor the fuel solution has considerable beta and gamma emission some of which causes decomposition of heavy water into deuterium and oxygen. For uranium fuel concentrations of the order of 10 gm U/kg D_2O , the G value for the reaction has been established as 0.4 molecules D_2O decomposed per 100 ev of absorbed energy. As batches of radioactive solution are added weekly to the storage vessel, its activity will gradually rise, and consequently the rate of water decomposition will increase. Over a period of one year, which was considered to be the longest period that any one tank would be used, the rate of decomposition rises to 0.0144 gm moles D_2O per minute.

It is important that the heavy water be recovered for its monetary value and to preclude build up of explosive concentrations of D_2 and O_2 in the storage vessel and its associated equipment. Flame recombination and catalytic recombination have been demonstrated as satisfactory methods of D_2O recovery, and, of these, catalytic recombination offers certain advantages in this system. A chamber of platinized alumina is installed convenient to the vapor phase of the storage vessel, and a continuous flow of vapor is maintained through the bed. The catalyst will activate the reactants even at room temperature and initiate the reaction; however, it is desirable to feed heated gas to the bed to avoid poisoning the catalyst by condensed moisture.

As a precaution against obtaining explosive concentrations in the vapor space, a constant boil-up of D_2O is maintained to dilute the stream to a low percentage of D_2 . A certain amount of D_2O will be vaporized by attenuation of radiation, but, in addition to this, the storage vessels can be heated externally to furnish more diluent. It has been established experimentally that concentrations of hydrogen and oxygen in stoichiometric mixtures in H_2O vapor less than 17 mole percent are non-explosive. Recombined D_2O and diluent D_2O are condensed and refluxed to the decay-storage vessel. This cycle is continued until it is desired to concentrate the storage vessel contents and recover heavy water. At this time refluxing is discontinued and D_2O is routed to a D_2O receiver for return to the reactor.

2.0 RECOMBINER DESIGN

2.1 Design Basis

The following data were available for use in calculating the sizes of the recombiner and condenser for the low pressure system of the HRT chemical processing plant. The expected composition and properties of the gas stream passing through the recombiner are:

Composition	Rate	gm moles/min	lbs/hr
D ₂ O	2.9-17.85 ft ³ /min at 210°F and 14 psia	2.455 - 15.12	6.5 - 40
D ₂ *		0.0042 - 0.0144	0.00224 - 0.00764
O ₂ *		0.0021 - 0.0072	0.00896 - 0.0306

ΔH_R of D ₂ + O ₂ at 18°C	= 57,781 cal/gm mole D ₂ O
Viscosity of gas stream	= 0.0127 cp
Thermal conductivity of gas stream	= 0.0137 BTU-ft/hr-ft ² °F
Molar heat capacity D ₂ (assumed same as H ₂)	= 6.62 + 0.00081 T
Molar heat capacity O ₂	= 8.27 + 0.000258 T
Molar heat capacity of D ₂ O vapor (assumed same as H ₂ O)	= 8.22 + 0.00015 T
Catalyst	= Platinized alumina
Max % D ₂ in vapor	= $\frac{(0.0144)(100)}{2.455 + 0.0144}$
	= 0.584 % mole
Gas temperature entering recombiner	= 293°F (145°C)

2.2 Catalyst Volume for D₂ + O₂ Recombination

Beta and gamma radiation from the decay of fission products in the storage tanks C-1 and C-2 causes decomposition of heavy water into its constituents D₂ and O₂. The recovery of the decomposition products is important not only from an economic consideration but also from the viewpoint of preventing an explosive atmosphere being formed in vapor spaces of the low pressure system. An experimental program ** of the MIT Practice School involving the use of platinized alumina catalyst for the recombination of D₂ and O₂ developed the data that essentially 100% conversion could be attained at

$$\text{space velocity} = 420,000 \text{ ft}^3 \text{ reactants (max) / ft}^3 \text{ catalyst-hr.}$$

* The decomposition rate of heavy water due to beta and gamma radiation has been calculated by W. L. Carter and published in ORNL-CF-55-2-119, HRP-CP: D₂O Decomposition in Chemical Plant from Beta and Gamma Radiation.

** Gaven, J. V., Bacastow, R. B. and Herrington, A. C., Catalytic Recombination of Hydrogen and Oxygen---II, KT-134, Aug. 29, 1952.

Designing for maximum conditions where

$$\text{gas rate} = 17.85 \text{ ft}^3/\text{min} (\text{D}_2, \text{O}_2 \text{ and } \text{D}_2\text{O}),$$

the calculated volume of catalyst

$$\begin{aligned} &= (17.85)(60) \text{ ft}^3/\text{hr} \times \text{ft}^3 \text{ catalyst-hr}/4.2 \times 10^5 \text{ ft}^3 \\ &= 2.55 \times 10^{-3} \text{ ft}^3 \\ &= 4.4 \text{ in.}^3 \end{aligned}$$

This is such a small unit that the actual size should be much larger to facilitate construction and produce a unit with sufficient cross section to minimize pressure drop. The fabricated unit has

$$V_{\text{actual}} = 41.6 \text{ in.}^3.$$

Since the volume of catalyst was calculated on the basis of maximum flow which the operator may find is never necessary to employ, this quantity of catalyst should suffice for all normal conditions. (Note: since this design was completed and the recombiner fabricated, a flowsheet change has reduced the normal operating rate to 10.3 lbs/hr which is 25.7% of the design figure.)

2.3 Recombiner Geometry

The volume of catalyst needed to recombine chemical plant gases is quite small compared to that required in the HRT low pressure system; therefore, it is not necessary to use an elaborate design as that shown for the HRT recombiner on drawing No. E-18228 and described in ORNL-CF-Memo 54-11-1*. Instead this design will employ a short section of standard weight stainless steel pipe packed with platinized alumina catalyst. Pressure drop considerations govern the cross section of the unit, as will be shown in Paragraph 5.1, instead of process reaction requirements. The completed design is shown in Figure 1. No provisions are being made for changing catalyst inside the unit; instead, if this is required, the entire unit will be removed at the flanged joints and replaced with a completely new assembly. Recombined gases discharge directly into the condenser.

Construction details of this unit can be found on drawing No. D-20253, Revision 4, HRT Core Processing Recombiner-Condenser H-1 and H-2.

3.0 RECOMBINER THERMODYNAMICS

3.1 Heat of Reaction

The recombination of D_2 and O_2 takes place at 293°F , which is the temperature to which the reactants are heated by the superheater. If it is assumed that the heat of formation of D_2O is the same as that of H_2O , at 18°C

$$\Delta H_{18} = 57,781 \text{ cal/gm mole}$$

* Aven, R. E., HRT Recombiner Condenser Design, ORNL CF 54-11-1, Nov. 1, 1954.

FIGURE 1

RECOMBINER UNIT FOR HRT
CHEMICAL PROCESSING PLANT

At 293°F (145°C) the heat of reaction can be represented as

$$\Delta H_{145} = \Delta H_{18} + \int_{291}^{418} \Delta C_p dT.$$

$$\Delta C_p = C_p \text{ of } D_2O - 1/2 C_p \text{ of } O_2 - C_p \text{ of } D_2$$

If it is assumed that the heat capacities of D_2O and D_2 are the same as those of H_2O and H_2 ,

$$\begin{aligned} \Delta C_p &= (8.22 + 0.00015 T) - (4.13 + 0.000129 T) - (6.62 + 0.00081 T) \\ &= -2.53 - 0.000789 T \end{aligned}$$

Therefore

$$\begin{aligned} \Delta H_{145} &= 57,781 - 2.53 \int_{291}^{418} dT - 0.000789 \int_{291}^{418} T dT \\ &= 57,781 - 2.53(418-291) - 0.000395 (418^2 - 291^2) \\ &= 57,781 - 321 - 35 \\ &= 57,425 \text{ cal/gm mole} \end{aligned}$$

For these conditions where there are 0.0144 gm moles D_2O /min formed

$$\begin{aligned} \text{heat liberated} &= 57,425 \times 0.0144 \times 60 \\ &= 49,600 \text{ cal/hr} \\ &= 197 \text{ BTU/hr} \end{aligned}$$

3.2 Temperature of Gas Leaving Recombiner

Assuming that the recombination is adiabatic, all of the heat of reaction goes to heat the gases which include the products of combustion, the inerts and the reactants. A heat balance for the reaction shows

$$N_{D_2O} \int_{T_0}^T C_p dT - N_{D_2} \int_{T_0}^T C_p dT - N_{O_2} \int_{T_0}^T C_p dT = \Delta H_{T_0}$$

$$\begin{aligned} N_{D_2O} &= \text{moles diluent } D_2O + \text{moles } D_2O \text{ reaction product} \\ &= 2.455 + 0.0144 \\ &= 2.4694 \text{ gm moles/min} \\ N_{D_2} &= 0.0144 \text{ gm moles/min} \\ N_{O_2} &= 0.0072 \text{ gm moles/min} \\ T_0 &= 145^\circ\text{C} (418^\circ\text{K}) \\ \Delta H_{T_0} &= 49,600 \text{ cal/hr} \\ &= 826.6 \text{ cal/min} \end{aligned}$$

Substituting these values in the equation,

$$\begin{aligned}
 & 2.4694 \int_{418}^T (8.22 + 0.00015 T) dT - 0.0144 \int_{418}^T (6.62 + 0.00081 T) dT \\
 & - 0.0072 \int_{418}^T (8.27 + 0.000258 T) dT = 826.6 \\
 & 2.4694 \left[8.22(T-418) + 0.000075(T^2 - 418^2) \right] - 0.0144 \left[6.62(T-418) \right. \\
 & \quad \left. + 0.000405(T^2 - 418^2) \right] - 0.0072 \left[8.27(T-418) + 0.000129(T^2 - 418^2) \right] = 826.6
 \end{aligned}$$

After performing the indicated operations and collecting terms,

$$T^2 + 160,157.7 T - 67,111,575.8 = 0.$$

From this it is found that the gases leaving the recombiner are at

$$\begin{aligned}
 T &= 418.04^\circ\text{K} \\
 &= 145.04^\circ\text{C}
 \end{aligned}$$

The temperature was calculated for the condition where the quantity of diluent D_2O is the least to give the greatest temperature rise. Obviously the heat of reaction is so small that its effect on the temperature of the gas stream can be neglected.

3.3 Heat Content of Gases Entering Recombiner

The heat content of the gases is calculated for the maximum flowrate using 32°F (0°C) as the datum plane.

$$\begin{aligned}
 \Delta H_v \text{ of } \text{D}_2\text{O} \text{ at } 14 \text{ psia} &= 971.9 \text{ BTU/lb} \times 0.919 \\
 &= 893.2 \text{ BTU/lb } \text{D}_2\text{O}
 \end{aligned}$$

In this case the ratio of the latent heats of vaporization of D_2O and H_2O is 0.919.

$$\begin{aligned}
 \Delta H_f \text{ of saturated liquid } \text{D}_2\text{O} &= 177.6 \text{ BTU/lb (assumed same as } \text{H}_2\text{O)} \\
 \text{Heat content of } \text{D}_2\text{O} \text{ vapor} &= (893.2 + 177.6) \text{ BTU/lb} \times 40 \text{ lb/hr} \\
 &= 42,850 \text{ BTU/hr}
 \end{aligned}$$

$$\begin{aligned}
 \text{Heat content of } \text{D}_2 &= N_{\text{D}_2} \int_{273}^{418} C_{p,d} dT \\
 &= 0.0144 \int_{273}^{418} (6.62 + 0.00081 T) dT \\
 &= 0.0144 \left[6.62(418-273) + 0.00081(418^2 - 273^2) \right] \\
 &= 13.8 \text{ cal/min} \\
 &= 3.6 \text{ BTU/hr}
 \end{aligned}$$

$$\begin{aligned}
 \text{Heat content of } O_2 &= N_{O_2} \int_{273}^{418} C_p dT \\
 &= 0.0072 \int_{273}^{418} (8.27 + 0.000258 T) dT \\
 &= 0.0072 [8.27(418-273) + 0.000258(418^2 - 273^2)] \\
 &= 0.186 \text{ cal/min} \\
 &= 0.048 \text{ BTU/hr}
 \end{aligned}$$

It is readily apparent that the heat content of D_2 and O_2 is negligible compared to that of D_2O .

3.4 Heat Content of Gas Leaving Recombiner

The heat content of the gas leaving the unit is the sum of the heat entering and the heat of reaction. The exothermic reaction contributes little to the total.

$$\begin{aligned}
 \text{Heat content of leaving gas} &= 42,850 + 197 \\
 &= 43,047 \text{ BTU/hr}
 \end{aligned}$$

This heat balance omits the negligible amount of heat brought in as sensible heat in D_2 and O_2 . The datum is 32°F (0°C).

3.5 Temperature of Catalyst Pellets

The quantity of heat released by the recombination reaction is so small that a difference between the temperature of the gas stream and the temperature of the pellet will hardly be detectable. Therefore this calculation has not been made, and it will be assumed that the catalyst will remain at essentially the gas stream temperature. It will be noticed in the sketch of Figure 1 that a thermocouple will be placed in the center of the catalyst bed. Although it was originally intended that this temperature would be a reaction zone temperature, it is more appropriate to expect this point to read gas stream temperatures. The chances are that the recombination reaction will be completed within the first inch or less of bed length.

It might be pointed out that a calculation of pellet temperature for the HRT recombining has been made by R. E. Aven and is reported in Appendix IV of ORNL CF Memo 54-11-1. *

4.0 CATALYST SPECIFICATIONS

Catalyst material will be the same as that specified for the HRT recombining. This material is $1/8"$ x $1/8"$ right circular cylinders of alumina coated with 0.3% weight of pure platinum.

* Aven, R. E., HRT Recombiner Condenser Design, ORNL CF 54-11-1, Nov. 1, 1954.

5.0 PRESSURE DROP THROUGH CATALYST

The pressure drop through the catalyst bed will be calculated for normal operating conditions as well as for the maximum flowrates to be encountered. Pressure drop in the low pressure area vapor system will have to be maintained at less than 10 ft D₂O to prevent process liquid being forced up the reflux line into the D₂O condensate receiver. Two valves and an entrainment separator in the vapor line preceding the recombiner will contribute to the pressure drop of the system, but these will be small compared to the drop through the recombiner. Therefore, the cross sectional area of the catalyst bed will be determined from pressure drop considerations to maintain the system pressure drop within the 10 ft D₂O limit. Several exploratory calculations were made before it was established that a 3-inch, Sch. 40, pipe would be required to satisfy the pressure drop conditions.

The schematic diagram in Figure 2 shows the position of the recombiner with respect to other components and indicates the flow pattern in the chemical plant low pressure system.

5.1 Pressure Drop at Maximum Flow Rate

Conditions at the entrance and exit of the recombiner can be diagrammatically represented as follows:

Pressure drop for fluids flowing through porous media can be calculated by the equation *

$$\Delta P = \frac{Lv^2 f \rho}{2g_c D_p}$$

The variables have the following significance.

- ΔP = pressure drop through bed, lb/ft²
- ρ = density of gas, lb/ft³
- v = velocity of gas (superficial), ft/sec
- L = length of bed, ft
- f = friction factor, dimensionless
- g_c = gravity constant, ft/sec²
- D_p = particle diameter, ft

* Brown, G. G. et al., Unit Operations, page 213, Fig. 221, John Wiley and Sons, New York, 1950.

FIGURE 2

SCHEMATIC FLOW DIAGRAM OF
CHEMICAL PLANT LOW PRESSURE
SYSTEM SHOWING LOCATION OF
RECOMBINER.

The value of f has to be calculated from porosity and sphericity data; whereas, other terms of this equation are

$$\begin{aligned} v &= (40/3600) \text{ lb/sec} \times 31 \text{ ft}^3/\text{lb} \times 144/7.39 \text{ ft}^2 \\ &= 6.72 \text{ ft/sec} \\ L &= 5 \text{ in.} = 0.417 \text{ ft} \\ \rho &= 1 \text{ lb}/31 \text{ ft}^3 = 0.0322 \text{ lb/ft}^3 \\ g_c &= 32.2 \text{ ft/sec}^2 \\ D_p &= 0.125 \text{ in.} = 0.0104 \text{ ft} \end{aligned}$$

Sphericity is defined as the area of the sphere having the same volume as the particle divided by the area of the particle. The particle is a right circular cylinder having dimensions

$$\begin{aligned} \text{diameter} &= 1/8 \text{ inch} \\ \text{height} &= 1/8 \text{ inch} \end{aligned}$$

Defining

$$\begin{aligned} A_p &= \text{area of particle} \\ A_s &= \text{area of sphere} \\ V_p &= \text{volume of particle} \\ V_s &= \text{volume of sphere} \\ r_s &= \text{radius of sphere} \end{aligned}$$

the following equations can be derived.

$$\begin{aligned} A_p &= 2\pi r (r + h) \\ V_p &= \pi r^2 h \\ A_s &= 4\pi r_s^2 \end{aligned}$$

By definition the volume of the sphere must equal the volume of the particle.

$$V_s = V_p = \pi r^2 h$$

The volume of the sphere is also

$$V_s = 4/3 \pi r_s^3$$

Equating V_s and V_p , there is obtained

$$\begin{aligned} 4/3 \pi r_s^3 &= \pi r^2 h \\ r_s &= (3r^2 h/4)^{1/3} \end{aligned}$$

Substituting the value of r_s into the equation for A_s ,

$$A_s = 4\pi(3r^2h/4)^{2/3}$$

Representing the sphericity by ψ , there obtains

$$\psi = \frac{A_s}{A_p} = \frac{4\pi(3r^2h/4)^{2/3}}{2\pi r(r+h)} = \frac{2(3r^2h/4)^{2/3}}{r(r+h)}$$

Numerical values to be substituted into this equation are

$$r = 1/16 \text{ inch}$$

$$h = 1/8 \text{ inch}$$

$$\psi = \frac{2\left(\frac{3}{256 \times 8 \times 4}\right)^{2/3}}{1/16(1/16 + 1/8)} = 0.874$$

Porosity is defined as the volume of void space in the bed divided by the volume of the bed. Letting X equal the porosity, it has been found experimentally that

$$X = 0.245$$

Using these values for ψ and X and Fig. 220, page 212, of Unit Operations, the friction factor F_f can be evaluated.

$$F_f = 19,000$$

Using Fig. 219, page 211, of Unit Operations, it can be determined that

$$F_{Re} = 58.5$$

The data are now known for calculating the Reynolds No.

$$Re = \frac{D_p F_{Re} v \rho}{\mu}$$

$$\mu = \text{viscosity} = 0.0127 \text{ cp}$$

$$Re = \frac{(0.0104)(58.5)(6.72)(0.0322)}{(0.0127)(0.000672)} = 15,420$$

For this Reynolds No. it is found on Fig. 221, page 213, of Unit Operations that

$$\frac{f}{F_f} = \frac{2g_c D_p (-\Delta P)}{L v^2 \rho F_f} = 0.029$$

$$\text{Therefore } f = 0.029 F_f = (0.029)(19,000) = 551$$

Rearranging the above equation,

$$\begin{aligned} -\Delta P &= \frac{f L v^2 \rho}{2 g_c D_p} \\ &= \frac{(551)(0.417)(6.72^2)(0.0322)}{(2)(32.2)(0.0104)} \\ &= 498 \text{ lb/ft}^2 \\ &= 3.46 \text{ psi} \\ &= 8.5 \text{ ft D}_2\text{O} \end{aligned}$$

This is the pressure drop through the unit at the maximum anticipated flow rate of 40 lb D₂O/hour. The only other pressure drop in the low pressure system of any consequence is through the valves at carrier D-1; this drop however is not sufficient to cause the total to exceed 10 ft D₂O.

It has now been determined that the catalyst bed cross section should be equivalent to that of a 3-inch, Sch. 40, pipe to accomodate the maximum boil-up rate while maintaining an upper limit on the pressure drop. The volume of catalyst brought about by using a 3-inch pipe is much more than the theoretical requirement; however, the total quantity is still small and is not a factor in the design.

5.2 Pressure Drop at Normal Flow Rate

The normal vapor rate through the recombiner will be that caused by boil-up from decay of radioactivity plus an arbitrary quantity of steam to keep the gases below the explosive limit. 15 psig steam in the heating coil of the decay storage vessels should give a vapor rate of around 10.3 lb/hr. Diagrammatically the conditions can be represented as follows:

As in the case of maximum flow the pressure drop will be calculated by the equation

$$-\Delta P = \frac{L v^2 \rho}{2 g_c D_p}$$

The terms are the same as defined above in paragraph 5.1 and have the following values:

$$\begin{aligned} L &= 0.417 \text{ ft} \\ P &= 0.0322 \text{ lb/ft}^3 \\ g_c &= 32.2 \text{ ft/sec}^2 \\ D_p &= 0.0104 \text{ ft} \\ V &= (10.3/3600) \text{ lb/sec} \times 31 \text{ ft}^3/\text{lb} \times (144/7.39) \text{ ft}^2 \\ &= 1.73 \text{ ft/sec} \end{aligned}$$

The friction factor, f , has to be evaluated from sphericity and porosity data which has been determined above. F_f and F_{Re} are not functions of the flow rate so their values are the same as determined for maximum flow.

$$\begin{aligned} F_f &= 19,000 \\ F_{Re} &= 58.5 \end{aligned}$$

The Reynolds number for the conditions becomes

$$\begin{aligned} Re &= \frac{D_p F_{Re} v}{\mu} \\ &= \frac{(0.0104)(58.5)(1.73)(0.0322)}{(0.0127)(0.000672)} \\ &= 3970 \end{aligned}$$

At this Reynolds number it can be determined from Fig. 221, page 213, of Unit Operations that

$$\frac{f}{F_f} = \frac{2g_c D_p (-\Delta P)}{L v^2 F_f} = 0.041$$

Rearranging the above equation and solving for $(-\Delta P)$,

$$\begin{aligned} -\Delta P &= \frac{0.041 L v^2 F_f}{2g_c D_p} \\ &= \frac{(0.041)(0.417)(1.73^2)(0.0322)(19,000)}{(2)(32.2)(0.0104)} \\ &= 46.8 \text{ lb/ft}^3 \\ &= 0.325 \text{ psi} \\ &= 9.6 \text{ in. D}_{20} \end{aligned}$$

This figure is well within the available head, ca. 10 ft, and since the recombiner is the component causing the largest portion of the pressure drop, the normal boil-up rate can be easily maintained.

6.0 HEATING MANTLE (H-9 and H-10) DESIGN

A heating mantle is used on the recombiner to keep the catalyst bed hot and dry since moisture on the catalyst surface effectively poisons the bed. The actual heat dissipation of the mantle is quite small since it is not required to supply heat for any process operation. Heat given up by the mantle is composed of radiation and conduction to the recombiner and radiation and convection losses to the cell. Therefore it has not been necessary to design into the unit any definite capacity for heat transfer because it acts only as a blanket.

One requirement of the heating mantle is that it not interfere with removing and installing a recombiner unit. This requirement is most easily met by designing the unit in the form of a cradle for fitting around the lower half of the recombiner. With this in mind the unit was fabricated as shown in Figure 3 of a piece of sheet metal, curved to fit the contour of the recombiner, and stainless steel tubing which has been silver soldered to the sheet metal. The tubing carries high pressure steam (250 psig) for heating the recombiner.

6.1 Heat Loss To Surroundings

The heat loss from the mantle will be essentially due to radiation and convection to the surroundings and the following data are known for the system. It will be assumed that the entire surface of the heating mantle is at steam temperature and that the convection pattern is the same as if the unit were a horizontal pipe.

The equation given by McAdams * will be used to calculate the convection coefficient:

$$\frac{h_c D_o}{k_f} = 0.53 \left[\left(\frac{D_o^3 \rho_f^2 \beta g \Delta t}{\mu_f^2} \right) \left(\frac{c_p \mu_f}{k_f} \right) \right]^{0.25}$$

Surface temperature = t_s = steam temperature = 406°F

Cell temperature = t = 100°F

Cell pressure = 7.5 psia

Film temperature = $t_f = t_s - 0.5(t_s - t)_m$
 = 406 - 0.5 (406 - 100)
 = 253°F

Outside diameter of unit = D_o = 3.62 in.
 = 0.3 ft

Thermal conductivity at t_f = k_f = 0.02 BTU-ft/hr ft²°F

Density of air at T_f = $\rho_f = P/NRT_f$
 = 7.5/(1)(0.37)(460 + 253)
 = 0.0284 lb/ft³

Coefficient of thermal expansion = $\beta = 1/T = 1/(460 + 100)$
 = 1.785 x 10⁻³

Conversion factor = $g = 4.17 \times 10^8$ ft/sec²
 $\Delta t = 406 - 100 = 306^\circ\text{F}$

Viscosity at t_f = μ_f = 0.022 cp
 = 0.0532 lb/ft-hr

Heat capacity = c_p = 0.25 BTU/lb-°F

* McAdams, W. H., Heat Transmission, 2nd edition, page 244, Eqn. 23, McGraw-Hill Book Co., Inc., New York, 1942.

PLAN

ELEVATION

FIGURE 3
HEATING MANTLE FOR RECOMBINER

Substituting these values in the above equation,

$$\begin{aligned}\frac{0.3 h_c}{0.02} &= 0.53 \left[\frac{(0.3)^3 (0.0284)^2 (0.001785) (4.17 \times 10^8) (306)}{(0.0532)^2} \right] \frac{(0.25)(0.0532)}{0.02} \Big]^{0.25} \\ &= 0.53(109 \times 10^4)^{0.25} \\ &= 17.1 \\ h_c &= (17.1)(0.02)/0.3 \\ &= 1.1 \text{ BTU/hr-ft}^2\text{-}^\circ\text{F}\end{aligned}$$

Additional heat loss is due to radiation from the surface of the stainless steel. For a surface temperature of 406°F and an ambient temperature of 100°F, a value of the radiation coefficient, h_r , can be read from Figure 27, page 63, of Heat Transmission.

$$h_r = 2.6 \text{ BTU/hr-ft}^2\text{-}^\circ\text{F}$$

However this value is for an emissivity of unity. Since stainless steel has an emissivity of 0.4, the true value is

$$\begin{aligned}h_r &= (0.4)(2.6) \\ &= 1.04 \text{ BTU/hr-ft}^2\text{-}^\circ\text{F}\end{aligned}$$

The equation giving the total heat loss from the system is

$$q = (h_c + h_r) A \Delta T_m.$$

Since the mantle covers only the lower half of the recombiner, the area is only one-half the outside surface area of an equivalent length of pipe.

$$\begin{aligned}A &= 0.19 \text{ ft}^2 \\ q &= (1.1 + 1.04)(0.19)(406-100) \\ &= 124 \text{ BTU/hr}\end{aligned}$$

6.2 Steam Consumption

The steam consumption can be estimated to be

$$\begin{aligned}\text{weight steam} &= 124 \text{ BTU/hr} \times \text{lb}/820.1 \text{ BTU} \\ &= 0.15 \text{ lb/hr}\end{aligned}$$

7.0 RECOMBINER OPERATION

The recombiner is operated at all times that there is fuel solution-solids concentrate stored in the chemical plant low pressure system so that there is continuous recombination of D_2 and O_2 . Since the unit is installed in series in the vapor line between the decay-storage vessel and condenser, all of the gases must pass through the catalyst. Because of the completeness and facility of the catalytic recombination of D_2 and O_2 no control over the flow of reactants is needed or required. It is probably appropriate to say that operating procedures are of a precautionary nature rather than a controlling nature. In this respect the following suggestions are made:

1. At least 8 hours before admitting radioactive fuel solution to the decay-storage vessel, turn 250 psig steam into the heating mantle to heat and dry the catalyst bed.
2. At the same time turn 250 psig steam into the superheater.
3. Shortly before dropping the contents of the underflow receiver, admit cooling water to the condenser (H-1 or H-2) and the cooling jackets (H-14 and H-16 or H-15 and H-17). Also admit 5-10 psig steam in the coil of the decay-storage vessel.
4. Be sure valve HCV T-3-2 or HCV T-4-2 is open. The choice of valve depends upon the decay-storage vessel that is being used.
5. Continue the use of steam and cooling water as long as any fuel solution is held up in the decay-storage vessel.
6. Pressure, temperature and flow measuring elements are installed in the gas stream at appropriate points. (See Dwg. E-24173). The operation of the system may be followed by these observations.

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