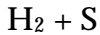


## ENERGY BALANCE

### Desulphurizer

Here Sulphur in Naphtha is made to react with Hydrogen in presence of catalysts to give Hydrogen Sulphide. This reaction takes place at a temperature of 623 K



Hydrogen and Naphtha are assumed to be stored at 303 K

Specific Heat capacity of naphtha is assumed to be 3.5235 kcal/kmol

Specific Heat capacity of Hydrogen = (6.62 + 0.00081T) kcal/kmol K

$$\Delta H = \sum_i n_i \int C_{pi} dT = 6.015915 \times 10^6 \text{ kcal/hr}$$

Hence heat required to raise their temperature from 303 K to 693 K

Assuming that the fuel used is Natural gas (calorific value = 39383.82 kJ/m<sup>3</sup>)

Thus amount of fuel needed to supply this quantity of heat = 152.75 m<sup>3</sup>/hr

Heat liberated within the reactor due to reaction = 1414.575 kcal/kmol

$$\Delta H_{sul} = \sum_i n_i \int_{623}^T C_{pi} dT = 15560.325$$

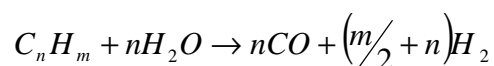
$\Delta H_r$  Hence total heat liberated = 11.1 x 1414.575 = 15560.325 kcal/hr

This heat released is utilized in increasing the temperature of the effluents

Solving the equation by a trial and error process we obtain the outlet temperature value as 623.8 K. Hence it is assumed that the outlet temperature is 623 K itself.

### Primary Reformer

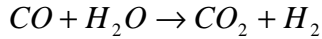
Within the reformer steam reacts with Naphtha and produces Carbon monoxide and Hydrogen. Side reactions also take place producing Carbon dioxide and



Methane. The exit gases from the reformer are at a temperature of 1093 K .

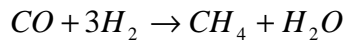
$$\Delta H_r \text{ is } 53.376 \times 10^3 \text{ kcal/kmol}$$

$$\Delta H_1 \text{ is } 10.29823 \times 10^7 \text{ kcal/hr}$$



$$\Delta H_r \text{ is } -7.834 \times 10^3 \text{ kcal/kmol}$$

$$\Delta H_2 \text{ is } -0.44489 \times 10^7 \text{ kcal/hr}$$



$$\Delta H_r \text{ is } -49.271 \times 10^3$$

$$\Delta H_3 \text{ is } -1.66672 \times 10^7 \text{ kcal/hr}$$

Heat required to raise the temperature of the reaction products from 673 K to 1093 K is

$$\Delta H_{ref} = \sum_i n_i \int_{623}^T n_i C_{pi} dT = 13.94433 \times 10^7 \text{ kcal/hr}$$

Thus total heat to be supplied is sum of all the heat requirements and the enthalpy of reactions =  $22.11061 \times 10^7$  kcal/hr

Fuel required for supplying this amount of heat is  $5614.13 \text{ m}^3/\text{hr}$  of natural gas.

### Secondary Reformer

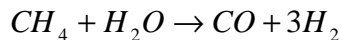
In the secondary reformer methane produced is converted into CO and H<sub>2</sub>.

Here air is mixed with the effluent stream in such a quantity that the exit stream contains 1:3 ratio of N<sub>2</sub>: H<sub>2</sub>. It is assumed that the inlet to the reformer enters at 1093 K and reaction takes place at this temperature.

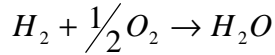
Heat required to raise the temperature of air from ambient conditions at 303 K to 1093 K is

$$\Delta H_{air} = \sum_i n_i \int_{623}^T n_i C_{pi} dT = 2.16325 \times 10^7 \text{ kcal/hr}$$

This heat is supplied by natural gas. Hence fuel required to provide this amount of heat is  $549.3 \text{ m}^3/\text{hr}$

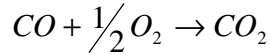


$$\Delta H \text{ is } 1.96936 \times 10^7 \text{ kcal/hr}$$



$\Delta H$  is  $-2.88989 \times 10^7$  kcal/hr

$\Delta H$  is  $-1.28508 \times 10^7$  kcal/hr



Total heat liberated during reaction is  $2.26464 \times 10^7$  kcal/hr. This heat is utilized in heating the effluents from 1093 K to the exit temperature.

$$\Delta H = \sum_i n_i \int_{1093}^T C_{pi} dT = 2.26464 \times 10^6 \text{ kcal/hr}$$

Thus by a trial and error process the temperature is found to be 1262 K.

### Heat Recovery between Secondary Reformer and HTSC

The stream entering the HTSC is at 693 K (value obtained from literature). Hence the stream is cooled from 1262 K to 693 K. The heat recovered from this stream is utilized in producing steam in quenchers.

Heat recovered is

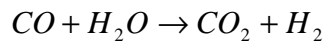
$$\Delta H = \sum_i n_i \int_{1093}^T C_{pi} dT = 2.26464 \times 10^7 \text{ kcal/hr}$$

Amount of steam at 373 K at 1 atm produced from this heat is

$$m_s = \frac{\Delta H}{C_{pw}\Delta T + \lambda_s} = 7401.87 \text{ kmol/hr}$$

### High Temperature Shift Converter

In the HTSC Carbon monoxide percentage is reduced to around 3% by volume of the exiting stream by converting it to Carbon dioxide by the water shift reaction.



$\Delta H_r$  is  $-7.963 \times 10^3$  kcal/kmol at 693 K

$\Delta H$  is  $-0.77180 \times 10^7$  kcal/hr .

This heat is utilized in increasing the heat content of the exit stream.

$$\Delta H = \sum_i n_i \int_{693}^T C_{pi} dT = 0.77180 \times 10^7 \text{ kcal/hr}$$

Solving for T by a trial and error process we get T = 700 K

### Heat recovery from HTSC to LTSC

The effluents from the HTSC are at a temperature of 700 K. The inlet stream into the LTSC is at a temperature of 523 K. The heat recovered from this stream is utilized in producing steam at 373 K and 1 atm in quencher.

Heat recovered

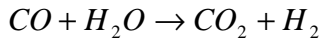
$$\Delta H = \sum_i n_i \int_{523}^{700} C_{pi} dT = 2.13555 \times 10^7 \text{ kcal/hr}$$

$$m_s = \frac{\Delta H}{C_{pw} \Delta T + \lambda_s} = 2181.36 \text{ kmol/hr}$$

Amount of steam produced will be

### Low Temperature Shift Converter

Here the Carbon monoxide content is reduced to 0.3% by volume of the total effluent gases. The reaction being exothermic heat is liberated which is used to increase the temperature of the outlet stream.



$\Delta H_r$  is  $-8.237 \times 10^3$  kcal/kmol at 523 K

$\Delta H$  is  $-0.27252 \times 10^7$  kcal/hr

The temperature of the outlet stream can be found out by making a heat balance

$$\Delta H = \sum_i n_i \int_{523}^T C_{pi} dT = 0.27252 \times 10^7 \text{ kcal/hr}$$

By a trial and error process the temperature is found to be 543 K.

### Condenser

The exit stream from the LTSC contains steam which would be an additional unwanted load on the absorber. Hence steam is removed by condensation and the dry gases leave the condenser at 493 K and are cooled to 373 K which is the temperature at which the gases enter the absorber. It is assumed that all steam produced is removed by condensation

Heat removed by condensation

$$Q_c = n_s \lambda_s = 4.16939 \times 10^7 \text{ kcal/hr}$$

### Heat Exchanger

The gases are cooled from 493 K to 373 K. The heat transferred during this process is

$$\Delta H = \sum_i n_i \int_{373}^{493} C_{pi} dT = 1.40919 \times 10^7 \text{ kcal/hr}$$

### Methanator

Here the trace quantities of CO and CO<sub>2</sub> are converted to Methane by reaction with H<sub>2</sub> in the presence of catalysts. The methanator operates at a temperature of 623K. The entering gases have to be heated from 373 K to 623 K.

Heat supplied

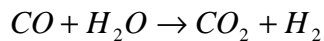
$$\Delta H = \sum_i n_i \int_{373}^{623} C_{pi} dT = 3.17869 \times 10^7 \text{ kcal/hr}$$

This heat may be supplied by condensing steam and utilizing the heat of condensation.

$$m_s = \frac{\Delta H}{\lambda_s} = 2970.6 \text{ kmol/hr}$$

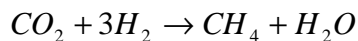
Amount of steam at 373 K and 1 atm required to heat the gases is

Reactions taking place in the methanator are



$\Delta H_r$  is  $-49.271 \times 10^3$  kcal/kmol at 623 K

$\Delta H$  is  $-1.36386 \times 10^6$  kcal/hr



$\Delta H_r$  is  $-39.433 \times 10^3$  kcal/kmol at 623 K

$\Delta H$  is  $-0.31561 \times 10^6$  kcal/hr

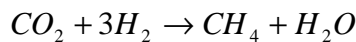
Total heat liberated within the methanator is used to increase the temperature of the exit gases.

$$\Delta H = \sum_i n_i \int_{623}^T C_{pi} dT = 1.67949 \times 10^6 \text{ kcal/hr}$$

The temperature of the exit stream found out by a trial and error process is 632 K. The gases from the Methanator are cooled, water removed by condensation and the dry gases cooled to 313 K and stored.

### Ammonia Synthesis Loop

The synthesis gas enters the converter at a temperature of 673 K. From the compressor section it is assumed that the gases come out at 313 K. The inlet stream is heated by the exit stream to the converter inlet temperature. The converter exit gas is then cooled to 313 K at which temperature at the high pressure of 50 MPa ammonia condenses.



$\Delta H_r$  is  $-7.301 \times 10^3$  kcal/kmol at 673 K

$\Delta H$  is  $-2.73698 \times 10^7$  kcal/hr

This heat produced is utilized in heating the exit gases to a final exit temperature, found by doing a heat balance.

$$\Delta H = \sum_i n_i \int_{673}^T C_{pi} dT = 2.73698 \times 10^7 \text{ kcal/hr}$$

The exit temperature from the condenser found out by a trial and error process is  $T = 778$  K

Ammonia condenses at a temperature of 313 K at the given pressure and mole fraction of 0.0952

Heat evolved by cooling gases from 778 K to 313 K

$$\Delta H = \sum_i n_i \int_{313}^{778} C_{pi} dT = 11.59567 \times 10^7 \text{ kcal/hr}$$

Heat needed to raise the temperature of synthesis gases from 313 K to 673 K is

$$\Delta H = \sum_i n_i \int_{313}^{673} C_{pi} dT = 9.45070 \times 10^7 \text{ kcal/hr}$$

Thus the excess heat, which is to be removed by the **Heat Exchanger**, is

$$\Delta H = 2.14497 \times 10^7 \text{ kcal/hr}$$

As the heat load is very high five HE in parallel are used. The inlet temperature of the hot gases found by a trial and error process is 405 K. The exit temperature is ammonia saturation temperature of 313 K.

Cooling water is assumed to enter at 293 K and exit at 308 K. The mass flow rate of cooling water is calculated and found to be 79.4 kg/s per HE.

**Condenser heat load may be calculated as**

$$Q_C = n_N \lambda_N = 2.02284 \times 10^7 \text{ kcal/hr}$$

Where  $\lambda_N$  is the latent heat of vaporization of ammonia at the given pressure  
= 4471.65 kcal/kmol

The condenser coolants used are refrigerants because water cannot be used to cool the contents.