HEAT BALANCE

Estimated heat capacities at different temperature:

In the reactor the inlet feed temp is 660^0C where the outlet temp is 600^0C. Hence for the energy balance about the reactor it is assumed that the physical properties are constant, ie values at average temp (600+660)/2 or 630^0C.

C_P of steam at 630^0C= 2.223 kJ/kg k

C_P of EB at 630^0C = 2.827 kJ/kg k

C_P of styrene at 630^0C = 2.610 kJ/kg k

Heat of formation (H_R)= $\mathbf{\Delta H_R|_{25} + \Sigma(\gamma_i * C_{Pi})|_{PORDUCT} - \Sigma(\gamma_i * C_{Pi})|_{REACTANT}}$

of styrene at 630^0 = 86.99 kJ/mol

- STEAM REQUIREMENT:

Let m is the mass of steam added to the feed per hour, in order to supply the heat of reaction in the reactor.

The steam and the reactant feed entered the reactor at a temperature 660^0C and leave at 600^0C temperature. Hence, it can be assumed that the temperature drop solely due to the reaction the approx. energy balance is made;

$$\mathbf{m * C_{Pw} * \Delta T + m_{ST} * C_{PST} * \Delta T + m_{EB} * C_{PEB} * \Delta T + Q_R = 0}$$

From the above balance the steam flow rate been calculated as;

m= 42930.83 kg/ hr

Henceforth the mol ratio of steam to hydrocarbon feed is approximately calculated = 11:1 (which is a close match to values obtained from literature[1])

- TEMPERATURE OF THE PREHEATED STREAM:

Let T be the temp of the preheated coming out of the preheater. Assume the mixing of the steam to hydrocarbon stream is adiabatic.

It is known that the temperature of the preheated steam is 800^0C and after mixing with the hydrocarbon stream, the temperature is dropped to 660^0C. Therefore, the heat balance can be given as:

42930.83*2.293*(800-660) =(5687.5*2.133+18187.5*2.267)(660-T)

i.e., T= 401.73^0 C

- TEMPERATURE DETERMINATION OF THE REACTION PRODUCT LEAVING THE PREHEATER:

Again, from mass balance about the reactor the wt fraction or composition of the stream leaving the reactor can be given as:

Steam= 47.64%

EB = 13.95%

Styrene=36.63%

Toluene= 0.67%

Benzene= 0.38%

Hydrogen=0.73%

Therefore the average C_p value been determined= 2.322 kJ/kg k

And the C_p value for the feed stream to the reactor= 2.235 kJ/kg k

The boiling point temp of the feed mixture at atmospheric pressure is calculated by using Roult's law is=138^0C

And the latent heat of vaporization of the mixture= 341.03 kJ/kg

The C_p value for the liquid mixture = 1.939 kJ/kg k

Let the outlet temp of the product stream from the preheater = t

Therefore, the approx energy balance can be given as:

66805.83*2.322*(630-t)= 23875*(1.939*(138-30)+341.03+2.235*(401-38))

i.e., t=454^{0}C

- <u>WASTE HEAT RECOVERY</u>

The excess of superheat from the stream leaving the reactor and through steam preheater, is been utilized in generating steam in boiler, which further fed back to the reactor system.

The temp of the product stream leaving the preheater=454^{0}C

& C_p value of the stream is =2.322 kJ/kg k

Let the boiler operated at a temp of 100^{0}C

The water feed to the condenser is at temp= 40^{0}C

The latent heat of vaporization of water at 100^{0}C is= 2260 kJ/kg

Therefore the amt. of steam generated is:

66805.83*2.322*(454-110) = M*(4.184*(100-40)+2260)

ie, M=21251 kg/hr

hence the amount of steam been generated= 21251 kg/hr

- <u>ENERGY REQUIRED FOR SUPER HEATING OF STEAM:</u>

The steam is been heated to a temperature of 800^{0}C in the super heater.

The specific heat of steam at 400^{0}C = 2.065 kJ/kg k

Therefore the heat requirement in the super heater=

Q= (42930.83/3600)*2.065*(100-60)

= 17,238 KW

- <u>ENERGY BALANCE FOR BENZENE-TOLUENE TOWER:</u>

The top condenser operator at the saturator temperature corresponding to 160 mm Hg pressure

The condenser is operated at 58^0C.

% COMPOSITION

WT. BASIS

	BENZENE	TOLUENE	ETHYL BENZENE	STYRENE	TEMP.(^{O}C) (APPROX)	MASS FLOW RATE KG/HR)
FEED	0.75	1.30	26.94	71.01	90	26545
DISTILLATE	50.0	40.0	10.0	0	58	313.17
RESIDUE	0.2	0.8	27.1	71.9	96	26231.5

And the average latent heat of vaporization = 401 kJ/kg

The distillate from the top of the tower =313.17 kg

Let, assume the reflux ratio to the tower maintained at=2.7(R)

Therefore the actual mass flow into the condenser= 313.17(1+2.7) kg = 1158.73 kg/hr

Let the inlet and outlet temperatures of water are 25^0C and 35^0C respectively.

Therefore, the water flow into the reactor can be calculated as:

M_w = 1158.73*401/ (3600*4.184*(35-25))

= 3.085 kg/s

$\therefore$,Q_c= 129 kW

let, T_R (reference temperature)= 273 k

avg. Cp measured at 40^oC:

$Cp,_{benzene}$= 1790 J/kg k

$Cp_{, toluene}$= 1742 J/kg k

$Cp,_{EB}$ = 1751 J/kg k

$Cp,_{styrene}$= 1790 J/kg k

$\therefore$, $Cp_{avg,feed} = \sum(x_i\ Cp_i)$ = 1779 J/kg k

$Cp_{avg,distillate}$ = 1767 J/kg k

$Cp_{avg,residue}$ = 1779 J/kg k

$Fh_f = F*(Cp_{avg,feed} *(T_f - T_R))$ = 1180.5 kW

$Wh_w = W*(Cp_{avg,residue}*(T_w - T_R))$= 1244.423 kW

$Dh_D = D*(Cp_{avg, distillate}*(T_D - T_R))$= 8.915 kW

Again assuming no heat loss in the distillation column:

$Q_B + Fh_f = Q_c + Wh_w + Dh_D$

$\therefore$, Q_B = 201.12 kW

therefore load on the toluene benzene tower reboiler= 201.12 kW

- <u>ETHYL BENZENE RECYCLE COLUMN:</u>

Assumption made: the amount of toluene benzene & heavy ends are negligible compare to ethyl benzene and styrene.

% COMPOSITION

WT. BASIS

	Ethyl benzene	styrene	Temperature (^oC)	Mass Flow

				rate (kg/hr)
Distillate	50	50	61.3	12880
Feed	27.1	72.9	96	26231.59
Residue	5.0	95	65	13348.98

$\therefore$, $Cp_{avg,feed} = \sum(x_i\ Cp_i) = 1779$ J/kg k

$Cp_{avg,distillate} = 1770$ J/kg k

$Cp_{avg,residue} = 1788$ J/kg k

Where, R= 3.99

$\lambda_{avg,top} = 392.5$ kJ/kg

$\therefore$, $Q_c = 6923$ kW

Again,

$Fh_f = F*(Cp_{avg,feed} *(T_f - T_R)) = 1244$ kW

$Wh_w = W*(Cp_{avg,residue}*(T_w - T_R))= 430$ kW

$Dh_D = D*(Cp_{avg,\ distillate}*(T_D - T_R))= 388$ kW

$\therefore$, $Q_B = Q_c + Wh_w + Dh_D - Fh_f = 6497$ kW

- FINAL PURIFIER TOWER:

% COMPOSITION

WT. BASIS

	STYRENE	HEAVY ENDS	TEMPERATURE (^{O}C)	MASS FLOW RATE (kg/HR)
FEED	95	5.0	65	13348.93
DISTILLATE	99.0	1.0	61	12500
RESIDUE	2.3	97.0	80	848.93

λ_{AVG} = 406 kJ/kg

Let, R= 1.2

Q_c = 2678 kW

$\therefore$, $Q_B = Q_c + Wh_w + Dh_D - Fh_f$ = 2676.5 kW